EPA/452/B-02-001

EPA AIR POLLUTION CONTROL COST MANUAL

Sixth Edition

EPA/452/B-02-001

January 2002

United States Environmental Protection Agency
Office of Air Quality Planning and Standards
Research Triangle Park, North Carolina 27711

This sixth edition of the EPA Air Pollution Control Cost Manual was prepared by the Air Quality Strategies and Standards Division of the Office of Air Quality Planning and Standards, U.S. Environmental Protection Agency, Research Triangle Park, NC 27711. Mention of trade names or commercial products is not intended to constitute endorsement or recommendation for use. Copies of this report are available through the OAQPS Clean Air Technology Center (MD-15), U.S. Environmental Protection Agency, Research Triangle Park NC 27711, or from the National Technical Information Service, 5285 Port Royal Road, Springfield VA 22161, (phone: 1-800-553-6847.)

Questions and comments should be addressed to the principal editor, Daniel C. Mussatti, OAQPS, phone 919-541-0032.

TABLE OF CONTENTS

EPA Air Pollution Control Cost Manual (Sixth Edition)

Section/Chapter	Title
Section 1	**Introduction**
Chapter 1	Background
Chapter 2	Cost Estimation: Concepts and Methodology
Section 2	**Generic Equipment and Devices**
Chapter 1	Hoods, Ductwork, and Stacks
Chapter 2	Fans, Pumps, and Motors [2]
Chapter 3	Permanent Total Enclosures [2]
Chapter 4	Monitors[1]
Section 3	**VOC Controls**
Section 3.1	**VOC Recapture Controls**
Chapter 1	Carbon Adsorbers
Chapter 2	Refrigerated Condensers
Section 3.2	**VOC Destruction Controls**
Chapter 1	Flares
Chapter 2	Incinerators
Section 4	**NO_x Controls**
Section 4.1	**NO_x Combustion Controls**
Chapter 1	Fuel Switching [2]
Chapter 2	Low NOx Burners and other Combustion Controls [2]
Chapter 3	Steam/Water Injection [2]
Chapter 4	Natural Gas Reburn [2]
Section 4.2	**NO_x Post-Combustion**
Chapter 1	Selective Non-Catalytic Reduction [1]
Chapter 2	Selective Catalytic Reduction [1]
Chapter 3	Non-Selective Catalytic Reduction [2]
Section 5	**SO_2 and Acid Gas Controls**
Section 5.1	**Pre-Combustion Controls**
Chapter 1	Fuel Substitution [2]
Section 5.2	**Post-Combustion Controls**
Chapter 1	Wet Scrubbers for Acid Gas
Chapter 2	Wet and Dry Scrubbers for SO_2 [2]
Section 6	**Particulate Matter Controls**
Chapter 1	Baghouses and Filters
Chapter 2	Wet Scrubbers for PM [2]
Chapter 3	Electrostatic Precipitators
Chapter 4	Inertial Impactors [2]
Section 7	**Mercury Controls** [1]

[1] New Chapter
[2] Planned Chapter

EPA/452/B-02-001

Section 1

Introduction

EPA/452/B-02-001

Chapter 1

Background

Daniel C. Mussatti
Innovative Strategies and Economics Group
Air Quality Strategies and Standards Division
Office of Air Quality Planning and Standards
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

January 2002

Contents

1.1 Role of Cost in Setting of Regulations 1-3

1.2 Purpose of the Manual 1-4

1.3 Organization of the Manual 1-5

1.4 Intended Users of the Manual 1-7

1.5 "Uniqueness" of the Manual 1-8

1.1 Role of Cost in Setting of Regulations

Cost has an important role in setting many state and federal air pollution control regulations. The extent of this role varies with the type of regulation. Some types of regulations, such as Maximum Achievable Control Technology (MACT) standards, explicitly use costs in determining their stringency. This use may involve a balancing of costs and environmental impacts, costs and dollar valuation of benefits, or environmental impacts and economic consequences of control costs. Other types of regulations (e.g., National Ambient Air Quality Standards), use cost analysis to choose among alternative regulations with the same level of stringency. For these regulations, the environmental goal is determined by some set of criteria which do not include costs. However, regulators use cost-effectiveness analysis to determine the minimum cost way of achieving the goal.

For some regulations, cost influences enforcement procedures or requirements for demonstration of progress towards compliance with an air quality standard. For example, the size of any monetary penalty assessed for noncompliance as part of an enforcement action must include the cost of the controls that were not installed by the noncompliant facility. For regulations without a fixed compliance schedule, demonstration of reasonable progress towards the goal is sometimes tied to the cost of attaining the goal on different schedules.

Cost is also a vital input to the EPA's standard setting and regulatory processes. Through various Executive Orders and acts, EPA has been charged with performing a number of detailed economic and benefit-cost analyses on each proposed rulemaking to assess their economic efficiency and assure the public the best possible regulation has been chosen from among alternative regulations. Cost also plays an input role in determining the economic impact of each regulatory alternative on sensitive populations, small businesses, employment, prices, and market and industry structure.

This Manual provides up-to-date information on point source and stationary area source air pollution controls for volatile organic compounds (VOCs), particulate matter (PM), oxides of nitrogen (NO_X), and some acid gasses (primarily SO_2 and HCl). It is not a source of information for non-stationary area (e.g. emissions from fugitive dust sources, agricultural sources) and mobile sources. Furthermore, this Manual does not directly address the controls needed to control air pollution at electrical generating units (EGUs) because of the differences in accounting for utility sources. Electrical utilities generally employ the EPRI Technical Assistance Guidance (TAG) as the basis for their cost estimation processes.[1] Finally, new and emerging technologies are not generally within the scope of this Manual. The control devices included in this Manual are generally well established devices with a long track record of performance.

[1]This does not mean that this Manual is an inappropriate resource for utilities. In fact, many power plant permit applications use the Manual to develop their costs. However, comparisons between utilities and across the industry generally employ a process called "levelized costing" that is different from the methodology used here.

1.2 Purpose of the Manual

The objectives of this Manual are two-fold: (1) to provide guidance to industry and regulatory authorities for the development of accurate and consistent costs (capital costs, operating and maintenance expenses, and other costs) for air pollution control devices, and (2) to establish a standardized and peer reviewed costing methodology by which all air pollution control costing analyses can be performed. To perform these objectives, this Manual, for the last twenty-five years, has compiled up-to-date information for "add-on" (downstream of an air pollution source) air pollution control systems and provided a comprehensive, concise, consistent, and easy-to-use procedure for estimating and (where appropriate) escalating these costs. Over time, the accessability of this Manual and its ease of use has significantly increased. Its early editions were only available in hard copy by request, mailed from the EPA's Office of Air Quality Planning and Standards in Research Triangle Park, North Carolina. Later editions became available electronically; first through the EPA's Technology Transfer network (TTN) bulletin board in the early nineties, later as a fully accessible series of documents on the Internet through the Agency's Clean Air Technology Center. The Manual is a living document, evolving continuously to meet the changing needs of its customers, and now, with supporting programs for the personal computer such as the CO$T-AIR spreadsheets and the Air Compliance Advisor that streamline and simplify the input of site-specific information, the Manual is even more accessible and important.

As always, to achieve its objectives, the Manual provides detailed engineering information that reflects the latest innovations in the industry and costing information that is up-to-date and relevant. The accuracy of the information in the Manual works at two distinct levels. From a regulatory standpoint, the Manual estimating procedure rests on the notion of the "study" (or rough order of magnitude - ROM) estimate, nominally accurate to within ± 30%. This type of estimate is well suited to estimating control system costs intended for use in regulatory development because they do not require detailed site-specific information necessary for industry level analyses. While more detailed data are available to the regulator, those data are generally proprietary in nature (which limits their ability to be published), costly to gather, and too time consuming to quantify. Therefore, for regulatory analysis purposes, study estimates offer sufficient detail for an assessment while minimizing its costs. The Manual and its supporting programs are also well suited to customization by industrial sources to provide more accurate assessments of control cost sizing and cost that can be used for scoping level decision making and planning purposes. While such customized analyses are by definition of greater accuracy than the generic study level analysis of the regulator, the Agency does not make any claim for a greater accuracy than the study level's nominal ±30 percent.

The Manual offers an additional, benefit to its users. When industry uses the Manual and its support programs to determine its control costs for permitting purposes, and the regulator uses the Manual (and its support programs) to validate industry's permit, the approval process can be faster and less expensive. With a common peer reviewed costing methodology used by

all parties, regulators and permitting authorities can minimize the time it takes to perform a permit review because the honest application of the methodology set out in this Manual by both industry and the regulator should provide results that are roughly similar. Differences in conclusions are reduced to a comparison of input parameters, rather than a protracted debate on the veracity of alternative models. This internal consistency allows industry and the regulatory community to work in partnership to bring industrial growth on line faster and produce needed pollution abatement sooner.

1.3 Organization of the Manual

This Manual is a revision of the fifth edition of the OAQPS Control Cost Manual, [1] which, in turn, was a revision of the edition completed in 1990. This sixth edition of the Manual includes sizing and costing procedures and data for the same eight types of add-on control devices and three kinds of auxiliary equipment available in the fifth edition, but beyond the necessary revisions, updates, and expansions of each of these chapters, the Manual has made a number of revolutionary changes.

As with earlier editions, this edition has been issued in self-contained chapters. Each chapter addresses a logically separate topic, which can be either of a general nature (e.g., this introduction) or of a more specific, equipment-oriented nature (e.g., fabric filters). To fully assess the sizing and cost of a particular air pollution control device, you only have to access one chapter - with one exception. For auxiliary equipment common to many different pollution control devices, that information has not been repeated in each chapter. Instead, auxiliary equipment is handled as a separate set of chapters in the section immediately following this introduction. The chapters which comprise this portion of the Manual are listed in Table 1.1, alongside the portions of the 1990 Manual they replace.

Each of these stand-alone chapters contains a:

- Process description, where the types, uses, and operating modes of the equipment item and (if applicable) its auxiliaries are discussed;
- Sizing (design) procedure, which enables one to use the parameters of the pollution source (*e.g.*, gas volumetric flow rate) to size the equipment item(s) in question;
- Capital and annual costing procedure and data for the equipment and suggested factors to use in estimating these costs from equipment design and operational (*e.g.*, operating hours) parameters. These costs are presented in both graphical and equation forms wherever possible.
- Example problems to illustrate the sizing and costing procedures presented in the chapter.

Table 1.1: Comparison of the the OAQPS Control Cost Manual (5th Edition) and the EPA Air Pollution Control Cost Manual (6th Edition)

Sixth Edition Section/Chapter	Title	Fifth Edition Chapter
Section 1	**Introduction**	
Chapter 1	Background	1
Chapter 2	Cost Estimation: Concepts and Methodology	2
Section 2	**Generic Equipment and Stacks**	
Chapter 1	Hoods, Ductwork, and Stacks	10
Chapter 2	Fans, Pumps, and Motors [2]	
Chapter 3	Permanent Total Enclosures [2]	
Chapter 4	Monitors [1]	
Section 3	**VOC Controls**	
Section 3.1	**VOC Recapture Controls**	
Chapter 1	Carbon Adsorbers	4
Chapter 2	Refrigerated Condensers	8
Section 3.2	**VOC Destruction Controls**	
Chapter 1	Flares	7
Chapter 2	Incinerators	3
Section 4	**NO_x Controls**	
Section 4.1	**NO_x Combustion Controls**	
Chapter 1	Fuel Switching [2]	
Chapter 2	Low NOx Burners and other Combustion Controls [2]	
Chapter 3	Steam/Water Injection [2]	
Chapter 4	Natural Gas Reburn [2]	
Section 4.2	**NO_x Post-Combustion**	
Chapter 1	Selective Non-Catalytic Reduction [1]	
Chapter 2	Selective Catalytic Reduction [1]	
Chapter 3	Non-Selective Catalytic Reduction [2]	
Section 5	**SO_2 and Acid Gas Controls**	
Section 5.1	**Pre-Combustion Controls**	
Chapter 1	Fuel Substitution [2]	
Section 5.2	**Post-Combustion Controls**	
Chapter 1	Wet Scrubbers for Acid Gas	9
Chapter 2	Wet and Dry Scrubbers for SO_2 [2]	
Section 6	**Particulate Matter Controls**	
Chapter 1	Baghouses and Filters	5
Chapter 2	Wet Scrubbers for PM [2]	
Chapter 3	Electrostatic Precipitators	6
Chapter 4	Inertial Impactors [2]	
Section 7	**Mercury Controls [2]**	

[1]New Chapter

[2]Planned Chapter

Finally, each chapter is supported by supplemental programs that streamline and expand upon the information in the Manual:

- for control cost analyses, we developed a series of Lotus spreadsheets (filename: COST.AIR.ZIP), that have been installed on the CATC website, allow the user to size and cost any of the control devices covered in the Manual, plus several others (e.g., venturi scrubbers) that are planned for future inclusion in the Manual; and

- further automating the costing procedure and offering important engineering information, we have also developed a highly automated program, the Air Compliance Advisor, which, following the input of detailed site-specific information on the facility, offers sizing and costing information on all potential controls.

The scope of the Manual has also changed over time. Initially, the Manual provided information on a small number of volatile organic compound (VOC) pollutants that were of primary interest to the abatement of ozone and hazardous air pollutants (HAPs). Over time, the Manual grew to include particulate matter (PM) and some acid gasses. As new pollutants and new devices were added, they were simply added on to the end of the chapter list, which resulted in an unsystematic approach to the discussion of pollution control. Now, in this latest edition of the Manual, we have branched out even farther with new chapters on NO_X post-combustion control devices, planned chapters on NO_X combustion controls, expansion of the acid gas scrubber chapter to include SO_2 more explicitly, and a new chapter for fine particulate controls. To accommodate these new chapters, we have also changed the look of the Manual, rearranging chapters into logical pollutant groupings and restructuring the numbering system to allow for new chapters under each of these pollutant headings. Further discussion of these changes can be found in the next section, below.

1.4 Intended Users of the Manual

The Manual provides comprehensive procedures and data for sizing and costing control equipment. Some of these procedures are based on rigorous engineering principles, such as the material and engineering balances presented for Thermal Incinerators (Section 3, Chapter 2). To fully appreciate and correctly apply these procedures the user must be able to understand them. Moreover, the user has to be able to exercise "engineering judgement" on those occasions when the procedures may need to be modified or disregarded. Typically, engineers and others with strong technical backgrounds possess this kind of knowledge. Hence, this Manual is oriented toward the technical not the non-technical user.

1.5 "Uniqueness" of the Manual

The Manual presents a unique approach to estimating air pollution control system sizing and costing methodologies from other cost-oriented reports, such as:

- The Cost Digest: Cost Summaries of Selected Environmental Control Technologies [2]
- A Standard Procedure for Cost Analysis of Pollution Control Operations [3]
- Handbook: Control Technologies for Hazardous Air Pollutants [4]

Although these reports (as well as many of the MACT Background Information Documents and other standards-supporting documents) contain costs for add-on control systems, they do not duplicate the Manual for one or more of the following reasons: (1) their costs have been based either wholly or partly on data in the previous Manuals; (2) they apply to specific source categories only, whereas the Manual data may be applied generally; (3) their estimating procedures and costs are of less than study estimate quality; or (4) they are not intended for estimating costs used in regulatory development. The Cost Digest, for example, is designed for use by non-technical personnel, contains procedures for making "order-of-magnitude" estimates (± 30% accuracy or worse). A Standard Procedure, conversely, was primarily intended for estimating costs for R&D cases (e.g., demonstration projects), where some site-specific data are available. Further, although the latter report contains a thorough list of equipment installation factors, it contains few equipment costs. The report, Handbook: Control Technologies, used data and estimating procedures from the 1990 Manual to provide sound generalized procedures for estimating costs for various types of control equipment. This edition of the Manual supplements this information. Also, since its inception, the Manual has been extensively used to support Agency regulatory development, state permitting programs, and other activities where current, consistent, and comprehensive control cost data are required.

One additional characteristic of the Manual must also be considered: the Manual is free. While other pollution control cost reports can costs hundreds of dollars, the EPA has always provided this Manual at no cost. This is especially important when we consider the increased use the Manual has received from academic institutions for master's and doctoral work by engineers, environmental engineers, and economists. In summation, the Manual remains a uniquely available, uniquely comprehensive, and uniquely accepted standard in the field of environmental pollution control sizing and costing.

References

[1] OAQPS Control Cost Manual (Fourth Edition), EPA, Office of Air Quality Planning and Standards, Emissions Standards Division, January 1990 (EPA 450/3-90-006 [NTIS/PB90-169954]).

[2] DeWolf, Glenn, et al. (Radian, Inc.), The Cost Digest: Cost Summaries of Selected Environmental Control Technologies, EPA, ORD, Office of Environmental Engineering and Technology, October 1984 (EPA-600/884-010).

[3] Uhl, Vincent W., A Standard Procedure for Cost Analysis of Pollution Control Operations, Volumes I and II, EPA, ORD, Industrial Environmental Research Laboratory, June 1979 (EPA-600/8-79-018a).

[4] Handbook: Control Technologies for Hazardous Air Pollutants, EPA, Office of Research and Development, Air and Energy Engineering Research Laboratory, June 1991 (EPA-625/6-91-014).

EPA/452/B-02-001

Section 1

Introduction

EPA/452/B-02-001

Chapter 2

Cost Estimation: Concepts and Methodology

Daniel C. Mussatti
William M. Vatavuk
Innovative Strategies and Economics Group
Air Quality Strategies and Standards Division
Office of Air Quality Planning and Standards
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

January 2002

Contents

2.1 Introduction 2-3

2.2 Types of Cost Estimates 2-3

2.3 Cost Categories Defined 2-5
2.3.1 Elements of Total Capital Investment 2-5
2.3.2 Elements of Total Annual Cost 2-7

2.4 Financial Concepts 2-10
2.4.1 Time Value of Money 2-11
2.4.2 Interest Rates 2-12
2.4.3 Prices and Inflation 2-13
2.4.4 Financial Analysis 2-15
2.4.4.1 Cash Flow 2-15
2.4.4.2 Payback 2-16
2.4.4.3 Net Present Value 2-18
2.4.4.4 Equivalent Uniform Annual Cash Flow and Annualization 2-20
2.4.4.5 Other Financial Analysis Tools 2-21
2.4.4.6 Economic versus Accounting Costs and Benefits 2-22

2.5 Estimating Procedure 2-23
2.5.1 Facility Parameters and Regulatory Options 2-24
2.5.2 Control System Design 2-25
2.5.3Sizing the Control System 2-26
2.5.4 Estimating Total Capital Investment 2-26
2.5.4.1 General Considerations 2-26
2.5.4.2 Retrofit Cost Considerations 2-28
2.5.5 Estimating Annual Costs 2-30
2.5.5.1 Raw Materials 2-30
2.5.5.2 Labor 2-31
2.5.5.3 Maintenance Materials 2-32
2.5.5.4 Utilities 2-32
2.5.5.5 Waste Treatment and Disposal 2-32
2.5.5.6 Replacement Materials 2-33
2.5.5.7 Overhead 2-34
2.5.5.8 Property Taxes, Insurance, and Administrative Charges 2-34

2.6 Example 2-34

References 2-38

APPENDIX A 2-40

2.1 Introduction

This chapter presents a methodology that will enable the user, having knowledge of the source being controlled, to produce study-level cost estimates for a control system applied to that source. The methodology, which applies to each of the control systems included in this Manual, is general enough to be used with other "add-on" systems as well. Further, the methodology can apply to estimating the costs of fugitive emission controls and for other non-stack abatement methods.

2.2 Types of Cost Estimates

As mentioned in Chapter 1.1, the costs and estimating methodology in this Manual are directed toward the "study" estimate with a nominal accuracy of ± 30% percent. According to Perry's Chemical Engineer's Handbook, a study estimate is "... used to estimate the economic feasibility of a project before expending significant funds for piloting, marketing, land surveys, and acquisition ... [However] it can be prepared at relatively low cost with minimum data."[1] Specifically, to develop a study estimate, the following must be known:

- Location of the source within the plant;
- Rough sketch of the process flow sheet (i.e., the relative locations of the equipment in the system);
- Preliminary sizes of, and material specifications for, the system equipment items;
- Approximate sizes and types of construction of any buildings required to house the control system;
- Rough estimates of utility requirements (e.g., electricity);
- Preliminary flow sheet and specifications for ducts and piping;
- Approximate sizes of motors required.[1]

In addition, the user will need an estimate of the labor hours required for engineering and drafting activities because the accuracy of an estimate (study or otherwise) depends on the amount of engineering work expended on the project. There are four other types of estimates, three of which are more accurate than the study estimate. Figure 2.1 below, displays the relative accuracy of each type of cost estimation process. The other processes are:[1]

- Order-of-magnitude. This estimate provides "a rule-of-thumb procedure applied only to repetitive types of plant installations for which there exists good cost history". Its error bounds are greater than ± 30%. (However, according to Perry's, "... no limits of accuracy can safely be applied to it.") The sole input required for making this level of estimate is the control system's capacity (often measured by the maximum volumetric flow rate of the gas passing through the system).

- Scope, Budget Authorization, or Preliminary. This estimate, nominally of ± 20% accuracy, requires more detailed knowledge than the study estimate regarding the site, flow sheet, equipment, buildings, etc. In addition, rough specifications for the insulation and instrumentation are also needed.

- Project Control or Definitive. These estimates, accurate to within ± 10%, require yet more information than the scope estimates, especially concerning the site, equipment, and electrical requirements.

- Firm or Contractor's or Detailed. This is the most accurate (± 5%) of the estimate types, requiring complete drawings, specifications, and site surveys. Consequently, detailed cost estimates are typically not available until right before construction, since "time seldom permits the preparation of such estimates prior to an approval to proceed with the project."[1]

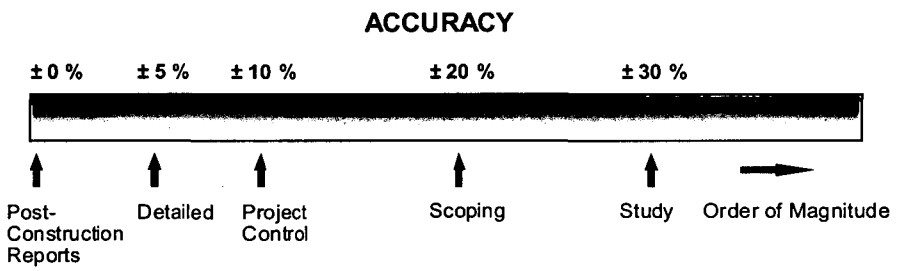


Figure 2.1: The Continuum of Accuracy for Cost Analyses

Study-level estimates are acceptable for regulatory development because they represent a compromise between the less accurate order-of-magnitude and the more accurate estimate types. The former is too imprecise to be of much value, while the latter are not only very expensive to make, but require detailed site and process-specific knowledge that most Manual users will not have. Over time, this Manual has become the standard for air pollution control costing methodologies for many regulatory agencies. For example, Virginia requires that the Manual to be used in making cost estimates for BACT and other permit applications, unless the permit applicant can provide convincing proof that another cost reference should be used.[1]

When used by industry to plan for the installation of a pollution control device, this Manual offers the user an opportunity for greater accuracy than that used by regulators. Since the industrial user will necessarily have much more detailed information than the generic cost and sizing information used in a study estimate, the methodology employed by this Manual can provide cost estimates

[1] Correspondence with William Vatavuk, former editor and author of the Manual, 12/24/01.

that approach those of a scoping study. However, the EPA does not claim cost estimates for industry at a greater than study level accuracy for industrial users, even though the anecdotal evidence from most testimonials volunteered by industrial users indicate a much greater than ±30 percent accuracy can be attained.

2.3 Cost Categories Defined

The terminology used in the earlier editions of this Manual were adapted from the American Association of Cost Engineers.[2]. However, different disciplines give different names to the same cost components and the objective of this edition is to reach out to a broader scientific audience. For example, engineers determine a series of equal payments over a long period of time that fully funds a capital project (and its operations and maintenance) by multiplying the present value of those costs by a capital recovery factor, which produces an Equivalent Uniform Annual Cash Flow (EUAC) value. This is identical to the process used by accountants and financial analysts, who adjust the present value of the project's cash flows to derive an annualized cost number.

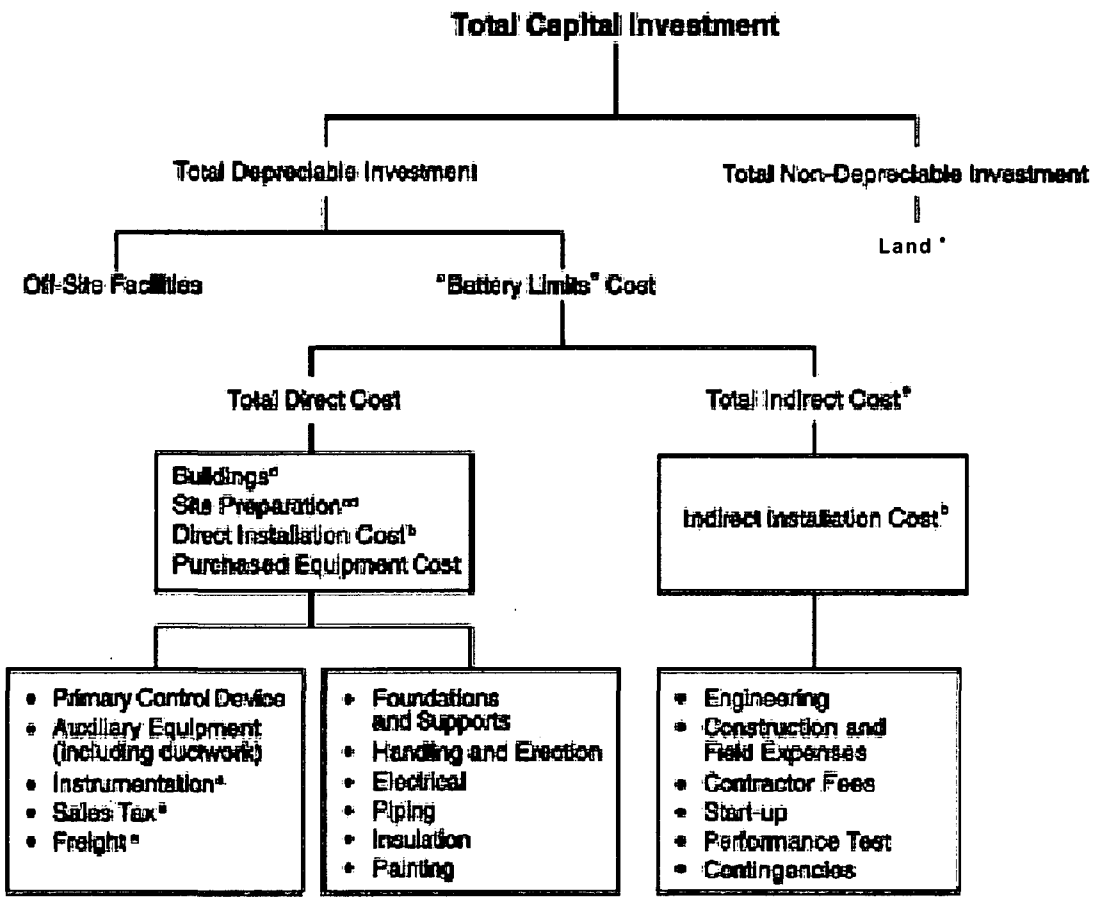
2.3.1 Elements of Total Capital Investment

Total capital investment (TCI) includes all costs required to purchase equipment needed for the control system (purchased equipment costs), the costs of labor and materials for installing that equipment (direct installation costs), costs for site preparation and buildings, and certain other costs (indirect installation costs). TCI also includes costs for land, working capital, and off-site facilities.

Direct installation costs include costs for foundations and supports, erecting and handling the equipment, electrical work, piping, insulation, and painting. Indirect installation costs include such costs as engineering costs; construction and field expenses (i.e., costs for construction supervisory personnel, office personnel, rental of temporary offices, etc.); contractor fees (for construction and engineering firms involved in the project); start-up and performance test costs (to get the control system running and to verify that it meets performance guarantees); and contingencies. Contingencies is a catch-all category that covers unforeseen costs that may arise, such as "... possible redesign and modification of equipment, escalation increases in cost of equipment, increases in field labor costs, and delays encountered in start-up."[2] Contingencies are not the same thing as uncertainty and retrofit factor costs, which are treated separately in this chapter.

The elements of total capital investment are displayed in Figure 2.2. Note that the sum of the purchased equipment cost, direct and indirect installation costs, site preparation, and buildings costs comprises the battery limits estimate. By definition, this is the total estimate "... for a specific job without regard to required supporting facilities which are assumed to already exist..."[2] at the plant. This would mainly apply to control systems installed in existing plants, though it could also apply to those systems installed in new plants when no special facilities for supporting the control system (i.e., off-site facilities) would be required. Off-site facilities include units to produce steam,

electricity, and treated water; laboratory buildings; and railroad spurs, roads, and other transportation infrastructure items. Pollution control systems do not generally have off-site capital units dedicated to them since pollution control devices rarely consume energy at that level. However, it may be necessary—especially in the case of control systems installed in new or "grass roots" plants—for extra capacity to be built into the site generating plant to service the system. (A venturi scrubber, which often requires large amounts of electricity, is a good example of this.) Note, however, that the capital cost of a device does not include utility costs, even if the device were to require an off-site facility. Utility costs are charged to the project as operating costs at a rate which covers both the investment and operating and maintenance costs for the utility. Operating costs are discussed in greater detail below.



[a]Typically factored from the sum of the primary control device and auxiliary equipment costs.
[b]Typically factored from the purchases equipment cost.
[c]Usually required only at "grass roots" installations.
[d]Unlike the other direct and indirect costs, costs for these items usually are not factored from the purchased equipment cost. Rather, they are sized and costed separately.
[e]Normally not required with add-on control systems.

Figure 2.2: Elements of Total Capital Investment

As Figure 2.2 shows, the installation of pollution control equipment may also require land, but since most add-on control systems take up very little space (a quarter-acre or less) this cost would be relatively small. Certain control systems, such as those used for flue gas desulfurization (FGD) or selective catalytic reduction (SCR), require larger quantities of land for the equipment, chemicals storage, and waste disposal. In these cases, especially when performing a retrofit installation, space constraints can significantly influence the cost of installation and the purchase of additional land may be a significant factor in the development of the project's capital costs. However, land is not treated the same as other capital investments, since it retains its value over time. The purchase price of new land needed for siting a pollution control device can be added to the TCI, but it must not be depreciated, since it retains its value forever. Instead, if the firm plans on dismantling the device at some future time, then the land should be either excluded from the analysis, or the value of the land should be included at the disposal point as an "income" to the project to net it out of the cash flow analysis (more on cash flow analyses later, in section 2.4.4.1).

One might expect initial operational costs (the initial costs of fuel, chemicals, and other materials, as well as labor and maintenance related to start-up) should be included in the operating cost section of the cost analysis instead of in the capital component, but such an allocation would be inappropriate. Routine operation of the control does not begin until the system has been tested, balanced, and adjusted to work within its design parameters. Until then, all utilities consumed, all labor expended, and all maintenance and repairs performed are a part of the construction phase of the project and are included in the TCI in the "Start-Up" component of the Indirect Installation Costs.

2.3.2 Elements of Total Annual Cost

Total Annual Cost (TAC) has three elements: direct costs (DC), indirect costs (IC), and recovery credits (RC), which are related by the following equation:

$$TAC = DC + IC - RC \tag{2.1}$$

Clearly, the basis of these costs is one year, as this period allows for seasonal variations in production (and emissions generation) and is directly usable in financial analyses. (See Section 2.3.) The various annual costs and their interrelationships are displayed in Figure 2.3.

Direct costs are those that tend to be directly proportional (variable costs) or partially proportional (semi-variable costs) to some measure of productivity - generally the company's productive output, but for our purposes, the proper metric may be the quantity of exhaust gas processed by the control system per unit time. Conceptually, a variable cost can be graphed in cost / output space as a positive sloped straight line that passes through the origin. The slope of the line is the factor by which output is multiplied to derive the total variable cost of the system. Semi-variable costs can be graphed as a positive sloped straight line that passes through the cost axis at a value greater than zero - that value being the "fixed" portion of the semi-variable cost and the slope of the line being analogous to that of the variable cost line discussed above.

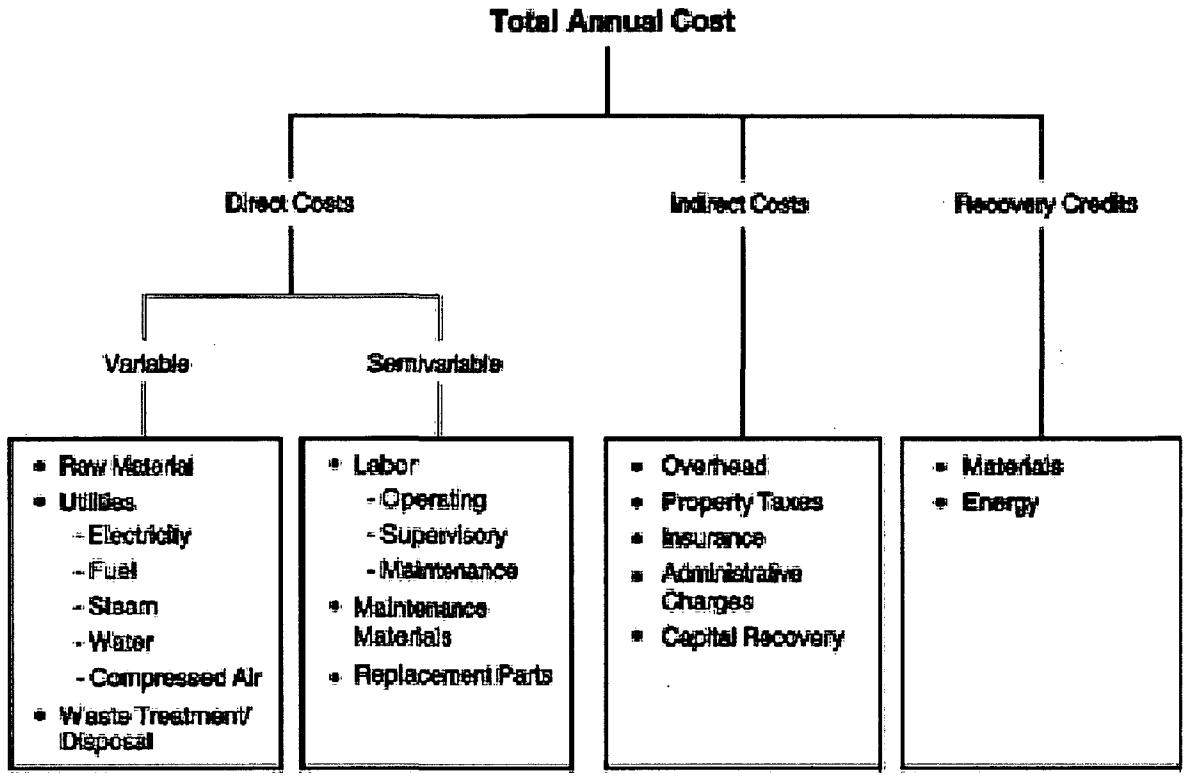


Figure 2.3: Elements of Total Annual Cost

In the graphical representation of variable and semi-variable costs in Figure 2.4, the blue line (lower line) indicates a variable cost function, with all of its value directly related to the level of output on the X-axis. At zero output, the variable cost function returns a variable cost of zero, as well. Alternatively, the upper, red line in Figure 2.4 shows a semi-variable cost, where even at an output level of zero, the system will still incur a cost of $50. The key difference between the two types of variable cost is that at an output level of zero, a semi-variable cost will still exist. An example would be a boiler producing process steam for only sixteen hours a day. During the time the boiler is idle, it costs less to keep the boiler running at some idle level than to re-heat it at the beginning of the next shift. Consequently, that idle level operation cannot be attributed to production and should be considered the fixed component of the semi-variable fuel cost of the boiler. Direct costs include costs for raw materials (reagents or adsorbers), utilities (steam, electricity, process and cooling water), waste treatment and disposal, maintenance materials (greases and other lubricants, gaskets, and seals), replacement parts, and operating, supervisory, and maintenance labor. Generally, raw materials, utilities, and waste treatment and disposal are variable costs, but there is no hard and fast rule concerning any of the direct cost components. Each situation requires a certain level of insight and expertise on the part of the analyst to separate out the cost components accurately.

Capital is depreciable, indicating that, as the capital is used, it wears out and that lost value cannot be recovered. Depreciation costs are a variable or semi-variable cost that is also included in the calculation of tax credits (if any) and depreciation allowances, whenever taxes are considered in a cost analysis. (However, taxes are not uniformly applied, and subsidies, tax moratoriums, and deferred tax opportunities distort how the direct application of a tax works.

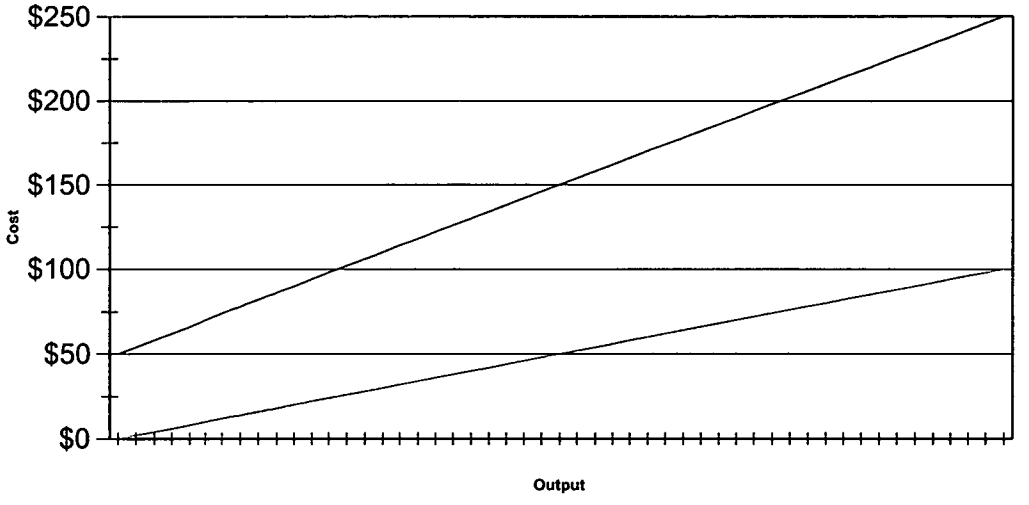


Figure 2.4: Graphical Comparison of Variable and Semi-variable Costs

Therefore, this Manual methodology does not consider income taxes.) Notice that when 100% of the system costs are depreciated, no salvage value can be taken for the system equipment at the conclusion of its useful life. This is a reasonable assumption for add-on control systems, as most of the equipment, which is designed for a specific source, cannot be used elsewhere without modifications. Even if it were reusable, the cost of disassembling the system into its components (i.e., "decommissioning cost") could be as high (or higher) than the salvage value. If a salvage value exists and will be recouped at the end of the useful life of the control, then that value must be included in the analysis. The exercise discussed later in this chapter employs a salvage value to illustrate its proper use.

Indirect, or "fixed", annual costs are independent of the level of production (or whatever unit of measure serves as the analytical metric) and, in fact, would be incurred even if the control system were shut down. Indirect costs include such categories as administrative charges, property taxes, insurance, and capital recovery. A fixed cost function added to Figure 2.4 would be a horizontal line appearing at the level of the fixed cost.

Finally, direct and indirect annual costs can be offset by recovery credits, taken for materials or energy recovered by the control system, which may be sold, recycled to the process, or reused elsewhere at the site. An example of such credits is the by-product of controlling sulfur with an FGD. As the lime or limestone reagent reacts with the sulfur in the exhaust gas stream, it becomes transformed into CaSO4 - gypsum - which can be landfilled inexpensively (a direct cost) or collected and sold to wallboard manufacturers (a recovery credit). These credits, must be calculated as net of any associated processing, storage, transportation, and any other costs required to make the recovered materials or energy reusable or resalable. Great care and judgement must be exercised in assigning values to recovery credits, since materials recovered may be of small quantity or of doubtful purity, resulting in their having less value than virgin material. Like direct annual costs, recovery credits are variable, in that their magnitude is directly proportional to level of production. A more thorough description of these costs and how they may be estimated is given in Section 2.4.

2.4 Financial Concepts

Engineers use a relatively small set of financial tools to assess alternative capital investments and to justify their selections to upper management. Most often, the engineer's purpose is to show how the recommended investment will improve the company's profitability. To a great extent, this sort of decision is voluntary. Adding a new assembly line or changing from one type of gasket material to another can be postponed or even rejected. This is not the case with pollution control devices, which are necessary for compliance with State and Federal pollution standards and generally have a deadline attached to their installation. Consequently, a decision to install device X may not originate with the engineer. Instead, the process may actually work backward, relative to the profitability motivated assessment of the engineer: the company's environmental manager could identify the need for pollution control equipment and then pass that decision on down to the engineer.

When air quality regulations limit the source's choice to only one control type, (e.g., when the regulation specifies the technology to be used), this Manual serves two functions. First, it ensures as complete a cost profile as possible has been taken for planning purposes. Second, identification of the appropriate control technology does not include site-specific requirements that need to be identified and costed out. This Manual provides engineering data for the proper sizing and design specification of the control. When the environmental manager can choose between alternative control technologies to achieve the same pollution abatement requirement, this Manual performs a third function by "normalizing" the financial data from each alternative so that a well reasoned selection can be made.

To fully assess the cost of a pollution control device, the reader must understand several financial analysis concepts. This section of the Manual discusses how these concepts fit together to provide the analyst with insight into the cost and selection of alternative pollution control equipment. Earlier editions of the Manual focused on the assessment of financial information from an engineering perspective. However, EPA has learned the audience that uses the Manual extends well beyond

the engineering discipline. Consequently, one of the key changes in this edition of the Manual involves adjusting the financial and economic data to reach that broader audience. Wherever earlier editions included engineering-specific names for financial terms, the engineering name remains, but the technical term from a financial perspective has been included, as well.

Section 2.4.1 discusses how the value of money changes over time and how that led to the derivation of interest and discount rates. Section 2.4.2 discusses the three kinds of interest rates that are important to this Manual, and how to select the right interest rate for your analysis. Section 2.4.3 describes how persistent increases in the general level of prices (inflation) are handled in this edition of the Manual and how to select and use a price index to translate the prices in this edition (2001 dollars) to your future analyses.

2.4.1 Time Value of Money

The costs and benefits of an investment occur over an extended period of time rather than at the moment of purchase. Consequently, financial analyses and benefit-cost studies must accommodate the future effects of current decisions. If individuals placed the same value on a dollar in the future that they placed on a dollar in the present, financial analysis could be simplified to the summation of all future costs and incomes derived from the investment. However, as the old saying goes: "A bird in the hand is worth two in the bush". Not only could the promise of a future dollar go unfulfilled, but the purchasing power of that dollar could decline. Furthermore, spending the dollar in the present offers immediate rewards that have to be postponed if the dollar is withheld until some future date. Therefore, individuals demand compensation to offset these concerns, thereby increasing the value of a future payment to more than a dollar. Conversely, to take payment today on a dollar promised for some time in the future, that same person will accept less than a full dollar because they could enjoy its benefits immediately without the risks of inflation or non-payment. This adjustment process is called the principle of the time value of money.

Adding more time to the delay of payment has a cumulative effect. For example, if an investment required an adjustment of ten percent for each year the decision maker has to postpone collection, a dollar would have to return \$1.10 at the end of the first year (\$1.00 times 110%), and \$1.21 for the decision maker to wait two years (\$1.00 times 110% times 110%). The formula for calculating the future value of a dollar invested today is:

$$FV = \$1 \times (1 + i)^n \tag{2.2}$$

where FV is the future value of the dollar invested, i is the interest rate, and n is the number of interest rate periods (typically years) before the investment has matured. Analogously, discounting future payments to the present has the same "accumulative" effect. For example, if a person wanted to be paid immediately, rather than wait one year for payment of a dollar, (at the same ten percent interest rate used above), they would be willing to accept \$0.92 (\$1 divided by 110%). To be paid

immediately for a dollar promised in two years, they would be willing to accept $0.83 ($1 divided by [110% times 110%]). The present value (PV) of a future dollar realized in year n can be calculated by the following formula:

$$PV = \frac{\$1}{(1+i)^n} \tag{2.3}$$

2.4.2 Interest Rates

Analysts use the interest rate to estimate the time value of money. It can be thought of as a return on investment or the cost of borrowing. A discount rate is an interest rate used to estimate the value of current payments in lieu of waiting until some time in the future. There are three types if interest rates that are important for this Manual: real, nominal, and social. The interest rate stated by lending institutions is a nominal interest rate. It is the cost of borrowing and the lender will have included in it a factor to account for anticipated changes in the general level of prices (inflation). Removing the inflation adjustment from the nominal interest rate yields the real rate of interest - the actual cost of borrowing. For example, say an investor borrows $100 at 10% from a bank for one year. At the end of the year, the investor must pay back $110 dollars. However, if during that year the inflation rate was six percent, the bank may receive $10 in interest, but it takes $106 to equal the purchasing power of the $100 loaned out the previous year. Consequently, it only made $4 in real interest. In equation form, the nominal interest rate (i) equals the real interest rate (i_r) plus the expected rate of inflation (p^e):

$$i = i_r + p^e \tag{2.4}$$

The interest rate employed in this Manual differs from that used in non-governmental financial analyses. It represents a social interest rate established by the Office of Management and Budget (OMB) for the comparison of public policy issues. Like a nominal or real interest rate, a social rate of interest compensates for the foregone benefits associated with spending a dollar today; but for slightly different reasons. Society as a whole has a collective rate of time preference that equates the value of future benefits with an equivalent level of benefits enjoyed now. This rate of preference (interest) would be the same as that which the market would assign to a completely riskless investment. In practice, that riskless investment is represented by the long-term interest rate on government bonds and securities. When determined in this manner, the social rate of discounting should be appropriate for the evaluation of social projects. However, the process is not that simple. Private funds bankroll public projects, and tax effects distort the true cost of borrowing. Furthermore, government securities play a macroeconomic role in the maintenance of

[1]This assumption introduces very little error.

the overall economy, fluctuating with the supply of money to stimulate and impede spending as conditions warrant. (For example, at this writing, the 30-year Treasury Bill rate of interest has fluctuated between 4.5 and 5.5 percent within the last 30 days.) Therefore, the interest rate of government securities may be a starting point, but they are not the same as the social discount rate.

In social terms, spending a dollar today on pollution control equipment means not making investments that could have provided immediate improvements to social welfare. For instance, the installation of a fabric filter baghouse on a coal-fired boiler will reduce the amount of particulate matter (*PM*) emitted into the atmosphere, but the steel used to build the baghouse could have been used to expand the factory or make ambulances or fire trucks. The social discount rate (a real interest rate) measures these kinds of foregone alternative uses. Unlike the return on investment for industry, which varies across industries (and even across firms within an industry), the foregone social benefits associated with an investment remain the same across industries. Also, since a change in the general level of prices affects everyone simultaneously, social rates of interest do not account for inflation. OMB sets the social interest rate for governmental analyses, and it is currently set at seven percent.[1]

When State, local, Tribal and other governmental authorities assess pollution control costs, the seven percent interest rate employed in this Manual should produce estimations comparable to those established by the Agency when it performs its own evaluations. However, the social rate of interest is probably not appropriate for industry. When choosing between alternative air pollution control devices, the industrial planner must not only take into consideration the costs of each device, they must also understand how the cost of each device fits into the financial structure of their business. Furthermore, a number of air regulations allow sources of pollution to petition for extensions on deadlines, variances from the regulation, or exemption from installing control devices, based upon the economic impact that equipment would have upon the source. In these cases, the source may find it useful to apply their own interest rate to the calculation of control costs. Common interest rates used by industry and accepted by the EPA for source petitions include the business' current borrowing rate, the current prime rate, and other acceptable industrial rates of return. Because industry may use an interest rate different from the EPA's seven percent social rate for its calculations, EPA facilitates the application of a customized interest rate in each chapter of the Manual by providing detailed explanations of all formulas and by allowing users of the CO$T-AIR spreadsheets and the Air Compliance Advisor program with the ability to change their interest rates to suit their specific situation.

2.4.3 Prices and Inflation

The prices in the Manual were not standardized. Some chapters had prices developed in the late 1990s, and other chapters had prices developed from as far back as 1985. Because these differences were not explicitly discussed in these earlier additions, the Agency attempted to

---

[1] www.epa.gov/ttn/catc/products.html#cccinfo

[2] U.S. Office of Management and Budget, Circular A-94: Guidelines and Discount Rates for Benefit Cost Analysis of Federal Programs. October 29, 1992. Prior to 1992, the OMB-determined social discount rate was 10%.

standardize all prices in this sixth edition of the Manual to reduce the chance for analytical error. Over the past two years, new values for equipment costs were developed by re-surveying affected industries and vendors. This effort updated all the costs to at least 1990.

To develop the costs used in each of the chapters of this Manual, we surveyed the largest possible group of vendors to determine an industry average price for each cost component. In many cases, this involved contact with hundreds of vendors and the assimilation of large amounts of data. In other cases, the pollution control equipment was supplied by only a few vendors, which limited the independence of our models. And, in still other cases, the number of existing manufacturers or the highly site-specific nature of their installation made it difficult for us to develop completely unbiased prices for some components.

Updating costs is an on-going effort at EPA with a goal of standardizing all costs to one base year. Each chapter of the Manual fully discloses the limitations of the costing information found in that chapter. This allows the analyst may make any adjustment they deem necessary.

Real and nominal prices act in the same way as real and nominal interest rates. Nominal prices are actual prices (i.e., the sticker price) and represent the value of a particular good at a particular point in time. Real prices remove the effect of inflation. Adjusting nominal prices to real prices involves establishing a base year for comparison purposes and then creating an adjustment factor for each year's prices relative to those in the base period. This adjustment factor is a price index (PI) that can then be used to adjust nominal prices to an equivalent base year value; derived through the following formula:

$$PI = \frac{price\ in\ given\ year}{price\ in\ base\ year} \qquad (2.5)$$

The Federal government and industry develop a variety of indexes tailored to the analysis of specific price issues. The most recognizable of these indexes are the Consumer Price Index (CPI) and the Producer Price Index (PPI), which investigate the change in prices across the entire economy. However, these indexes are often too general for the specific needs of industry. For pollution control purposes, OAQPS has developed and maintained the Vatavuk Air Pollution Control Cost Indexes (VAPCCI) which provides an estimate of the change in prices for the purchase of pollution control devices. The VAPCCI can be found on the Internet, at the OAQPS web site and the Technology Transfer Network (TTN) web site. [1] Other indexes are also available from industry and academic sources through the Internet, industry publications, trade journals, and financial institutions. When choosing the right price index for your analysis, employ the "ABC Principle" - that the index is Accepted by industry or financial institutions, it is Bias-free, and it is Conservative. Bias is a statistical sampling or testing error caused by systematically favoring some outcomes over others. It is a reflection of the judgement and opinion of the analyst but is not the same thing as "professional judgement." Bias most commonly appears as a "self-fulfilling prophesy" that incorrectly validates an incorrect assumption on the analyst's part. "Conservative" often gets interpreted as "careful" - an intentional over- or under-estimation of the actual value to avoid the

negative effects of errors. Selection of an index employing these standards will most likely not cause estimation error or be challenged by persons or agencies overseeing your work.

2.4.4 Financial Analysis

Once you have amassed all of the necessary information on the design, installation, operation, and revenue of a possible capital investment, what does that tell you about how that investment will affect the overall financial health of your firm? When comparing two different investment opportunities, how do you distill all of these data into one comprehensive and coherent form so that an informed decision can be made? This section deals with a number of the concepts and operations that will help you answer these questions.

When alternative investment opportunities exist - or, for our purposes, when more than one pollution control device may be used - the selection of the most appropriate alternative depends on that alternative's effect on the firm's profitability. Consequently, financial analysts have created a set of tools that provide insight into the potential financial consequences associated with an investment. While no single tool works in all instances, applying several of these tools can provide the financial manager with sufficient insight for a meaningful decision to be made.

Survey evidence indicates most analysts use more than one tool to make financial decisions.[2] The remainder of this section discusses each of these tools and describes their relative strengths and weaknesses. The most fundamental analysis needed is that of cash flow, which formalizes the expected inflows of revenue and outflows of expenses associated with an investment alternative. Pollution control devices do not typically generate revenues, but environmental cost accountants still begin their evaluation of pollution control alternatives through cash flow analysis as a precursor to the application of other tools.[3] The next section discusses cash flow analysis and how it applies to pollution control equipment. Probably the most important tool in the analyst's arsenal is net present value (*NPV*) since it acts as the foundation for a number of related analyses, including benefits/cost analysis. The sections after cash flow analysis discuss these common financial analysis tools. At the end of this chapter is an exercise that applies these concepts.

2.4.4.1 Cash Flow

Incomes and expenditures take place over the life of an investment (its planning horizon), the amounts and timing of which constitute the cash flows of the project. Pollution control system costing always includes expenditures but may not necessarily have incomes. For a control to be income generating, it must reduce production cost (through fewer inputs or product reformulation), or it must capture and recover a pollutant with recyclable characteristics (e.g., solvent recovery).

---

[2] c.f., R.S. Kaplan and A.A. Atkinson, Advanced Management Accounting, 2nd ed., Engelwood Cliffs, NJ: Prentice Hall, 1989.

[3] U.S.Environmental Protection Agency, Environmental Cost Accounting for Capital Budgeting: A Benchmark Survey of Management Accounting, #EPA742-R-95-005. Washington D.C., U.S. EPA Office of Pollution Prevention and Toxics, 1995.

For illustration purposes, consider a hypothetical series of cash flows for a project with an operational life of ten years. The data for this figure can be found in Table 2.1, below and will be used as the basis for further financial discussion to follow.

Table 2.1: Hypothetical Cash Flow

	Year										
	0	1	2	3	4	5	6	7	8	9	10
Expenses	-250	-33	-30	-30	-30	-30	-30	-30	-32	-34	-39
Revenues	0	50	65	65	65	65	65	65	62	58	50
Net Cash Flow	-250	17	35	35	35	35	35	35	30	24	11

Figure 2.5 is the cash flow information in graphical form. Expenses are solid red bars extending below the line and incomes are solid blue bars above it. Figure 2.5 displays net cash flow - the difference between incomes and expenses - as white bars on the graph. Typical of many equipment-related cash flows, the greatest cost occurs at time zero, when the control is purchased and installed. In the first years of operation, costs tend to be relatively high for operating and maintaining new equipment, due to balancing and breaking-in conditions. After that, costs tend to drop and remain fairly constant until the equipment approaches the end of its useful life, when operations and maintenance costs tend to rise again. In the hypothetical example, the control generates income that offsets the costs of operations and maintenance during the life of the equipment. This is not the typical situation for pollution control devices, but is used in this hypothetical example because it allows us to illustrate other financial concepts.

2.4.4.2 Payback

Probably the simplest form of financial analysis is the payback period analysis, which simply takes the capital cost of the investment and compares that value to the net annual revenues that investment would generate. If net annual revenues are the same every year, the revenue can simply be divided into the total capital investment to calculate the payback period. If the annual net revenues differ, then the values need to be summed sequentially until the revenue exceeds total capital investment. The payback decision rule is to select that investment with the shortest payback time. For instance, consider the hypothetical example above. It has a capital investment of $250, and costs between $30 and $39 per year to operate and maintain. The project also has an expected revenue generating capacity of between $50 and $65 per year, for annual net revenues of between $11 and $35 per year. If all revenues and expenses were equal, say to the steady state values of $65 and $30, respectively, the project would

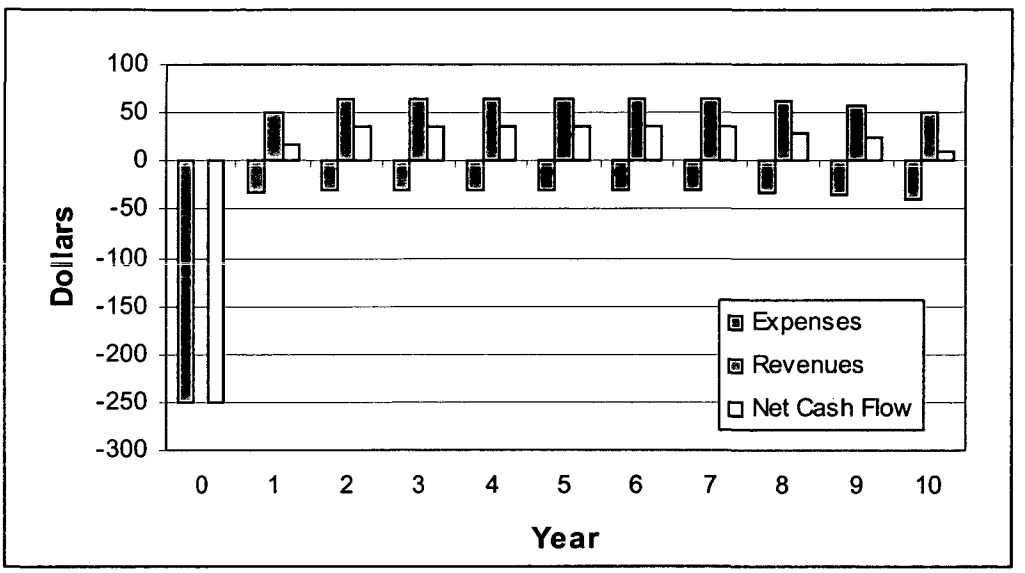


Figure 2.5: Hypothetical Cash Flow Diagram

have had a net annual revenue of $35 per year, suggesting the investment would "pay itself back" in a little more than seven years ($250 divided by $35/yr). However, with uneven revenues and costs, we have to sum the net revenues to determine when they exceed the initial capital cost - in year eight, for the hypothetical example.

For most pollution control devices, payback analysis is not possible because the device does not produce revenue. However, for a limited number of devices (FGD units, VOC recovery devices, etc.), the device may produce a saleable product that produces a revenue stream. In these cases, payback is a very limited tool and offers only the grossest of estimates with regard to relative profitability, for the following reasons. First, payback ignores the magnitude and direction of cash flows in all of the years in the planning horizon beyond the payback period. A project that paid for itself in five years and produced revenues in all years after payback would have the same payback value as one that paid back in the same time yet incurred huge losses in all subsequent years!

Second, payback does not take into account the time value of money. Consider the example above. Applying the social discount rate of seven percent, the capital investment will never pay for itself. The net revenues for the ten years of the payback period in real "year zero" dollars are $15.89, $30.57, $28.57, $26.70, $24.95, $23.32, $21.80, $17.46, $13.05, $5.59, respectively, leaving the payback of investment short in real terms by more than 20 percent! [1]

Next, even if the payback analysis is performed in real terms rather than in nominal terms,

---

[1] Apply the discounting factors found in Appendix A at the end of this chapter to derive these values.
[2] The author invites the reader to construct the appropriate cash flow analyese spreadsheets as an exercise.

the process still fails when comparing dissimilar projects. Payback considers all projects with equal payback periods to be equally feasible, regardless of other characteristics. Referring again to our hypothetical example, assume the equipment has a useful life of ten years and a payback value of seven. Furthermore, consider a second alternative investment, again with a payback period of seven years, but now the equipment has a useful life of only seven years. For simplicity sake, assume all years for both investments have a net revenue of $35.[2] While payback considers both projects to be equal, the second project would actually cost more over the ten year period because it would have to be replaced at the end of its useful life. Therefore, when using the payback method, the analyst must be sure to standardize all of the alternatives to the length of the longest planning horizon among the choices.

2.4.4.3 Net Present Value

To evaluate alternative pollution control devices, the analyst must be able to compare them in a meaningful manner. Since different controls have different expected useful lives and will result in different cash flows, the first step in comparing alternatives is to normalize their returns using the principle of the time value of money discussed in section 2.4.1. The process through which future cash flows are translated into current dollars is called present value analysis. When the cash flows involve income and expenses, it is also commonly referred to as net present value analysis. In either case, the calculation is the same: adjust the value of future money to values based on the same (generally year zero of the project), employing an appropriate interest (discount) rate and then add them together. The decision rule for NPV analysis is that projects with negative NPVs should not be undertaken; and for projects with positive NPVs, the larger the net present value, the more attractive the project.

Derivation of a cash flow's net present value involves the following steps:

- Identification of alternatives - for example, the choice between a fabric filter bag house and an electrostatic precipitator (ESP) for removing particulate matter (PM) from a flue gas stream.

- Determination of costs and cash flows over the life of each alternative - each of the subsequent chapters of this Manual offer detailed costing information on specific air pollution control devices and equipment, and the supplemental programs that support the Manual, the Air Compliance Advisor (ACA) program, and the CO$T-Air spreadsheets provide the same costing information electronically.

- Determination of an appropriate interest (discount) rate - for States, local, Tribal, and other environmental management organizations, the EPA's seven percent social discount rate will probably be the most appropriate. Industrial users of this Manual should consult with their financial officers and / or trade association for input. Section 2.3.2 in this chapter, discusses some of the issues that govern industry's choice of an appropriate interest rate. If no private defendable discount rate can be identified, then the industry analyst should fprobably use the Agency's social rate for its own analyses.

- For each alternative: Calculate a discounting factor for each year over the life of the equipment - For example, the EPA's seven percent discount rate produces discount factors of: 0.9346, 0.8734, 0.8163, 0.7629, and 0.7130 for the 1st, 2nd, 3rd, 4th, and 5th years of a piece of equipment's life, respectively. Table A.1 in Appendix A displays discount factors for interest rates from 5.5 to 15 percent, in half-percent increments for 25 years.

- For each year's cash flows, sum all incomes and expenses to determine the net cash flow for that year in nominal terms (see section 2.4.4.1).

- Multiply each years' net cash flow by the appropriate discount factor.

- Sum the discounted net cash flows to derive the net present value.

- Compare the net present values from each alternative. Higher net present values indicate better investment opportunities, relative to the other alternatives in the decision set.

The net present value of a stream of cash flows over the life of an investment can be calculated using equation 2.6:

$$NPV = \sum_{t=0}^{n} NCF_t \left(\frac{i}{1-(1+i)^{-t}} \right) \qquad (2.6)$$

where NCF_t represents the net cash flow for year t, and i is the interest rate.

However, net present value analysis has limitations. Consider a hypothetical firm investigating the installation of two alternative air pollution controls. One alternative would be to retrofit existing equipment that has only five years of useful life remaining. The other strategy would be to salvage existing equipment and replace it with new state-of-the-art components that pollute less. This strategy would have a useful life of 20 years. If the retrofitting process has a higher net present value, does that mean it is the better choice? Not necessarily, because the new equipment would have to be purchased in five years, anyway, resulting in two sets of investments. If the new equipment strategy had a higher NPV, would it be a better choice? Not necessarily, since the firm would have to scrap existing equipment to install the new system. Furthermore, timing the installation of the state-of-the-art controls with the replacement of the fully depreciated equipment postpones the investment for five years (allowing it to decline in value in real terms) and avoids scrapping equipment that is still useful.

One solution to the timing problem described above would be to extend the retrofit alternative by adding 15 years' worth of equipment replacement at the end of the retrofit period. This would provide a more meaningful comparison, since it would be for similar time periods and would also include the two expenditures for equipment necessary for the retrofit to occur. Another solution would be to augment the information received from net present value analysis by employing other financial analysis tools.

Calculating NPV is not difficult, but it does require some subjective decision-making by the analyst to establish the appropriate discount rate or rates to apply. Since market forces typically incorporate inflation adjustments into investment returns and borrowing costs, often the discount rate is keyed to a standard reference rate. As with choosing a cost index, when choosing the appropriate rate for a present value analysis, apply the "ABC Principle" - that the rate is Accepted, Bias-free, and Conservative. As a general rule of thumb, the higher the discount rate, the less expensive (more discounted) the present value will be. Therefore, the ideal discount rate would be one that which matches the highest rate of return the firm can obtain by investing, or the lowest rate at which it can borrow money, whichever is higher. If the analyst chooses a discount rate lower than this "ideal" rate, then the net present value of the investment will appear high, relative to other investment opportunities and the firm will run the risk of taking on an investment decision that has been artificially inflated to make it more attractive. Conversely, if the discount rate chosen is too high, then the net present value of the investment will appear unfavorable, relative to other investment opportunities and the firm risks rejecting a potentially favorable investment. The same arguments hold for pollution control decisions. When applying for a permit to build or operate a source of pollution, the application of an inappropriate discount rate can lead to erroneous petitions for waivers or variances. Therefore, the EPA pays particular attention to the interest / discount rates used in operating and new source review permits.

2.4.4.4 Equivalent Uniform Annual Cash Flow and Annualization

When you purchase a new home, you do not immediately pay for the cost of construction, maintenance, upkeep, and all of the other expenses you will incur over the next thirty years. Instead, you probably borrowed the money from a lender to make the initial purchase and will pay annual expenses as the occur. Net present value analysis allows us to evaluate between investments by summing the present value of al future incomes and expenses, but that does not give us an insight into the expected cash flows that will actually occur. So, instead of paying up front for all the future costs of installation, maintenance, and operation of a pollution control device (NPV analysis), what if the payments could be equalized (in constant net present value dollars) over the life of the control? A common engineering tool for this sort of evaluation is called the equivalent uniform annual cash flow (EUAC) method.[3] EUAC works best when the is only one capital investment to incorporate and annual cash flows are constant or normalized to one year, typically year zero. When comparing EUACs for competing systems, analysts should avoid comparing systems with widely differing useful lives. Comparing EUACs for systems with lives differing by two or three years may be reasonable, but beyond that range, comparisons become problematic. This Manual does not recommend the use of EUAC by itself and only when the useful lives of alternative controls are very similar.

Annualization is a process similar to EUAC but is not limited to constant cash flows. It involves determining the net present value of each alternative equipment investment and then determining the equal (in nominal terms) payment that would have to be made at the end of each year to attain the same level of expenditure. In essence, annualization involves establishing an annual "payment" sufficient to finance the investment for its entire life, using the formula:

$$PMT = NPV\left(\frac{i}{1-(1+i)^{-n}}\right) \qquad (2.7)$$

where *PMT* is the equivalent uniform payment amount over the life of the control, *n*, at an interest rate, *i*. *NPV* indicates the present value of the investment as defined above in equation 2.6.

Engineering texts call this payment the capital recovery cost (*CRC*), which they calculate by multiplying the *NPV* of the investment by the capital recovery factor (*CRF*):

$$CRC = NPV \times CRF \qquad (2.8)$$

where *CRF* is defined according to the formula:

$$CRF = \left(\frac{i\,(1+i)^{n}}{(1+i)^{n}-1}\right) \qquad (2.8a)$$

The *CRF* equation is a transformation of the *PMT* form in equation 2.7 and returns the same information. Table A.2 in Appendix A lists the *CRF* for discount rates between 5.5 percent and 15 percent for annualization periods from one to 25 years.

2.4.4.5 Other Financial Analysis Tools

Many firms make investment decisions based upon the return on investment (*ROI*) of the proposed capital purchase, rather than the magnitude of its net present value. In and of itself, the ROI of an investment opportunity is of little use. For most pollution control investments, *ROI* analysis does not provide much in the way of useful information because, like a payback analysis, it must have positive cash flows to work properly. Calculated by dividing annual net income by the investment's capital cost, results in a percentage of the investment that is returned each year. The decision rule one should apply for *ROI* analysis is if the resulting percentage is at least as large as some established minimum rate of return, then the investment would be worth while. However, different industries require different rates of return on investments, and even within an industry, many different rates can be found. Analysts should consult with their firm's financial officers or an industrial association to determine what percentage would apply.

In its simplest terms, internal rate of return (IRR) is a special case of net present value analysis used to separate "good" investment opportunities from "bad". In fact, many trade organizations publish standard IRR rates for their particular industry. Projects with an IRR less than the industry standard should be rejected as not providing sufficient income to make them worth while; and projects with IRRs greater than the industry standard should be considered good investment opportunities. NPV analysis is actually a series of current values, each one associated with a different interest rate. For each interest rate chosen, the NPV of the same investment will differ, increasing from a negative NPV at very low interest rates to a positive NPV at higher rates. For each investment analyzed, the interest rate that results in a net present value of exactly zero is the investment's IRR. However, the application of IRR depends on having positive cash flows, which again limits their use in analyzing pollution control alternatives; but, when there are positive cash flows, IRR can provide useful information.

Twenty years ago, IRR was not easily used because there is no direct method for deriving it. Instead, a project's IRR had to be determined manually by an iterative process that could take many hours to perform. Today, although the mathematical processes behind determining an IRR have not changed, the convenience of computers have made it much easier to perform. Most spreadsheet programs available today offer an IRR calculator within their financial tools. One of the biggest problems with applying the internal rate of return methodology happens when the relevant cash flow switches between positive and negative. When this occurs it is possible to derive two or more different IRRs for the same project. When that happens, IRR is not applicable for determining the acceptance of independent projects or for identifying the best investment risk out of a group of potential projects.

The benefit-cost ratio of an investment is defined as the ratio of the discounted benefits to the discounted cost, each evaluated at the same constant dollar rate - generally in year zero dollars. With benefits in the numerator of the ratio, the criterion for accepting a project on the basis of the benefit-cost ratio is whether or not the benefit-cost ratio is greater than or equal to one (i.e., benefits are greater than costs). However, as with the payback analysis and financial tools that rely on incomes, benefit cost ratios can be problematic when applied to pollution control devices and evaluated from a strict financial standpoint.

2.4.4.6 Economic versus Accounting Costs and Benefits

From a strict financial standpoint, many of the tools discussed above do not have an application to industry when evaluating pollution control devices. This is not the case when the analysis is being performed at the regulatory level. In these cases, all of the above tools can prove beneficial - provided the analyst includes the appropriate set of costs and benefits (incomes). Clearly, benefit cost analysis is a powerful regulatory tool for evaluating pollution control equipment when assessed from an economic perspective, where the external costs and benefits of the device

---

[2] Retrofit factors for specific applications (coal-fired boiler controls) have been developed. See references [14] and [15].

can be more easily quantified. That is because economic costs are not the same as accounting costs and economic benefits are not the same as accounting benefits.

Accounting costs are those costs included in a financial statement, ledger, or other accounting record. They "account for" the transfer of funds between one entity and another. However, economic costs are a much broader cost category. While they include accounting costs, other typical economic costs a regulator may encounter when assessing pollution control issues would include external costs - the cost incurred by others and not part of the accounting system of the firm. For example, a boiler may produce large particles of unburned or partially burned fuels (soot). While the owner of the boiler pays for the cost of that fuel through higher fuel costs, it does not include the cost of cleaning that soot off of buildings and houses upwind of the plant. The owner also does not have to pay for the asthma medicine for affected people who suffer respiratory problems because of that soot, nor does it compensate them for the discomfort of that asthma attack. The first of these economic costs is fairly straight forward and the economic literature has many examples of how to approximate it. The second is a health issue that can also be approximated, although only after a great deal of study and analysis. The third cost, compensation for discomfort, is a psychic cost and is extremely difficult to quantify. However these and many other similar costs should be considered by the regulatory analyst when assessing the usefulness of a pollution control alternative.

Similar to economic and accounting costs, accounting benefits (revenues, avoided production costs) are a subset of economic benefits. Pollution control devices reduce pollution and their installation reduces the occurrence of these economic costs, so the regulatory analyst would include among the benefits of the device the avoided economic costs derived from the pollutant. In other words, a soot free building does not have to pay for cleaning - and that avoided cost is considered a benefit of the device. Similarly, not having an asthma attack is also considered a benefit of the device. When we perform an economic assessment of a pollution control device, we such as a bag house for capturing soot before it enters the atmosphere, we look at the benefit of avoiding these economic costs.

When performing an economic assessment of a pollution control alternative, the analyst can apply economic costs and benefits to payback (to establish a "social payback" period), net present value analysis (for benefit cost analyses or to compare to the social discount rate through ROI or IRR). Without going into detail on the science of economic assessment, the analyst should be able to go back to each of the discussions above and readily see how to apply these simple accounting tools to an economic study.

2.5 Estimating Procedure

The estimating procedure used in the Manual consists of five steps: (1) obtaining the facility parameters and regulatory options for a given facility; (2) roughing out the control system design; (3) sizing the control system components; (4) estimating the costs of these individual components; and (5) estimating the costs (capital and annual) of the entire system.

2.5.1 Facility Parameters and Regulatory Options

Obtaining the facility parameters and regulatory options involves not only assembling the parameters of the air pollution source (i.e., the quantity, temperature, and composition of the emission stream(s)), but also compiling data for the facility's operation. (Table 2.2 lists examples of these.) We identify two facility parameters: intensive (with values independent of quantity or dimensions) and extensive (size-dependent variables, such as the gas volumetric flow rate).

Regulatory options are usually specified by others (generally a regulatory authority) and are often technology driven, typically defining allowable ways to achieve a predetermined emission limit. These options range from "no control" to a requirement for the system to reach the maximum control technically achievable. The options allowed will depend, firstly, on whether the emission source is a point source (a stack or other identifiable primary source of pollution), a fugitive source (a process leak or other source of pollution that could not reasonably pass through a stack, chimney, vent, or other functionally-equivalent opening) or an area fugitive source (an unenclosed or partly enclosed area, such as a storage pile or a construction site). Stacks are normally controlled by "add-on" devices - the primary focus of this Manual. (However, some of these devices can be used to control process fugitive emissions in certain cases, such as a fabric filter used in conjunction with a building evacuation system.) Add-on pollution controls are normally used to meet a specified emission limit, although in the case of particulate emissions, they may also be required to meet an opacity level.

Table 2.2: Facility Parameters and Regulatory Options

Facility Parameters	Regulatory Options
Intensive	**No control**
Facility status (new or existing, location)	
Gas Characteristics (temperature, pressure, moisture control)	**ìAdd-onî devices**
Pollutant concentration(s) and/or particle size distribution	Emission limits
	Opacity limits
Extensive	**Process modification**
Facility capacity	Raw material changes
Facility life	Fuel substitution
Exhaust gas flow rate	
Pollutant emission rate(s)	**Source/Feedstock pretreatment**
	Coal desulfurization
	Wet dust suppression

2.5.2 Control System Design

Roughing out the control system design involves deciding what kinds of systems will be priced (a decision that will depend on the pollutants to be controlled, exhaust gas stream conditions, and other factors), and what auxiliary equipment will be needed. When specifying the auxiliary equipment, several questions need to be answered:

- What type of hood (if any) will be needed to capture the emissions at the source?
- Will a fan be needed to convey the exhaust through the system? Does the system require any other auxiliary equipment, such as a pump to inject liquids into the exhaust gas stream?
- Does the exhaust stream pose any hazard to the materials of the hoods, ducts, fans, and other auxiliary equipment? Is the exhaust caustic or acidic? Is it abrasive? Does the treatment of the exhaust render it caustic or acidic?
- Does the exhaust stream require any pre-treatment (e.g., cyclone or another pre-cleaner) before it enters the control device?
- Will the captured pollutants be disposed of or recycled? How will this be done?
- Can the on-site capacity (e.g., utilities, stockpiling space) accommodate the added requirements of the control system?

The kinds of auxiliary equipment selected will depend on the answers to these and other site-specific questions. However, regardless of the source being controlled, each system will likely contain, along with the control device itself, the following auxiliaries:

- Hood, or other means for capturing the exhaust;
- Ductwork, to convey the exhaust from the source to, through, and from the control system;
- Fan system (fan, motor, starter, inlet/outlet dampers, etc.), to move the exhaust through the system and to prevent pressure drop within the system due to the pollution control system;
- Stack, for dispersing the cleaned gas into the atmosphere.

2.5.3 Sizing the Control System

Once the system components have been selected, they must be sized. Sizing is probably the most critical step because the assumptions made in this step will more heavily influence capital investment than any other. Table 2.3 lists examples of these parameters. Also listed in Table 2.3 are general parameters which must be specified before the purchased cost of the system equipment can be estimated. Note that, unlike the control device parameters, these parameters may apply to any kind of control system. They include materials of construction (which may range from carbon steel to various stainless steels to fiberglass- reinforced plastic), presence or absence of insulation, and the economic or useful life of the system. As indicated in Section 2.4.2, this last parameter is required for estimating the annual capital recovery costs. The lifetime not only varies according to the type of the control system, but with the severity of the environment in which it is installed. Each of the control-specific chapters of this Manual and the Air Compliance Advisor (ACA) program include a comprehensive list of the specific parameters that must be considered for each device.

Table 2.3: Examples of Typical Control Device Parameters [11]

General	Device-Specific
Material of construction: carbon steel	Gas-to-cloth ratio (ìcritical parameterî): 3.0 to 1
Insulated? Yes	Pressure drop: 6.0 in w.c. (inches water column)
Economic life: 20 years	Construction: standard (vs. custom)
Redundancy[a]: none	Duty: continuous (vs. intermittent)
	Filter type: shaker
	Bag material: polyester, 16-oz.

[a] Refers to whether there are any extra equipment items installed (e.g., fans) to function in case the basic items become inoperative, so as to avoid shutting down the entire system.

2.5.4 Estimating Total Capital Investment

2.5.4.1 General Considerations

The fourth step is estimating the purchased equipment cost of the control system equipment. As discussed in Section 2.2, total direct cost includes purchased equipment cost, which in turn, is the sum of the base equipment cost (control device plus auxiliaries), freight, instrumentation, and sales tax. The values of these installation factors depend on the type of the control system installed and are, therefore, listed in the individual Manual chapters dedicated to them. These costs are available from this Manual for the most commonly used add-on control devices and auxiliary equipment, with each type of equipment covered in a separate chapter (see Table of Contents and the discussion in Chapter 1). Total Direct Cost also includes Direct Installation Cost, which contains many of the cost categories included in Section 2 of this Manual, Generic Equipment and Devices.

Most of the costs in each of the subsequent sections of this Manual were derived from data obtained from control equipment vendors. For many control devices there are many vendors, which allowed us to offer highly representative costs, based upon the average cost of components submitted by those vendors in response to Agency survey efforts. [7] For items that are mass produced or "off-the-shelf" equipment, vendors provided a written quotation listing their costs, model designations, date of quotation, estimated shipment date, and other information. For other equipment there are not many vendors or we did not receive many responses to our inquiries. In these cases, we offer costs that are as representative as possible and the cost discussion in that control's particular chapter offers an appropriate caveat to the analyst.

For some controls, no amount of vendor data would have made our cost numbers more accurate because the control in question is either so large or so site-specific in design that suppliers design, fabricate, and construct each control according to the specific needs of the facility. For these devices (specifically, SCR reactors and FGD units), the Manual deviates from its standard approach of providing study level costs and, instead, provides a detailed description of the factors that influence the TCI for the analyst to consider when dealing with a vendor quotation. For these kinds of controls, the vendor may still give quotations, but will likely take much longer to do so and may even charge for this service, to recoup the labor and overhead expenses of his estimating department. When performing a cost analysis, the cost of the quotation is a part of the TCI.

Generally, vendor quotes are "F.O.B." (free-on-board) the vendor, meaning that no taxes, freight, or other charges are included. For these equipment, the analyst must take care to identify and include the cost of transportation, taxes, and other necessary charges in the TCI (see Figure 2.1). The costs of freight, instrumentation, and sales tax are calculated differently from the direct and indirect installation costs. These items are developed by multiplying the base equipment cost (F.O.B. the vendor) by an industry-accepted factor. Unlike other estimating factors that differ from system to system, installation factors are essentially equal for all control systems. Table 2.4, below, displays values for these factors.

Table 2.4: Cost Ranges for Freight, Sales Tax, and Instrumentation

	% of Total Equipment Cost, FOB	
Cost	Range	Typical
Freight	0.01 ñ 0.10	0.05
Sales Tax	0 ñ 0.08	0.03
Instrumentation	0.05 ñ 0.30	0.10

To some extent, the application of an appropriate factor requires the subjective application of the analyst's best judgement. For example, the range in freight costs is, in part, a function of the distance between the vendor and the site. The lower end of the factor range represents shorter distance deliveries, while the upper end of the range would reflect freight charges to remote locations such as Alaska and Hawaii.[6] The sales tax factors simply reflect the range of local and state tax rates currently in effect in the United States.[8] In some locations, and for many institutional and governmental purchases, sales taxes do not apply; (hence the zero value at the low end of the sales tax factor range). The range of instrumentation factors is also quite large. For systems requiring only simple continuous or manual control, the lower factor would apply. However, if the control is intermittent and/or requires safety backup instrumentation, the higher end of the range would be applicable.[6] Finally, some "package" control systems (e.g., incinerators covered in Chapter 3) have built-in controls, with instrumentation costs included in the base equipment cost. In those cases, the instrumentation factor to use would, of course, be zero.

2.5.4.2 Retrofit Cost Considerations

Probably the most subjective part of a cost estimate occurs when the control system is to be installed on an existing facility. Unless the original designers had the foresight to include additional floor space and room between components for new equipment, the installation of retrofitted pollution control devices can impose an additional expense to "shoe-horn" the equipment into the right locations. For example, an SCR reactor can occupy tens of thousands of square feet and must be installed directly behind a boiler's combustion chamber to offer the best environment for NOx removal. Many of the utility boilers currently considering an SCR reactor to meet the new federal NOx limits are over thirty years old - designed and constructed before SCR was a proven technology in the United States. For these boilers, there is generally little room for the reactor to fit in the existing space and additional ductwork, fans, and flue gas heaters may be needed to make the system work properly.

To quantify the unanticipated additional costs of installation not directly related to the capital cost of the controls themselves, engineers and cost analysts typically multiply the cost of the system by a retrofit factor. The proper application of a retrofit factor is as much an art as it is a science, in that it requires a good deal of insight, experience, and intuition on the part of the analyst. The key behind a good cost estimate using a retrofit factor is to make the factor no larger than is necessary to cover the occurrence of unexpected (but reasonable) costs for demolition and installation. Such unexpected costs include - but are certainly not limited to - the unexpected magnitude of anticipated cost elements; the costs of unexpected delays; the cost of re-engineering and re-fabrication; and the cost of correcting design errors.

The magnitude of the retrofit factor varies across the kinds of estimates made as well as across the spectrum of control devices. At the study level, analysts do not have sufficient information to fully assess the potential hidden costs of an installation. At this level, a retrofit factor of as much as 50 percent can be justified. Even at detailed cost level (± 5 percent accuracy), vendors will not be able to fully assess the uncertainty associated with a retrofit situation and will include a retrofit

factor in their assessments. For systems installed at the end of the stack, such as flares, retrofit uncertainty is seldom a factor. In these cases, an appropriate retrofit factor may be one or two percent of the TCI. In complicated systems requiring many pieces of auxiliary equipment, it is not uncommon to see retrofit factors of much greater magnitude can be used.

Since each retrofit installation is unique, no general factors can be developed. A general rule of thumb as a starting point for developing an appropriate retrofit factor is: The larger the system, the more complex (more auxiliary equipment needed), and the lower the cost level (eg. study level, rather than detailed), the greater the magnitude of the retrofit factor. Nonetheless, some general information can be given concerning the kinds of system modifications one might expect in a retrofit:

1. Auxiliary equipment. The most common source of retrofit-related costs among auxiliary equipment types comes from the ductwork related costs. In addition, to requiring very long duct runs, some retrofits require extra tees, elbows, dampers, and other fittings. Furthermore, longer ducts and additional bends in the duct cause greater pressure drop, which necessitates the upgrading or addition of fans and blowers.

2. Handling and erection. Because of a "tight fit," special care may need to be taken when unloading, transporting, and placing the equipment. This cost could increase significantly if special means (e.g., helicopters) are needed to get the equipment on roofs or to other inaccessible places.

3. Piping, Insulation, and Painting. Like ductwork, large amounts of piping may be needed to tie in the control device to sources of process and cooling water, steam, etc. Of course, the more piping and ductwork required, the more insulation and painting will be needed.

4. Site Preparation. Site preparation includes the surveying, clearing, leveling, grading, and other civil engineering tasks involved in preparing the site for construction. Unlike the other categories, this cost may be very low or zero, since most of this work would have been done when the original facility was built. However, if the site is crowded and the control device is large, the size of the site may need to be increased and then site preparation may prove to be a major source of retrofit-related costs.

5. Off-Site Facilities. Off-site facilities should not be a major source of retrofit costs, since they are typically used for well-planned activities, such as the delivery of utilities, transportation, or storage.

6. Engineering. Designing a control system to fit into an existing plant normally requires extra engineering, especially when the system is exceptionally large, heavy, or utility-consumptive. For the same reasons, extra supervision may be needed when the installation work is being done.

7. Lost Production. The shut-down for installation of a control device into the system should be a well-planned event. As such, its cost should be considered a part of the indirect installation cost (start-up). However, unanticipated problems with the installation due to retrofit-related conditions can impose significant costs on the system. (For example, consider a pollution control device to be installed in the middle of a stack. After shutting down the plant, removing a section of the stack reveals it has been worn too thin to weld the device to it, necessitating the fabrication and replacement of a major portion of the stack.) The net revenue (i.e., gross revenue minus the direct costs of generating it) lost during this unanticipated shutdown period is a bonafide retrofit expense.

Due to the uncertain nature of many estimates, analysts may want to add an additional contingency (i.e., uncertainty) factor to their estimate. However, the retrofit factor is a kind of contingency factor and the cost analyst must be careful to not impose a double penalty on the system for the same unforseen conditions. Retrofit factors should be reserved for those items directly related to the demolition, fabrication, and installation of the control system. A contingency factor should be reserved (and applied to) only those items that could incur a reasonable but unanticipated increase but are not directly related to the demolition, fabrication, and installation of the system. For example, a hundred year flood may postpone delivery of materials, but their arrival at the job site is not a problem unique to a retrofit situation.

2.5.5 Estimating Annual Costs

Determining the total annual cost is the last step in the estimating procedure. As mentioned in Section 2.3 the TAC is comprised of three components—direct and indirect annual costs and recovery credits. Unlike the installation costs, which are factored from the purchased equipment cost, annual cost items are usually computed from known data on the system size and operating mode, as well as from the facility and control device parameters.

Following is a more detailed discussion of the items comprising the total annual cost. (Values/factors for these costs are also given in the chapters for the individual devices.)

2.5.5.1 Raw Materials

Raw materials are generally not required with control systems. Exceptions would be chemicals used in gas absorbers or venturi scrubbers as absorbents or to neutralize acidic exhaust gases (e.g., hydrochloric acid). Chemicals may also be required to treat wastewater discharged by scrubbers or absorbers before releasing it to surface waters. If the source uses the same raw

materials for production, the analyst must be careful to include only are only those costs that are attributable to the raw materials needed by the control device. Quantities of chemicals required are calculated via material balances, with an extra 10 to 20% added for miscellaneous losses. Costs for chemicals are available from the Chemical Marketing Reporter and similar publications.

2.5.5.2 Labor

The amount of labor required to operate and maintain a pollution control system depends on its size, complexity, level of automation, and operating mode (i.e., batch or continuous). The labor is usually figured on an hours-per-shift basis. As a rule, though, data showing explicit correlations between the labor requirement and capacity are hard to obtain. One non-linear correlation found in the literature is shown below:[11]

$$\frac{L_2}{L_1} = \left(\frac{V_2}{V_1}\right)^y \qquad (2.9)$$

where

L_1, L_2 = labor requirements for systems 1 and 2
V_1, V_2 = capacities of systems 1 and 2 (as measured by the gas flow rate, for instance)
y = 0.2 to 0.25 (typically)

The exponent in Equation 2.9 can vary considerably. Conversely, in many cases, the amount of operator labor required for a system will be approximately the same regardless of its size.

Maintenance labor is calculated in the same way as operating labor and is influenced by the same variables. The maintenance labor rate, however, is normally higher than the operating labor rate, mainly because more skilled personnel are required. Many cost studies use a flat ten percent premium over the operations labor wage rate for maintenance labor costs.[12] A certain amount must also be added to operating labor to cover supervisory requirements. Generally, cost estimates include supervisory labor as a flat fifteen per cent of the operating labor requirement.[12] To obtain the annual labor cost, multiply the operating and supervisory labor requirements(labor-hr/operating-hr) by the respective wage rates (in $/labor-hr) and the system operating factor (number of hours per year the system is in operation). Wage rates also vary widely, depending upon the source category, geographical location, etc. These data are tabulated and periodically updated by the U.S. Department of Labor, Bureau of Labor Statistics, in its Monthly Labor Review and in other publications. This Manual uses labor rates that are representative of the industries at the national level. The supplemental CO$T-AIR spreadsheets and the Air Compliance Advisor (ACA) incorporate these rates as defaults. For regulatory cost assessments, these wages (adjusted for inflation through an appropriate cost index) should be adequate for study level purposes. For industry users of this manual, the CO$T-AIR spreadsheets and the ACA can be customized to include site-specific labor rates and improve the accuracy of the analysis.

Finally, note that the wage rates used by the Manual and its supplemental programs are base labor rates, which do not include payroll and plant overhead. Wages found in reports from the Bureau of Labor Statistics or some other reliable source may or may not include overhead. The analyst must be careful to apply overhead and other wage adjustment factors uniformly. (See the discussion on Overhead, below.)

2.5.5.3 Maintenance Materials

Maintenance also requires maintenance materials—oil, other lubricants, duct tape, etc., and a host of small tools. The costs for these items can be figured individually, but since they are normally so small, they are usually factored from the maintenance labor. Reference [11] suggests a factor of 100% of the maintenance labor to cover the maintenance materials cost.

2.5.5.4 Utilities

This cost category covers many different items, ranging from electricity to compressed air. Of these, only electricity is common to all control devices, where fuel oil and natural gas are generally used only by incinerators; water and water treatment, by venturi scrubbers, quenchers, and spray chambers; steam, by carbon adsorbers; and compressed air, by pulse-jet fabric filters. Techniques and factors for estimating utility costs for specific devices are presented in their respective sections. However, because nearly every system requires a fan to convey the exhaust gases to and through it, a general expression for computing the fan electricity cost (C_e) is given here:[6]

$$C_e = \frac{0.746\ Q\ \Delta P\ s\ \theta\ p_e}{6356\ \eta} \qquad (2.10)$$

where

Q = gas flow rate (actual ft^3/min, acfm)
ΔP = pressure drop through system (inches of water, column) (Values for ΔP are given in the chapters covering the equipment items.)
s = specific gravity of gas relative to air (1.000, for all practical purposes)
θ = operating factor (hr/yr)
η = combined fan and motor efficiency (usually 0.60 to 0.70)
p_e = electricity cost ($/kwhr)

A similar expression can be developed for calculating pump motor electricity requirements.

2.5.5.5 Waste Treatment and Disposal

Though often overlooked, there can be a significant cost associated with treating and/or disposing of waste material captured by a control system that neither can be sold nor recycled to the process. Liquid waste streams, such as the effluent from a gas absorber, are usually processed

before being released to surface waters. The type and extent of this processing will, of course, depend on the characteristics of the effluent. For example, the waste can first be sent to one (or more) clarifiers, for coagulation and removal of suspended solids. The precipitate from the clarifier is then conveyed to a rotary filter, where most of the liquid is removed. The resulting filter cake is then disposed of, via landfilling, for example.

The annual cost of this treatment can be relatively high—$1.00 to $2.00/thousand gallons of treated material or more.[13] The (non-hazardous) solid waste disposal costs (via landfilling, for example) typically would add another $20 to $30/ton of disposed material.[14] This, however, would not include transportation to the disposal site. Disposal of hazardous waste (which may not be landfilled) can be much more costly—$200 to $300/ton or more. More information on these technologies and their costs is found in References [13] and [14].

2.5.5.6 Replacement Materials

The cost or maintenance materials is a component of the operations and maintenance function of the system and is not the same thing as the system's replacement materials cost, which is the cost of such items as carbon (for carbon absorbers), bags (for fabric filters) and catalyst (for catalytic incinerators), along with the labor for their installation. Because replacement materials last for more than a year but are consumed by the system, they cannot be included in the general maintenance and operations costs, which are annual in nature. Instead, these costs must be annualized by first determining the life of the material, then applying the appropriate capital recovery factor to that cost to determine its annualized value (see section 2.4.5.3, above). The annual cost of the replacement materials is a function of the initial parts cost, the parts replacement labor cost, the life of the parts, and the interest rate, as follows:

$$CRC_p = \left(C_p + C_{pl}\right) CRF_p \qquad (2.11)$$

where

CRC_p = capital recovery cost of replacement parts ($/yr)
C_p = initial cost of replacement parts, including sales taxes and freight ($)
C_{pl} = cost of parts-replacement labor ($)
CRF_p = capital recovery factor for replacement parts (defined in Section 2.3).

The useful life of replacement materials is generally less than the useful life of the rest of the control system - typically two to five years. Consequently, the annualization of the system's replacement materials must be done separately from the annualization of the control system itself. Furthermore, the annualized cost of the pollution control system should be performed net of the cost of the replacement materials needed at the beginning of operations to prevent double counting. Replacement materials labor will vary, depending upon the amount of the material, its workability, accessibility of the control device, and other factors. The cost of replacement materials labor should be included in the cost of the materials before annualization.

2.5.5.7 Overhead

This cost is easy to calculate, but often difficult to comprehend. Much of the confusion surrounding overhead is due to the many different ways it is computed and to the several costs it includes, some of which may appear to be duplicative.

There are, generally, two categories of overhead, payroll and plant. Payroll overhead includes expenses directly associated with operating, supervisory, and maintenance labor, such as: workmen's compensation, Social Security and pension fund contributions, vacations, group insurance, and other fringe benefits. Some of these are fixed costs (i.e., they must be paid regardless of how many hours per year an employee works). Payroll overhead is traditionally computed as a percentage of the total annual labor cost (operating, supervisory, and maintenance).

Conversely, plant (or "factory") overhead accounts for expenses not necessarily tied to the operation and maintenance of the control system, including: plant protection, control laboratories, employee amenities, plant lighting, parking areas, and landscaping. Some estimators compute plant overhead by taking a percentage of all labor plus maintenance materials [11], while others factor it from the total labor costs alone.[2]

For study estimates, it is sufficiently accurate to combine payroll and plant overhead into a single indirect cost. This is done in this Manual. Also, overhead is factored from the sum of all labor (operating, supervisory, and maintenance) plus maintenance materials, the approach recommended in reference [11]. The factors recommended therein range from 50 to 70% [11] An average value of 60% is used in this Manual. For more accurate assessments by industrial users of the Manual, the CO$T-AIR spreadsheets and the ACA allow for customization of these factors.

2.5.5.8 Property Taxes, Insurance, and Administrative Charges

These three indirect operating costs are factored from the system total capital investment, at 1, 1, and 2%, respectively. Property taxes and insurance are self-explanatory. Administrative charges covers sales, research and development, accounting, and other home office expenses. (It should not be confused with plant overhead, however.) For simplicity, the three items are usually combined into a single, 4% factor. This is standard approach used in in all OAQPS cost analyses and by this Manual.

2.6 Example

All-American Electrical (AAE) [1] operates a single 600 MWe tangentially fired high sulfur bituminous coal-fired boiler to produce steam to power it generators. It emits an uncontrolled 90,000 tons of sulfur per year, and because it is planning on a major renovation, it must install devices to reduce its sulfur emissions to less than 900 tons per year (99 percent removal efficiency). After careful study of the available technologies, AAE has determined that either a wet limestone

flue gas desulfurization (FGD) unit or a wet buffered lime FGD would be the most logical choice to achieve such a high removal rate. For simplification purposes we will assume either device would have an operating life of five years, after which the scrubbers could be sold as scrap for a salvage value of about $500,000. Table 2.5, below, displays the capital and annual costs associated with each of the alternative devices.

Table 2.5: Capital, O&M, and Parasitic Energy Costs of Alternative FGD Controls

	Wet Limestone FGD	Wet Buffered Lime FGD
Capital Cost	$200,000,000	$180,000,000
Annual O&M Costs		
Fixed O&M Costs [a]	$2,000,000	$2,100,000
Reagent	$1,200,000	$3.750,000
Auxiliary Power	$1,300,000	$1,150,000
Annual Gypsum Sales	$1,200,000	$600,000
Parasitic Power [b]	$950,000	$375,000
Salvage Value (after 10 years)	$500,000	$500,000

[a] Estimated at 1% of capital cost

[b] In many systems, the insertion of a pollution control device causes the system to lose productive capacity. This can be caused by the device creating obstructions in the flue, temperature losses that create imbalances, or other physical changes that affect performance. these losses are collectively termed "parasitic power" losses.

From the information in Table 2.5, neither device can be shown to be superior to the other. It costs $20 million less to install a wet buffered lime scrubber, but a buffered lime FGD would cost over three times as much each year for the purchase of the lime, relative to the cost of the reagent in a limestone FGD. Each FGD has similar fixed O&M costs, but because a buffered lime FGD uses much less reagent, it requires less power to run - about half the power demand and about 40 percent of the productive loss of the limestone FGD. While these factors indicate
the wet buffered lime FGD may be a better alternative, the use of less reagent also means the production of less gypsum by-product - for about half the expected revenue generating capability of a limestone system. To make our selection, we must rely upon our financial tools.

The exercise does not lend itself to a payback analysis, even though there are revenues to be generated from the sale of the scrubber's byproduct. So long as annual costs exceed annual revenues, payback will not an alternative because there will be no net revenue to help offset the capital costs of the project. Furthermore, even if one were to ignore the cost component of the cash flow, the revenues from most pollution control devices are so low that their payback values are meaningless. For instance, the limestone and buffered lime scrubbers in this exercise have a simple payback (without considering costs) of 167 and 300 years,

respectively. Consequently, the analyst must look to the more sophisticated tools available: cash flow analysis and net present value.

Table 2.6 shows the hypothetical cash flows from each alternative control in nominal dollars. You will notice that the cost for O&M and the revenues from selling the gypsum by-product are constant over time. That is because we have ignored any inflation rate change in prices and have created our cash flow analysis in real terms. This is the preferred way to approach this kind of analysis, since it relies on the most accurate information available (current prices) and does not try to extrapolate those prices into the future. Because we will perform our cash flow analysis in real dollars, we must use the real interest rate to determine net present values. We will assume AAE can borrow funds at will at a nominal interest rate of nine percent and that informed sources expect the inflation rate over the relevant range to be, on average, four percent. Consequently, the real rate of interest is (nine percent minus four percent) five percent. Using real dollars for revenues and costs and then using nominal interest rates for our discounting factors (nine percent) would have led to an understatement of the net present value of the projects, making them appear less beneficial to AAE.

Translating the costs in each future year to year zero values means applying the factors found in Table A.1 from Appendix A. From the 10 percent column, we applied the factors 0.90909, 0.82645, 0.75131, 0.68301, and 062092, respectively, to the net costs of years 1, 2, 3, 4, and 5 to determine the year zero costs, and then sum all of the values to derive the net present value for each control alternative. Based upon the information developed in the cash flow analysis and the NPV calculation, which control device is the best one for AAE to install? The answer is still not evident! Even with a twenty million dollar capital cost savings, the net present value of the wet buffered lime FGD is only about a half million dollars more expensive than the wet limestone FDG! This is a function of the other cash flow components - the higher operating cost of the buffered lime system versus the higher revenue generating capacity of the limestone FGD, both of which work to almost completely eliminate the capital cost advantage of the buffered lime scrubber. Clearly, relying on just the sticker price of the two units could have driven us to a potentially bad decision. So now what? Payback analysis does not offer any help, (nor will IRR, which also relies upon a positive net cash flow to work). Cash flow analysis tells us that, within our study-level estimation range, the two devices are almost identical. That in and of itself is important information, because the environmental engineer can be fairly certain that whichever device they choose, the affect of that choice on his company will be about the same. That leaves them free to look at other considerations: Twice as much limestone means twice as much storage and twice as much stockpiling of the gypsum by-product. Is that an important factor? Limestone is more caustic than buffered lime, but it takes less equipment to operate the system. Should the engineer opt for simplicity in design or potentially higher rates of repair? These are the sort of considerations that can now come into play in making a decision, now that the relative values of each device has been determined.

This does not mean that our process has failed. Far from it. If our input assumptions have been made correctly, then we have determined that from a cost standpoint, there does not seem to be an appreciably different risk to choosing one device over the other. However, other considerations may play a roll in making the choice clearer. For instance, the limestone scrubber will produce about twice as much gypsum as the wet buffered lime scrubber. Does the storage, transportation, or marketability of that amount of gypsum create a problem? Likewise, it takes about three times as much limestone to remove the same amount of sulfur, relative to the amount of lime needed, but the lime costs between five and seven times as much as the limestone. Do these considerations clarify the choice? Finally, the power demands for each device differ significantly, both in terms of operation and in lost productive capacity. Perhaps these considerations will make one device more attractive to the firm. The bottom line is that there is no clear-cut "cookbook" process through which the analyst will be able to make the right informed decision each time, and the formalized costing methodology employed by the Manual is only a part of that process. However, if the Manual's methodology is followed rigorously and in an unbiased manner, then the analyst can feel safe about the ROM-level cost of his alternative projects and can then move on to a more formal cost determination with the help of an engineering or consulting firm.

2.6: Cash Flow Analyses Exercise (in thousands of dollars)

	0	1	2	3	4	5	6	7	8	9	10
Limestone Scrubber											
Income											
Gypsum Sales	0	1,200	1,200	1,200	1,200	1,200	1,200	1,200	1,200	1,200	1,200
Salvage Value	0	0	0	0	0	0	0	0	0	0	500
Expenses											
Capital Investment	200,000	0	0	0	0	0	0	0	0	0	0
Annual O&M Costs	0	4,500	4,500	4,500	4,500	4,500	4,500	4,500	4,500	4,500	4,500
Parasitic Power	0	950	950	950	950	950	950	950	950	950	950
Net Annual Cost	-200,000	-4,250	-4,250	-4,250	-4,250	-4,250	-4,250	-4,250	-4,250	-4,250	-3,750
Present Value	-200,000	-4,048	-3,855	-3,671	-3,496	-3,330	-3,171	-3,020	-2,877	-2,740	-2,302
NPV	-232,510										
Buffered Lime Scrubber											
Income											
Gypsum Sales	0	600	600	600	600	600	600	600	600	600	600
Salvage Value	0	0	0	0	0	0	0	0	0	0	500
Expenses											
Capital Investment	180,000	0	0	0	0	0	0	0	0	0	0
Annual O&M Costs	0	7,000	7,000	7,000	7,000	7,000	7,000	7,000	7,000	7,000	7,000
Parasitic Power	0	375	375	375	375	375	375	375	375	375	375
Net Annual Cost	-180,000	-6,775	-6,775	-6,775	-6,775	-6,775	-6,775	-6,775	-6,775	-6,775	-6,275
Present Value	-180,000	-6,452	-6,145	-5,852	-5,574	-5,308	-5,056	-4,815	-4,586	-4,367	-3,852
NPV	-232,008										

References

[1] Perry, Robert H., and Chilton, Cecil H., Perry's Chemical Engineers' Handbook (Fifth Edition), McGraw-Hill, New York, 1973, pp. 25-12 to 25-16.

[2] Humphries, K. K. and Katell, S., Basic Cost Engineering, Marcel Dekker, New York, 1981, pp. 17-33.

[3] Grant, E.L., Ireson, W.G., and Leavenworth, R.S., Principles of Engineering Economy, Sixth Edition, John Wiley & Sons, New York, 1976.

[4] CEIS (OAQPS) Cost Guidance Memo #2: "Implementing OMB Circular A-94: Guidelines and Discount Rates for Benefit Cost Analysis of Federal Programs," March 18, 1993.

[5] Scheraga, Joel D., Draft of "Supplemental Guidelines on Discounting in the Preparation of Regulatory Impact Analyses", Office of Policy, Planning and Evaluation, U.S. EPA, March 31, 1989.

[6] Vatavuk, W.M. and Neveril, R.B., "Estimating Costs of Air-Pollution Control Systems—Part I: Parameters for Sizing Systems," Chemical Engineering, October 6, 1980, pp. 165-168.

[7] Pollution Equipment News 1996 Buyer's Guide, Rimbach Publishing, Pittsburgh, 1996.

[8] Internal Revenue Service, Form 1040, 1985.

[9] Shattuck, D.M., et al., Retrofit FGD Cost-Estimating Guidelines. Electric Power Research Institute, Palo Alto, CA (CS-3696, Research Project 1610-1), October 1984.

[10] Kaplan, N., et al., "Retrofit Costs of SO_2 and NO_x Control at 200 U.S. Coal-Fired Power Plants," Pittsburgh Coal Conference, 1990.

[11] Peters, M.S. and Timmerhaus, K.D., Plant Design and Economics for Chemical Engineers (Third Edition), McGraw-Hill, New York, 1980.

[12] Vatavuk, W.M. and Neveril, R.B., "Estimating Costs of Air Pollution Control Systems—Part II: Factors for Estimating Capital and Operating Costs," Chemical Engineering, November 3, 1980, pp. 157-162.

[13] Vatavuk, W.M. and Neveril, R.B., "Estimating Costs of Air-Pollution Control Systems—Part XVII: Particle Emissions Control," Chemical Engineering, April 2, 1984, pp. 97-99.

[14] The RCRA Risk-Cost Analysis Model, U.S. Environmental Protection Agency, Office of Solid Waste, January 13, 1984.

APPENDIX A

Net Present Value and Capital Recovery Factor Tables

Table A.1 displays the amount an individual would be willing to accept today for a dollar promised in the future. Select the year in which the dollar is supposed to be paid from the leftmost column and the discount rate from the top row. The value where the column and row intersect is the present value of that future dollar. For instance, if you were promised a dollar twelve years from now, and you believed the interest rate over that period would be 9.5 percent, then you would be willing to accept 33.7 cents for that dollar today.

Table A.1: Present Value Factors for a Dollar to Be Paid Now Instead of in a Future Year

	5.50%	6.00%	6.50%	7.00%	7.50%	8.00%	8.50%	9.00%	9.50%	10.00%
1	0.94787	0.9434	0.93897	0.93458	0.93023	0.92593	0.92166	0.91743	0.91324	0.90909
2	0.89845	0.89	0.88166	0.87344	0.86533	0.85734	0.84946	0.84168	0.83401	0.82645
3	0.85161	0.83962	0.82785	0.8163	0.80496	0.79383	0.78291	0.77218	0.76165	0.75131
4	0.80722	0.79209	0.77732	0.7629	0.7488	0.73503	0.72157	0.70843	0.69557	0.68301
5	0.76513	0.74726	0.72988	0.71299	0.69656	0.68058	0.66505	0.64993	0.63523	0.62092
6	0.72525	0.70496	0.68533	0.66634	0.64796	0.63017	0.61295	0.59627	0.58012	0.56447
7	0.68744	0.66506	0.64351	0.62275	0.60275	0.58349	0.56493	0.54703	0.52979	0.51316
8	0.6516	0.62741	0.60423	0.58201	0.5607	0.54027	0.52067	0.50187	0.48382	0.46651
9	0.61763	0.5919	0.56735	0.54393	0.52158	0.50025	0.47988	0.46043	0.44185	0.4241
10	0.58543	0.55839	0.53273	0.50835	0.48519	0.46319	0.44229	0.42241	0.40351	0.38554
11	0.55491	0.52679	0.50021	0.47509	0.45134	0.42888	0.40764	0.38753	0.36851	0.35049
12	0.52598	0.49697	0.46968	0.44401	0.41985	0.39711	0.3757	0.35553	0.33654	0.31863
13	0.49856	0.46884	0.44102	0.41496	0.39056	0.3677	0.34627	0.32618	0.30734	0.28966
14	0.47257	0.4423	0.4141	0.38782	0.36331	0.34046	0.31914	0.29925	0.28067	0.26333
15	0.44793	0.41727	0.38883	0.36245	0.33797	0.31524	0.29414	0.27454	0.25632	0.23939
16	0.42458	0.39365	0.3651	0.33873	0.31439	0.29189	0.2711	0.25187	0.23409	0.21763
17	0.40245	0.37136	0.34281	0.31657	0.29245	0.27027	0.24986	0.23107	0.21378	0.19784
18	0.38147	0.35034	0.32189	0.29586	0.27205	0.25025	0.23028	0.21199	0.19523	0.17986
19	0.36158	0.33051	0.30224	0.27651	0.25307	0.23171	0.21224	0.19449	0.17829	0.16351
20	0.34273	0.3118	0.2838	0.25842	0.23541	0.21455	0.19562	0.17843	0.16282	0.14864
21	0.32486	0.29416	0.26648	0.24151	0.21899	0.19866	0.18029	0.1637	0.1487	0.13513
22	0.30793	0.27751	0.25021	0.22571	0.20371	0.18394	0.16617	0.15018	0.1358	0.12285
23	0.29187	0.2618	0.23494	0.21095	0.1895	0.17032	0.15315	0.13778	0.12402	0.11168
24	0.27666	0.24698	0.2206	0.19715	0.17628	0.1577	0.14115	0.1264	0.11326	0.10153
25	0.26223	0.233	0.20714	0.18425	0.16398	0.14602	0.13009	0.11597	0.10343	0.0923

Table A.1: Continued

	10.50%	11.00%	11.50%	12.00%	12.50%	13.00%	13.50%	14.00%	14.50%	15.00%
1	0.90498	0.9009	0.89686	0.89286	0.88889	0.88496	0.88106	0.87719	0.87336	0.86957
2	0.81898	0.81162	0.80436	0.79719	0.79012	0.78315	0.77626	0.76947	0.76276	0.75614
3	0.74116	0.73119	0.7214	0.71178	0.70233	0.69305	0.68393	0.67497	0.66617	0.65752
4	0.67073	0.65873	0.64699	0.63552	0.6243	0.61332	0.60258	0.59208	0.58181	0.57175
5	0.607	0.59345	0.58026	0.56743	0.55493	0.54276	0.53091	0.51937	0.50813	0.49718
6	0.54932	0.53464	0.52042	0.50663	0.49327	0.48032	0.46776	0.45559	0.44378	0.43233
7	0.49712	0.48166	0.46674	0.45235	0.43846	0.42506	0.41213	0.39964	0.38758	0.37594
8	0.44989	0.43393	0.4186	0.40388	0.38974	0.37616	0.36311	0.35056	0.3385	0.3269
9	0.40714	0.39092	0.37543	0.36061	0.34644	0.33288	0.31992	0.30751	0.29563	0.28426
10	0.36845	0.35218	0.33671	0.32197	0.30795	0.29459	0.28187	0.26974	0.25819	0.24718
11	0.33344	0.31728	0.30198	0.28748	0.27373	0.2607	0.24834	0.23662	0.2255	0.21494
12	0.30175	0.28584	0.27083	0.25668	0.24332	0.23071	0.2188	0.20756	0.19694	0.18691
13	0.27308	0.25751	0.2429	0.22917	0.21628	0.20416	0.19278	0.18207	0.172	0.16253
14	0.24713	0.23199	0.21785	0.20462	0.19225	0.18068	0.16985	0.15971	0.15022	0.14133
15	0.22365	0.209	0.19538	0.1827	0.17089	0.15989	0.14964	0.1401	0.1312	0.12289
16	0.2024	0.18829	0.17523	0.16312	0.1519	0.1415	0.13185	0.12289	0.11458	0.10686
17	0.18316	0.16963	0.15715	0.14564	0.13502	0.12522	0.11616	0.1078	0.10007	0.09293
18	0.16576	0.15282	0.14095	0.13004	0.12002	0.11081	0.10235	0.09456	0.0874	0.08081
19	0.15001	0.13768	0.12641	0.11611	0.10668	0.09806	0.09017	0.08295	0.07633	0.07027
20	0.13575	0.12403	0.11337	0.10367	0.09483	0.08678	0.07945	0.07276	0.06666	0.0611
21	0.12285	0.11174	0.10168	0.09256	0.08429	0.0768	0.07	0.06383	0.05822	0.05313
22	0.11118	0.10067	0.09119	0.08264	0.07493	0.06796	0.06167	0.05599	0.05085	0.0462
23	0.10062	0.09069	0.08179	0.07379	0.0666	0.06014	0.05434	0.04911	0.04441	0.04017
24	0.09106	0.0817	0.07335	0.06588	0.0592	0.05323	0.04787	0.04308	0.03879	0.03493
25	0.0824	0.07361	0.06579	0.05882	0.05262	0.0471	0.04218	0.03779	0.03387	0.03038

Table A.2 displays the annual payment you would have to make for a specific number of years to equal the present value of a single dollar borrowed today. Select the number of years you will make payments from the leftmost column and the discount rate from the top row. The value where the column and row intersect is annual payment on that borrowed dollar. For example, if you plan on making equal payments for twelve years at 9.5 percent interest to repay a dollar borrowed toaday, you would make annual payments of 14.3 cents.

Table A.2: Capital Recovery Factors for Equal Payments on a Dollar over a Number of Years

	5.50%	6.00%	6.50%	7.00%	7.50%	8.00%	8.50%	9.00%	9.50%	10.00%
1	1.055	1.06	1.065	1.07	1.075	1.08	1.085	1.09	1.095	1.1
2	0.54162	0.54544	0.54926	0.55309	0.55693	0.56077	0.56462	0.56847	0.57233	0.57619
3	0.37065	0.37411	0.37758	0.38105	0.38454	0.38803	0.39154	0.39505	0.39858	0.40211
4	0.28529	0.28859	0.2919	0.29523	0.29857	0.30192	0.30529	0.30867	0.31206	0.31547
5	0.23418	0.2374	0.24063	0.24389	0.24716	0.25046	0.25377	0.25709	0.26044	0.2638
6	0.20018	0.20336	0.20657	0.2098	0.21304	0.21632	0.21961	0.22292	0.22625	0.22961
7	0.17596	0.17914	0.18233	0.18555	0.1888	0.19207	0.19537	0.19869	0.20204	0.20541
8	0.15786	0.16104	0.16424	0.16747	0.17073	0.17401	0.17733	0.18067	0.18405	0.18744
9	0.14384	0.14702	0.15024	0.15349	0.15677	0.16008	0.16342	0.1668	0.1702	0.17364
10	0.13267	0.13587	0.1391	0.14238	0.14569	0.14903	0.15241	0.15582	0.15927	0.16275
11	0.12357	0.12679	0.13006	0.13336	0.1367	0.14008	0.14349	0.14695	0.15044	0.15396
12	0.11603	0.11928	0.12257	0.1259	0.12928	0.1327	0.13615	0.13965	0.14319	0.14676
13	0.10968	0.11296	0.11628	0.11965	0.12306	0.12652	0.13002	0.13357	0.13715	0.14078
14	0.10428	0.10758	0.11094	0.11434	0.1178	0.1213	0.12484	0.12843	0.13207	0.13575
15	0.09963	0.10296	0.10635	0.10979	0.11329	0.11683	0.12042	0.12406	0.12774	0.13147
16	0.09558	0.09895	0.10238	0.10586	0.10939	0.11298	0.11661	0.1203	0.12403	0.12782
17	0.09204	0.09544	0.09891	0.10243	0.106	0.10963	0.11331	0.11705	0.12083	0.12466
18	0.08892	0.09236	0.09585	0.09941	0.10303	0.1067	0.11043	0.11421	0.11805	0.12193
19	0.08615	0.08962	0.09316	0.09675	0.10041	0.10413	0.1079	0.11173	0.11561	0.11955
20	0.08368	0.08718	0.09076	0.09439	0.09809	0.10185	0.10567	0.10955	0.11348	0.11746
21	0.08146	0.085	0.08861	0.09229	0.09603	0.09983	0.1037	0.10762	0.11159	0.11562
22	0.07947	0.08305	0.08669	0.09041	0.09419	0.09803	0.10194	0.1059	0.10993	0.11401
23	0.07767	0.08128	0.08496	0.08871	0.09254	0.09642	0.10037	0.10438	0.10845	0.11257
24	0.07604	0.07968	0.0834	0.08719	0.09105	0.09498	0.09897	0.10302	0.10713	0.1113
25	0.07455	0.07823	0.08198	0.08581	0.08971	0.09368	0.09771	0.10181	0.10596	0.11017

Table A.2: Continued

	10.50%	11.00%	11.50%	12.00%	12.50%	13.00%	13.50%	14.00%	14.50%	15.00%
1	1.105	1.11	1.115	1.12	1.125	1.13	1.135	1.14	1.145	1.15
2	0.58006	0.58393	0.58781	0.5917	0.59559	0.59948	0.60338	0.60729	0.6112	0.61512
3	0.40566	0.40921	0.41278	0.41635	0.41993	0.42352	0.42712	0.43073	0.43435	0.43798
4	0.31889	0.32233	0.32577	0.32923	0.33271	0.33619	0.33969	0.3432	0.34673	0.35027
5	0.26718	0.27057	0.27398	0.27741	0.28085	0.28431	0.28779	0.29128	0.29479	0.29832
6	0.23298	0.23638	0.23979	0.24323	0.24668	0.25015	0.25365	0.25716	0.26069	0.26424
7	0.2088	0.21222	0.21566	0.21912	0.2226	0.22611	0.22964	0.23319	0.23677	0.24036
8	0.19087	0.19432	0.1978	0.2013	0.20483	0.20839	0.21197	0.21557	0.2192	0.22285
9	0.17711	0.1806	0.18413	0.18768	0.19126	0.19487	0.19851	0.20217	0.20586	0.20957
10	0.16626	0.1698	0.17338	0.17698	0.18062	0.18429	0.18799	0.19171	0.19547	0.19925
11	0.15752	0.16112	0.16475	0.16842	0.17211	0.17584	0.1796	0.18339	0.18722	0.19107
12	0.15038	0.15403	0.15771	0.16144	0.16519	0.16899	0.17281	0.17667	0.18056	0.18448
13	0.14445	0.14815	0.1519	0.15568	0.1595	0.16335	0.16724	0.17116	0.17512	0.17911
14	0.13947	0.14323	0.14703	0.15087	0.15475	0.15867	0.16262	0.16661	0.17063	0.17469
15	0.13525	0.13907	0.14292	0.14682	0.15076	0.15474	0.15876	0.16281	0.1669	0.17102
16	0.13164	0.13552	0.13943	0.14339	0.14739	0.15143	0.1555	0.15962	0.16376	0.16795
17	0.12854	0.13247	0.13644	0.14046	0.14451	0.14861	0.15274	0.15692	0.16112	0.16537
18	0.12586	0.12984	0.13387	0.13794	0.14205	0.1462	0.15039	0.15462	0.15889	0.16319
19	0.12353	0.12756	0.13164	0.13576	0.13993	0.14413	0.14838	0.15266	0.15698	0.16134
20	0.12149	0.12558	0.1297	0.13388	0.1381	0.14235	0.14665	0.15099	0.15536	0.15976
21	0.11971	0.12384	0.12802	0.13224	0.13651	0.14081	0.14516	0.14954	0.15396	0.15842
22	0.11813	0.12231	0.12654	0.13081	0.13512	0.13948	0.14387	0.1483	0.15277	0.15727
23	0.11675	0.12097	0.12524	0.12956	0.13392	0.13832	0.14276	0.14723	0.15174	0.15628
24	0.11552	0.11979	0.1241	0.12846	0.13287	0.13731	0.14179	0.1463	0.15085	0.15543
25	0.11443	0.11874	0.1231	0.1275	0.13194	0.13643	0.14095	0.1455	0.15008	0.1547

EPA/452/B-02-001

Section 2

Generic Equipment and Devices

EPA/452/B-02-001

Chapter 1

Hoods, Ductwork and Stacks

William M. Vatavuk
Innovative Strategies and Economics Group
Air Quality Strategies and Standards Division
Office of Air Quality Planning and Standards
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

December 1995

Contents

1.1 Introduction 1-3

1.2 Equipment Description 1-3
1.2.1 Hoods 1-4
1.2.1.1 Types of Hoods 1-4
1.2.2 Ductwork 1-6
1.2.2.1 Ductwork Components 1-8
1.2.3 Stacks 1-12

1.3 Design Procedures 1-13
1.3.1 Design Fundamentals 1-13
1.3.1.1 The Bernoulli Equation 1-13
1.3.1.2 Pressure: Static, Velocity, and Total 1-17
1.3.1.3 Temperature and Pressure Adjustments 1-20
1.3.2 Hood Design Procedure 1-21
1.3.2.1 Hood Design Factors 1-21
1.3.2.2 Hood Sizing Procedure 1-26
1.3.3 Ductwork Design Procedure 1-28
1.3.3.1 Two Ductwork Design Approaches 1-28
1.3.3.2 Ductwork Design Parameters 1-28
1.3.3.3 Ductwork Pressure Drop 1-31
1.3.4 Stack Design Procedures 1-35
1.3.4.1 Calculating Stack Diameter 1-36
1.3.4.2 Calculating Stack Height 1-37
1.3.4.3 Calculating Stack Draft 1-38

1.4 Estimating Total Capital Investment 1-40
1.4.1 Equipment Costs 1-40
1.4.1.1 Hood Costs 1-40
1.4.1.2 Ductwork Costs 1-41
1.4.1.3 Stack Costs 1-46
1.4.2 Taxes, Freight, and Instrumentation Costs 1-59
1.4.3 Purchased Equipment Cost 1-50
1.4.4 Installation Costs 1-50

1.5 Estimating Total Annual Cost 1-51
1.5.1 Direct Annual Costs 1-51
1.5.2 Indirect Annual Costs 1-52
1.5.3 Total Annual Cost 1-51

1.6 Acknowledgements 1-55

1.1 Introduction

Most control devices are located some distance from the emission sources they control. This separation may be needed for several reasons. For one thing, there may not be enough room to install the control device close to the source. Or, the device may collect emissions from several sources located throughout the facility and, hence, must be sited at some convenient, equidistant location. Or, it may be that required utility connections for the control device are only available at some remote site. Regardless of the reason, the waste gas stream must be conveyed from the source to the control device and from there to a stack before it can be released to the atmosphere.

The type of equipment needed to convey the waste gas are the same for most kinds of control devices. These are: (1) hoods, (2) ductwork, (3) stacks, and (4) fans. Together, these items comprise a ventilation system. A hood is used to capture the emissions at the source; ductwork, to convey them to the control device; a stack, to disperse them after they leave the device; and a fan, to provide the energy for moving them through the control system. This section covers the first three kinds of equipment. However, because they constitute such a broad and complex subject, fans will be dealt with in a another section of this Manual to be developed in the future. Only short stacks (100-120 feet high or less) are covered. Typically, short stacks are included with packaged control systems or added to them. So-called "tall stacks" ("chimneys"), used at power plants or other sources where the exhaust gases must be dispersed over great distances, will not be discussed in this section.

This section presents all the information one would need to develop study (± 30%-accurate) cost estimates for hoods, ductwork, and stacks. Accordingly, the following sections include: (1) descriptions of the types of equipment used in air pollution control ventilation systems, (2) procedures for sizing (designing) this equipment, and (3) methodologies and data for estimating their capital and annual costs. Throughout the chapter are several illustrations (example problems) that show the reader how to apply the various sizing and costing methodologies.

1.2 Equipment Description

In this section, the kinds of hoods, ductwork, and stacks used in air pollution control systems are described, each in a separate subsection. These descriptions have been based on information obtained from standard ventilation and air pollution control references, journal articles, and equipment vendors.

1.2.1 Hoods

Of the several components of an air pollution control system, the capture device is the most important. This should be self-evident, for if emissions are not efficiently captured at the source they cannot be conveyed to and removed by a control device. There are two general categories of capture devices: (1) direct exhaust connections (DEC) and (2) hoods. As the name implies, a DEC is a section of duct (typically an elbow) into which the emissions directly flow. These connections often are used when the emission source is itself a duct or vent, such as a process vent in a chemical manufacturing plant or petroleum refinery. (See following discussion on "Ductwork".)

Hoods comprise a much broader category than DECs. They are used to capture particulates, gases, and/or mists emitted from a variety of sources, such as basic oxygen steelmaking furnaces, welding operations, and electroplating tanks. The hooded processes are generally categorized as either "hot" or "cold," a delineation that, in turn, influences hood selection, placement, and design.

The source characteristics also influence the materials from which a hood is fabricated. Mild (carbon) steel is the material of choice for applications where the emission stream is noncorrosive and of moderate temperature. However, where corrosive substances (e.g., acid gases) are present in high enough concentrations, stainless steels or plastics (e.g., fiberglass-reinforced plastic, or FRP) are required. As most hoods are custom-designed and built, the vendor involved would determine which material would be optimal for a given application.

1.2.1.1 Types of Hoods

Although the names of certain hoods vary, depending on which ventilation source one consults, there is general agreement as to how they are classified. There are four types of hoods: (1) enclosures, (2) booths, (3) captor (capture) hoods, and (4) receptor (receiving) hoods.[1,2]

Enclosures are of two types: (1) those that are completely closed to the outside environment and (2) those that have openings for material input/output. The first type is only used when handling radioactive materials, which must be handled by remote manipulators. They are also dust- and gas-tight. These kinds of enclosures are rarely used in air pollution control. The second type, have applications in several areas, such as the control of emissions from electric arc furnaces and from screening and bin filling operations. They are equipped with small wall openings (natural draft openings—"NDOs") that allow for material to be moved in or out and for ventilation. However, the area of these openings must be small compared with the total area of the enclosure walls (typically, 5% or less).

Another application of total enclosures is in the measurement of the capture efficiency of volatile organic compound (VOC) control devices. Capture efficiency is that fraction of all VOCs generated at, and released by, an affected facility that is directed to the control device. In this application, a total enclosure is a temporary structure that completely surrounds an emitting process so that all VOC emissions are captured for discharge through ducts or stacks. The air flow through the total enclosure must be high enough to keep the concentration of the VOC mixture inside the enclosure within both the Occupational Safety and Health Administration (OSHA) health requirement limits and the vapor explosive limits. (The latter are typically set at 25% of the lower explosive limit (LEL) for the VOC mixture in question.) In addition, the overall face velocity of air flowing through the enclosure must be at least 200 ft/min.[3]

The surfaces of temporary total enclosures are usually constructed either of plastic film or of such rigid materials as insulation panels or plywood. Plastic film offers the advantages of being lightweight, transparent, inexpensive, and easy to work with. However, it is flimsy, flammable, and has a relatively low melting point. In addition, the plastic must be hung on a framework of wood, plastic piping, or scaffolding.

Although rigid materials are more expensive and less workable than plastic, they are more durable and can withstand larger pressure differentials between the enclosure interior and exterior. Total enclosure design specifications (which have been incorporated into several EPA emission standards) are contained in the EPA report, The Measurement Solution: Using a Temporary Total Enclosure for Capture Testing.[4]

Booths are like enclosures, in that they surround the emission source, except for a wall (or portion thereof) that is omitted to allow access by operators and equipment. Like enclosures, booths must be large enough to prevent particulates from impinging on the inner walls. They are used with such operations (and emission sources) such as spray painting and portable grinding, polishing, and buffing operations.

Captor hoods (also termed active or external hoods) do not enclose the source at all. Consisting of one to three sides, they are located at a short distance from the source and draw the emissions into them via fans. Captor hoods are further classified as side-draft/backdraft, slot, downdraft, and high-velocity, low-volume (HVLV) hoods.

A side-draft/back-draft hood is typically located to the side/behind of an emission source, but as close to it as possible, as air velocities decrease inversely (and sharply) with distance. Examples of these include snorkel-type welding hoods and side shake-out hoods.

A slot hood operates in a manner similar to a side-draft/back-draft. However, the inlet opening (face) is much smaller, being long and narrow. Moreover, a slot hood is situated at the periphery of an emission source, such as a narrow, open tank. This type of hood is also employed with bench welding operations.

While slot and side-draft/back-draft hoods are located beside/behind a source, a downdraft hood is situated immediately beneath it. It draws pollutant-laden air down through the source and, thence, to a control device. Applications of down-draft hoods include foundry shake-out and bench soldering and torch cutting operations.

HVLV hoods are characterized by the use of extremely high velocities (capture velocities) to collect contaminants at the source, and by the optimal distribution of those velocities across the hood face. To maintain a low volumetric flow rate, these hoods are located as close to the source as possible, so as to minimize air entrainment.

The last category is receptor hoods (passive or canopy hoods). A receptor hood typically is located above or beside a source, to collect the emissions, which are given momentum by the source. For example, a canopy hood might be situated directly above an open tank containing a hot liquid (a buoyant source). With entrained air, vapors emitted from the liquid would rise into the hood. Here, the canopy hood would function as a passive collector, as the rising gases would be drawn into the hood via natural draft. (See Figure 1.1.)

Receptor hoods are also used with nonbuoyant sources, sources from which emissions do not rise. But are "thrown off" from a process, such as a swing grinder. The initial velocity of the emissions typically is high enough to convey them into a receiving hood.[5]

1.2.2 Ductwork

Once the emission stream is captured by either a hood or a direct exhaust connection, it is conveyed to the control device via ductwork. The term "ductwork" denotes all of the equipment between the capture device and the control device. This includes: (1) straight duct; (2) fittings, such as elbows and tees; (3) flow control devices (e.g., dampers); and (4) duct supports. These components are described in Section 1.2.2.1.)

In air pollution control systems, the fan is usually located immediately before or after the control device. Consequently, most of the ductwork typically is under a negative static pressure, varying from a few inches to approximately 20 inches of water column. These pressure conditions dictate the type of duct used, as well as such design parameters as the wall thickness (gauge). For instance, welded duct is preferable to spiral-wound duct in vacuum applications.[6]

Ductwork is fabricated from either metal or plastic, the choice of material being dictated by the characteristics of the waste gas stream, structural considerations, purchase and installation costs, aesthetics, and other factors. Metals used include carbon steel (bare or galvanized), stainless steel, and aluminum. The most commonly used plastics are polyvinyl chloride (PVC) and fiberglass-reinforced plastic (FRP), although polypropylene (PP) and linear polyethylene (LPE) also can also be applied. However, one serious drawback to PP and LPE is that both are combustible.[7]

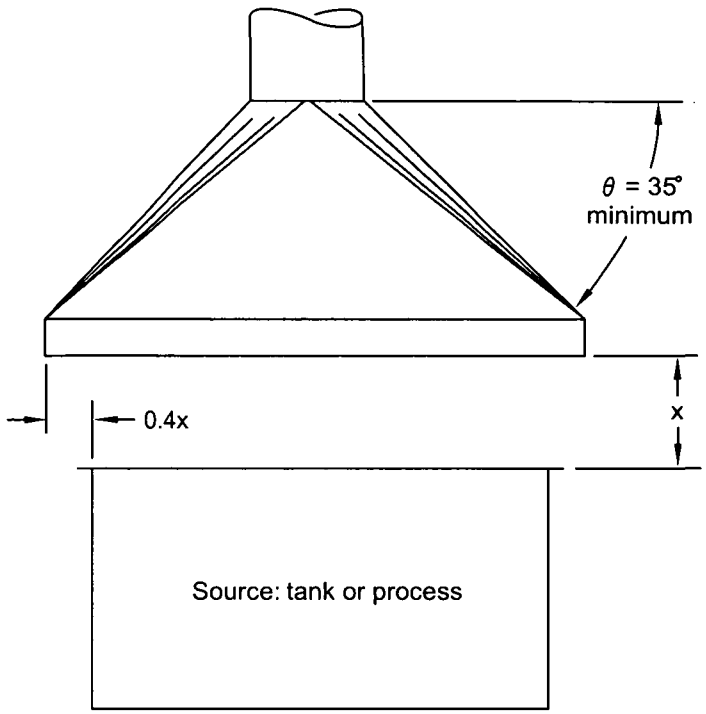


Figure 1.1: Typical Canopy Hood Installation

PVC and other plastic ductwork are resistant to a variety of corrosive substances, from aqua regia to 95% sulfuric acid. But plastic ductwork cannot tolerate environmental temperatures above 150°F.[8] Metal ductwork can handle temperatures up to approximately 1000°F, but only certain alloys can tolerate corrosive streams.

In terms of construction, ductwork can be either rigid or flexible. As the name implies, rigid ductwork, whether metal or plastic, has a fixed shape. Conversely, flexible ductwork can be bent to accommodate situations where space is limited or where the layout is so convoluted that rigid fittings cannot meet construction requirements. Usually circular in cross-sectional shape, flexible duct can be fabricated from metals or plastic and can be either insulated or uninsulated.

Rigid ductwork is fabricated into circular, flat oval, or square/rectangular cross-sectional shapes. Of these, circular duct is most commonly used in air pollution control systems. Although square/rectangular duct is advantageous to use when space is limited, round duct offers several advantages. It resists collapsing, provides better transport conditions, and uses less metal than

square/rectangular or flat oval shapes of equivalent cross-sectional area.[9] Unless otherwise noted, the following discussion pertains to rigid, circular duct, as this is the type most commonly used in air pollution control.

Rigid metal circular duct is further classified according to method of fabrication. Longitudinal seam duct is made by bending sheet metal into a circular shape over a mandrel, and buttwelding the two ends together. Spiral seam duct is constructed from a long strip of sheet metal, the edges of which are joined by an interlocking helical seam that runs the length of the duct. This seam is either raised or flush to the duct wall surface.

Fabrication method and cross-sectional shape are not the only considerations in designing ductwork, however. One must also specify the diameter; wall thickness; type, number, and location of fittings, controllers, and supports; and other parameters. Consequently, most ductwork components are custom designed and fabricated, so as to optimally serve the control device. Some vendors offer prefabricated components, but these are usually common fittings (e.g., 90° elbows) that are available only in standard sizes (e.g., 3- to 12-inch diameter).[10,11]

If either the gas stream temperature or moisture content is excessive, the ductwork may need to be insulated. Insulation inhibits heat loss/gain, saving energy (and money), on the one hand, and prevents condensation, on the other. Insulation also protects personnel who might touch the ductwork from sustaining burns. There are two ways to insulate ductwork. The first is to install insulation on the outer surface of the ductwork and cover it with a vapor barrier of plastic or metal foil. The type and thickness of insulation used will depend on the heat transfer properties of the material. For instance, one vendor states that 4 inches of mineral wool insulation is adequate for maintaining a surface ("skin") temperature of 140°F (the OSHA workplace limit) or lower, provided that the exhaust gas temperature does not exceed 600°F. [12]

The second way to insulate ductwork is by using double-wall, insulated duct and fittings. Double-wall ductwork serves to reduce both heat loss and noise. One vendor constructs it from a solid sheet metal outer pressure shell and a sheet metal inner liner with a layer of fiberglass insulation sandwiched between. The insulation layer is typically 1-inch, although 2- and 3-inch thicknesses are available for more extreme applications. The thermal conductivities of these thicknesses are 0.27, 0.13, and 0.09 Btu/hr-ft^2-°F, respectively.[13]

1.2.2.1 Ductwork Components

As discussed above, a ductwork system consists of straight duct, fittings, flow control devices, and supports. Straight duct is self-explanatory and easy to visualize. The "fittings" category, however, encompasses a range of components that perform one or more of the following functions: change the direction of the ducted gas stream, modify the stream velocity, tie it to another duct(s), facilitate the connection of two or more components, or provide for expansion/contraction when thermal stresses arise.

The most commonly used fittings are elbows ("ells"). These serve to change the gas stream direction, typically by 30°, 45°, 60°, or 90°, though they may be designed for other angles as well. The elbow centerline radius determines the rate at which this directional change occurs. (See Figure 1.2.) The standard centerline radius (R_{cl}) is 1.5 times the elbow cross-sectional diameter (D_c). However, in "long-radius" elbows, in which the directional change is more gradual than in standard elbows, R_{cl} is greater than or equal to 2 times D_c.[14]

A T-shaped fitting ("tee") is used when two or more gas streams must be connected. In straight tees, the streams converge at a 90° angle, while in angle tees ("laterals", "wyes") the connection is made at 30°, 45°, 60°, or some other angle. (See Figure 1.2.) Tees may have one "tap" (connection) or two, and may have either a straight or a "conical" cross-section at either or both ends. Crosses are also used to connect duct branches. Here, the two branches intersect each other at a right angle.

Reducers (commonly called "expansions" or "contractions") are required whenever ducts of different diameter must be joined. Reducers are either concentric or eccentric in design. In concentric reducers, the diameter tapers gradually from the larger to smaller cross section. However, in eccentric reducers, the diameter decreases wholly on one side of the fitting.

Dampers control the volumetric flowrate through ventilation systems. They are usually delineated according to the flow control mechanism (single blade or multiblade), pressure rating (low/light or high/heavy), and means of control (manual or automatic). In single blade dampers, a circular plate is fastened to a rod, one end of which protrudes outside the duct. In the most commonly used type of single blade damper (butterfly type), this rod is used to control the gas flow by rotating the plate in the damper. Fully closed, the damper face sits perpendicular to the gas flow direction; fully open, the face is parallel to the gas flow lines. Several single blade "control" dampers are depicted in Figure 1.2.

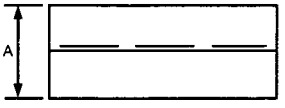
Blast gate dampers control the flow by sliding the damper blade in and out of the duct. Blast gates are often used to control the flow of air streams containing suspended solids, such as in pneumatic conveyors. In these respects, butterfly dampers and blast gates are analogous, respectively, to the globe valves and quick-opening gate valves that are used to regulate liquid flow in pipes.

Multiblade (louvered) dampers operate on the same principal as single blade dampers. However, instead of using a single blade or plate to control the gas flow, multiblade dampers employ slats that open and close like venetian blinds.[15] Louvered dampers typically are used in very large ducts where a one-piece damper blade would be too difficult to move.

Manually-controlled dampers simply have a handle attached to the control rod which is used to adjust the gas flow by hand. If automatic control is needed, a pneumatic or electronic

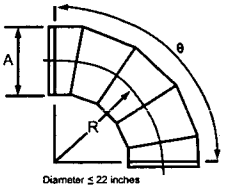
LONGITUDINAL SEAM DUCT

(Solid welded longitudinal seam)



DIMENSIONS:
8-inch minimum
90-inch maximum

GORED ELBOW



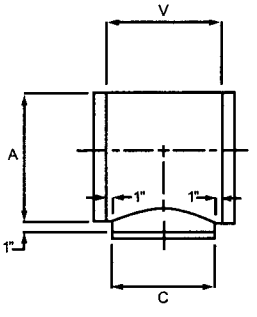
DIMENSIONS:
R = 1.5A

Where:

θ	Number of gores
0 - 35°	2
36 - 71°	3
72 - 90°	5

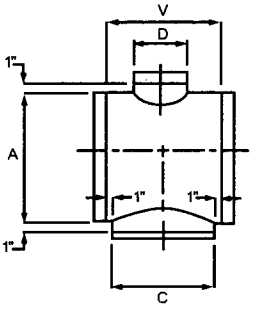
For elbows where θ exceeds 90°, add one gore for each additional 18° or fraction thereof.

STRAIGHT TEE



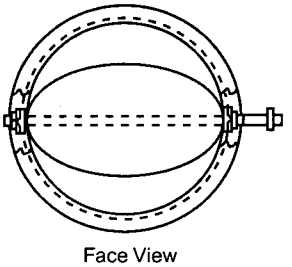
DIMENSIONS:
V = C + 2
Maximum C = A

STRAIGHT 90° CROSS

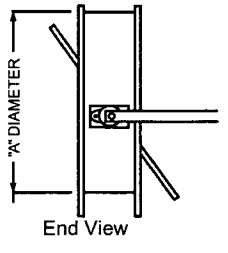


DIMENSIONS:
V = C + 2
Maximum C or D = A

DAMPERS

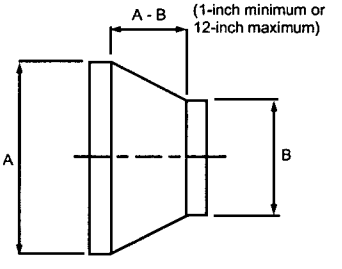


Face View



End View

CONCENTRIC REDUCER



ECCENTRIC REDUCER

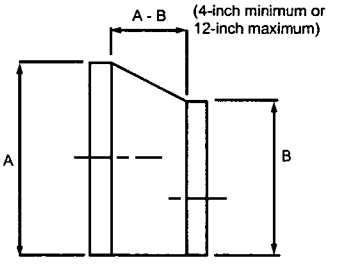


Figure 1.2: Selected Circular Ductwork Components[14]

actuator is used. The actuator receives a pneumatic (pressurized air) or electrical signal from a controller and converts it to mechanical energy which is used, in turn, to open/close the damper via the damper rod. In this respect, an actuated damper is analogous to an automatic control valve.[16] For example, an automatic damper may be used to control the dilution air flow rate to an incinerator combustion chamber. This flow rate, in turn, would depend on the combustibles concentration (i.e., percentage of lower explosive limit—%LEL) in the inlet waste gas stream. If this concentration deviates from a predetermined amount ("set point"), a signal is sent from the measuring device via the controller to the automatic damper to increase/decrease the dilution air flow rate so as to maintain the desired %LEL.

Expansion joints are installed, especially in longer metal duct runs, to allow the ductwork to expand or contract in response to thermal stresses. These fittings are of several designs. One type, the bellows expansion joint, consists of a piece of flexible metal (e.g., 304 stainless steel) that is welded to each of two duct ends, connecting them. As the temperature of the duct increases, the bellows compresses; as the duct temperature decreases, the bellows expands. Another commonly used expansion joint consists of two flanges between which is installed a section of fabric. Like the bellows expansion joint, it compresses as the duct temperature increases, and vice-versa. The temperature dictates the type of fabric used. For instance, silicone fiberglass and aramid fiber cloth can be used for duct temperatures of up to 500°F., while coated fiberglass cloth is needed to accommodate temperatures of 1000°F.[17]

The last component to consider is the ductwork support system. However, it is far from being the least important. As the Sheet Metal and Air Conditioning Contractors' National Association (SMACNA) HVAC Duct Construction Standards manual states, "The selection of a hanging system should not be taken lightly, since it involves not only a significant portion of the erection labor, but also because [the erection of] an inadequate hanging system can be disastrous." As a rule, a support should be provided for every 8 to 10 feet of duct run.[18] Ductwork can be suspended from a ceiling or other overhead structure via hangers or supported from below by girders, pillars, or other supports.

A suspension arrangement typically consists of an upper attachment, a hanger, and a lower attachment. The upper attachment ties the hanger to the ceiling, etc. This can be a concrete insert, an eye bolt, or a fastener such as a rivet or nailed pin. The hanger is generally a strap of galvanized steel, round steel rod, or wire that is anchored to the ceiling by the upper attachment. The type of hanger used will be dictated by the duct diameter, which is proportional to its weight per lineal foot. For instance, wire hangers are only recommended for duct diameters up to 10 inches. For larger diameters (up to 36 inches), straps or rods should be used. Typically, a strap hanger is run from the upper attachment, wrapped around the duct, and secured by a fastener (the lower attachment). A rod hanger also extends down from the ceiling. Unlike strap hangers, they are fastened to the duct via a band or bands that are wrapped around the circumference. Duct of diameters greater than 3 feet should be supported with two hangers, one on either side of the duct, and be fastened to two

circumferential bands, one atop and one below the duct.[19] Moreover, supports for larger ductwork should also allow for both axial and longitudinal expansion and contraction, to accommodate thermal stresses.[20]

1.2.3 Stacks

Short stacks are installed after control devices to disperse the exhaust gases above ground level and surrounding buildings. As opposed to "tall" stacks, which can be up to 1,000 feet high, short stacks typically are no taller than 120 feet. Certain packaged control devices come equipped with short ("stub") stacks, with heights ranging from 30 to 50 feet. But if such a stack is neither provided nor adequate, the facility must erect a separate stack to serve one or more devices. Essentially, this stack is a vertical duct erected on a foundation and supported in some manner. For structural stability, the diameter of the stack bottom is slightly larger than the top diameter, which typically ranges from 1 to 7 feet.[21]

A short stack may be fabricated of steel, brick, or plastic (e.g., fiberglass-reinforced plastic, or FRP). A stack may be lined or unlined. The material selection depends on the physical and chemical properties of the gas stream, such as corrosiveness and acidity, as well as the temperature differential between the gas stream and the ambient air. Liners of stainless steel, brick, or FRP usually are used to protect the stack against damage from the gas stream. They are much easier and less expensive to replace than the entire stack. Alternatively, the interior of an unlined stack may be coated with zinc (galvanized), aluminum, or another corrosion-resistant material, but a coating does not provide the same protection as a liner and does not last as long.

Short stacks are either self-supporting (free-standing), supported by guy wires, or fastened to adjacent structures. The type of support used depends on the stack diameter, height and weight, the wind load, local seismic zone characteristics, and other factors.

Auxiliary equipment for a typical stack includes an access door, a sampling platform, ladders, lightning protection system, and aircraft warning lights. The access door allows for removal of any accumulated materials at the bottom of the stack and provides access to the liner for repair or replacement. Local and state air pollution control regulations also may require the permanent installation of sampling platforms for use during periodic compliance tests, while ladders are used both during stack sampling and maintenance procedures. The lightning protection system is needed to prevent damage to the stack and immediate surroundings during electrical storms. Lastly, aircraft warning lights are required by local aviation authorities.[23] Altogether, these auxiliaries can add a large amount to the base stack cost.

1.3 Design Procedures

As stated above, a hood, ductwork, and a stack are key elements in any air pollution control system. Because each of these elements is different, both in appearance and function, each must be designed separately. But at the same time, these elements comprise a system, which is governed by certain physical laws that serve to unite these elements in "common cause." Thus, before the individual design procedures for hoods, ductwork, and stacks are described, ventilation fundamentals will be presented. These fundamentals will cover basic fluid flow concepts and how they may be applied to air pollution control ventilation systems. Nonetheless, these concepts will be given as straightforwardly as possible, with the aim of making the design parameters easy to understand and compute.

1.3.1 Design Fundamentals

1.3.1.1 The Bernoulli Equation

The flow of fluids in any hood, duct, pipe, stack, or other enclosure is governed by a single relationship, the familiar Bernoulli equation. Put simply and ideally, the Bernoulli equation states that the total mechanical energy of an element of flowing fluid is constant throughout the system. This includes its potential energy, kinetic energy, and pressure energy. However, as no system is ideal, the Bernoulli equation must be adjusted to take into account losses to the surroundings due to friction. Gains due to the energy added by fans, pumps, etc., also must be accounted for. For a pound mass (lb_m) of fluid flowing in a steady-state system the adjusted Bernoulli equation is:[24]

$$\int_1^2 v\,dp + \Delta z\left(\frac{g}{g_c}\right) + \frac{\Delta\left(u^2\right)}{2g_c} = W - F \qquad (1.1)$$

where

v	=	specific volume of fluid (ft^3/lb_m)
p	=	static pressure—gauge (lb_f/ft^2)
z	=	height of fluid above some reference point (ft)
u	=	fluid velocity through duct, hood, etc. (ft/sec)
g	=	gravitational acceleration (ft/sec^2)
g_c	=	gravitational constant (32.174 ($[lb_m\text{-}ft/sec^2]/lb_f$)
W	=	work added by fan, etc. ($ft\text{-}lb_f/lb_m$)
F	=	energy lost due to friction ($ft\text{-}lb_f/lb_m$)

Each of the terms on the left hand side of Equation 1.1 represents an energy change to a pound mass of fluid between two locations in the system—points "1" and "2." The work (W) and friction (F) terms denote the amounts of energy added/lost between points 1 and 2.

Note that the units of each term in Equation 1.1 are "ft-lb_f/lb_m," energy per unit mass. In the English system of units, "lb_f" and "lb_m" are, for all intents, numerically equivalent, since the ratio of the gravitational acceleration term (g) to the gravitational constant (g_c) is very close to 1. In effect the equation unit are "feet of fluid" or "fluid head in feet." In air pollution control situations, the fluid often has the properties of air. That is because the contaminants in the waste gas stream are present in such small amounts that the stream physical properties approximate those of pure air.

Because air is a "compressible" fluid, its specific volume is much more sensitive to changes in pressure and temperature than the specific volume of such "incompressible" fluids such as water. Hence, the "vdp" term in Equation 1.1 has to be integrated between points 1 and 2. However, in most air pollution control ventilation systems neither the pressure nor the temperature changes appreciably from the point where the emissions are captured to the inlet of the control device. Consequently, the specific volume is, for all practical purposes, constant throughout the ventilation system, and one does not have to integrate the vdp term. With this assumption, the first term in Equation 1.1 becomes simply:

$$\int_1^2 v\,dp \;=\; v\int_1^2 dp \;=\; v\Delta p \qquad (1.2)$$

Illustration: VOC emitted by an open tank is captured by a hood and conveyed, via a blower, through 150 feet of 12-inch diameter ductwork to a refrigerated condenser outdoors. The blower, which moves the gas through the hood, ductwork, and condenser, is located immediately before the inlet to the condenser. Thus, the entire ventilation system is under vacuum. The stream temperature and absolute pressure are 100°F and approximately 1 atmosphere (14.696 lb_f/in^2), respectively. The elevation of the refrigerated condenser inlet is 30 feet below that of the tank. The air velocity at the source is essentially zero, while the duct transport velocity is 2,000 ft/min. The static gauge pressure increases from -0.50 in. w.c. (water column) at the source to 4.5 in. w.c. at the blower outlet. Finally, the calculated friction loss through the ductwork and hood totals 1.25 in. w.c. Calculate the amount of mechanical energy that the blower adds to the gas stream. Assume that the gas temperature remains constant throughout.

Solution:

First, develop a factor to convert "inches of water" to "feet of air":

$$Feet\ of\ air = (Inches\ of\ water)\left(\frac{1\,ft}{12\,in}\right)\left(\frac{v_{a100}}{v_{w100}}\right) \qquad (1.3)$$

where

v_{w100} = specific volume of water @ 100°F = 0.01613 ft^3/lb_m
v_{a100} = specific volume of air @ 100°F, 1 atmosphere

Because the system absolute pressure is close to atmospheric, the waste gas behaves as an ideal gas. Thus, the specific volume can be calculated from the ideal gas law:

$$v_a = \frac{RT}{pM} \tag{1.4}$$

where

R = ideal gas constant = 1,545 ft-lb_f/(lb_m-mole)(ER)
T = absolute temperature of gas = 100 + 460 = 560ER
M = molecular weight of gas (air) = 28.85 lb_m/lb_m-mole
p = absolute pressure = 2,116 lb_f/ft^2

Substituting, we obtain:

$$v_a = 14.17 \frac{ft^3}{lb_m}$$

Finally, substitution of these values for v_a and v_w into Equation 1.3 yields:

$$Feet\ of\ air\ (@\ 100\ °F,\ 1\ atm) = 73.207 \times inches\ of\ water$$

Compute the changes in the mechanical energy terms and the friction losses between the hood inlet (point 1) and the blower outlet/condenser inlet (point 2):

Pressure: $v\Delta p = (4.5 - [-0.50]\ in.\ w.c.)\left(\frac{73.207\ ft\ air}{in.\ w.c.}\right) = 336.0\ ft\ air$

Potential: Δz = -30 ft air (point 2 is below point 1)

Kinetic: $$\Delta u^2 / 2g_c = \left(\frac{\left[2{,}000 \ \frac{ft}{min} \right]}{\left[\frac{60 \ \frac{ft}{min}}{1 \ \frac{ft}{sec}} \right]} \right) \times \left(\frac{1}{2}\right) \left(\frac{32.174 \left[lb_m - \frac{ft}{sec^2} \right]}{lb_f} \right)^{-1} = 17.3 \ ft \ air$$

Friction losses: $F = 1.25 \ in. \ w.c \times 73.207 = 91.5 \ ft \ air$

Substitute above results into Equation 1.1 and solve for W, the fan energy added:

$$366.0 + (-30) + 17.1 = W - 91.5, \text{ or}$$

$$W = 44.6 \ \frac{ft - lb_f}{lb_m \ air} = 6.07 \ in. \ w.c.$$

To convert the fan energy input, W, to horsepower (hp_f), we would have to multiply it by the air mass flow rate (lb_m/sec), and divide the result by the horsepower conversion factor, 550 ft-lb_f/sec-hp. However, the mass flow rate is just the volume flow rate (Q, ft^3/sec) divided by the specific volume:

$$hp_f = W \left(\frac{Q}{v_a} \right) \left(\frac{1}{550} \right) = 0.001818 \ \frac{WQ}{v_a} \qquad (1.5)$$

In turn, Q is a function of the duct velocity (u_t, ft/sec) and duct diameter (D_d, ft):

$$Q = u_t \left(\frac{\pi D_d^{\,2}}{4} \right) \qquad (1.6)$$

Equation 1.6 applies, of course, only to circular ducts.

If we combine Equations 1.5 and 1.6 and substitute the inputs for this illustration, we obtain:

$$hp_f = 444.6 \left(\frac{2{,}000}{60} \right) \left(\frac{\pi}{4} \right) (1)^2 \left(\frac{1}{14.17} \right) \left(\frac{1}{550} \right) = 1.49 \ hp$$

Some observations about this illustration:

- Recall that the precise units for W and the other terms in Equation 1.1 are "ft-lb_f per b_m air," which, for convenience, have been shortened to "ft air". Thus, they measure energy, not length.

- Compared to the pressure energy and friction terms, the potential and kinetic energy terms are small. Had they been ignored, the results would not have changed appreciably.

- The large magnitude of the pressure and friction terms clearly illustrates the importance of keeping one's units straight. As shown in step (1), one inch of water is equivalent to over 73 feet of air. However, as Equation 1.3 indicates, the pressure corresponding to equivalent heights of air and water columns would be the same.

- The fan power input depends not just on the total "head" (ft air) required, but also on the gas flow rate. Also, note that the horsepower computed via Equation 2.5 is a theoretical value. It would have to be adjusted to account for the efficiencies of the fan and fan motor. The fan efficiency ranges from 40 to 70 percent, while the motor efficiency is typically 90 percent. These efficiencies are usually combined into a single efficiency (ϵ, fraction), by which the theoretical horsepower is divided to obtain the actual horsepower requirement.

1.3.1.2 Pressure: Static, Velocity, and Total

Although it is more rigorous and consistent to express the Bernoulli equation terms in terms of feet of air (or, precisely, ft-lb_f/lb_m of air), industrial ventilation engineers prefer to use the units "inches of water column (in. w.c.)." These units were chosen because, as the above illustration shows, results expressed in "feet of air" are often large numbers that are cumbersome to use. In addition, the total pressure changes in ventilation systems are relatively small, compared to those in liquid flow systems. Total pressure changes expressed in inches of mercury would be small numbers which are just as awkward to work with as large numbers. Hence, "inches of water" is a compromise, as values expressed in this measurement unit typically range from only 1 to 10. Moreover, practical measurement of pressure changes is done with water-filled manometers.

In the previous paragraph, a new quantity was mentioned, total pressure (TP). Also known as the "impact pressure", the total pressure is the sum of the static gauge (SP) and velocity pressures (VP) at any point within a duct, hood, etc., all expressed in in. w.c.[25] That is:

$$TP = SP + VP \qquad (1.7)$$

where

$SP = (cf)vp$
$VP = (cf)u^2/2g_c$

The "*cf*" in the expressions for *SP* and *TP* is the factor for converting the energy terms from "ft air" to "in. w.c.", both at standard temperature and absolute pressure (70°F, 1 atmosphere). (Again, keep in mind that, regardless of what units *SP* or *VP* are expressed in, the actual units are "energy per unit mass.") This conversion factor, *cf*, would be obtained via rearranging Equation 1.3:

$$cf = \frac{in.w.c.}{ft.\ air} = 12\left(\frac{v_{w70}}{v_{a70}}\right) \tag{1.8}$$

where

v_{w70} = specific volume of water at 70°F = 0.01605 (ft^3/lb_m)
v_{a70} = specific volume of air at 70°F = 13.41 (ft^3/lb_m)

Thus: $cf = 0.01436$ in. w.c./ft air

Clearly, the conversion factor varies as a function of temperature and pressure. For instance, at 100°F and 1 atmosphere, $cf = 1/73.207 = 0.01366$.

Conspicuously absent from Equation 1.7 is the potential energy term, "$z(g/g_c)$". This omission was not inadvertent. In ventilation systems, the potential energy (*P.E.*) is usually small compared to the other terms. (For example, see illustration above.) The *P.E.* is, of course, a function of the vertical distance of the measurement point in question from some datum level, usually the ground. At most, that distance would amount to no more than 20 or 30 feet, corresponding to a *P.E.* of approximately 0.3 to 0.4 in. w.c. Consequently, we can usually ignore the *P.E.* contribution in ventilation systems without introducing significant error.

The static gauge pressure in a duct is equal in all directions, while the velocity pressure, a function of the gas velocity, varies across the duct face. The duct velocity is highest at the center and lowest at the duct walls. However, for air flowing in a long, straight duct, the average velocity across the duct (u_t) approximates the center line velocity (u_{cl}).[26] This is an important point, for the average duct velocity is often measured by a pitot tube situated at the center of the duct.

By substituting for *cf* in Equation 1.7, we can obtain a simple equation that relates velocity to velocity pressure (VP) at standard conditions:

$$VP = \frac{0.0143 u_t^{\,2}}{2 g_c} \tag{1.9}$$

Solving:

$$u_t \left(\frac{ft}{sec} \right) = 66.94 \ (VP)^{1/2} \tag{1.10}$$

Or:

$$u_t \left(\frac{ft}{min} \right) = 4{,}016 \ (VP)^{1/2} \tag{1.11}$$

Incidentally, these equations apply to any duct, regardless of its shape.

As Burton describes it, static gauge pressure can be thought of as the "stored" energy in a ventilation system. This stored energy is converted to the kinetic energy of velocity and the losses of friction (which are mainly heat, vibration, and noise). Friction losses fall into several categories:[27]

- Losses through straight duct
- Losses through duct fittings—elbow tees, reducers, etc.
- Losses in branch and control device entries
- Losses in hoods due to turbulence, shock, vena contracta
- Losses in fans
- Losses in stacks

These losses will be discussed in later sections of this chapter. Generally speaking, much more of the static gauge pressure energy is lost to` friction than is converted to velocity pressure energy. It is customary to express these friction losses (ΔSP_f) in terms of the velocity pressure:

$$F = \Delta SP_f = kVP \tag{1.12}$$

where

k = experimentally-determined loss factor (unitless)

Alternatively, Equations 1.11 and 1.12 may be combined to express F (in. w.c.) in terms of the average duct velocity, u_t (ft/min):

$$F = (6.200 \times 10^{-8})\, k u_t^2 \tag{1.13}$$

1.3.1.3 Temperature and Pressure Adjustments

Equations 1.8 to 1.13 were developed assuming that the waste gas stream was at standard temperature and pressure. These conditions were defined as 70°F and 1 atmosphere (14.696 lb_f/in^2), respectively. While 1 atmosphere is almost always taken as the standard pressure, several different standard temperatures are used in scientific and engineering calculations: 32°F, 68°F, and 77°F, as well as 70°F. The standard temperature selected varies according to the industry or engineering discipline in question. For instance, industrial hygienists and air conditioning engineers prefer 70°F as a standard temperature, while combustion engineers prefer 77°F.

Before these equations can be used with waste gas streams which are not at 70°F and 1 atmosphere, their variables must be adjusted. As noted above, waste gas streams in air pollution control applications obey the ideal gas law. From this law the following adjustment equation can be derived:

$$Q_2 = Q_1\left(\frac{T_2}{T_1}\right)\left(\frac{P_1}{P_2}\right) \tag{1.14}$$

where

Q_2, Q_1 = gas flow rates at conditions 2 and 1, respectively (actual ft^3/min)

T_2, T_1 = absolute temperatures at conditions 2 and 1, respectively (°R)

P_2, P_1 = absolute pressures at conditions 2 and 1, respectively (atm)

However, according to Equation 1.6:

$$Q = u_t\left(\frac{\pi D_d^2}{4}\right)$$

If Equations 1.6 and 1.14 were combined, we would obtain:

$$u_{t2} = u_{t1}\left(\frac{T_2}{T_1}\right)\left(\frac{P_1}{P_2}\right)\left(\frac{D_{d2}{}^2}{D_{d1}{}^2}\right) \tag{1.15}$$

This last expression can be used to adjust u_t in any equation, as long as the gas flow is in circular ducts.

1.3.2 Hood Design Procedure

1.3.2.1 Hood Design Factors

When designing a hood, several factors must be considered:[28]

- Hood shape
- Volumetric flow rate
- Capture velocity
- Friction

Each of these factors and their interrelationships will be explained in this section.

As discussed in Section 1.2.1, the hood shape is determined by the nature of the source being controlled. This includes such factors as the temperature and composition of the emissions, as well as the dimensions and configuration of the emission stream. Also important are such environmental factors as the velocity and temperature of air currents in the vicinity.

The hood shape partly determines the volumetric flow rate needed to capture the emissions. Because a hood is under negative pressure, air is drawn to it from all directions. Consider the simplest type of hood, a plain open-ended duct. Now, envision an imaginary sphere surrounding the duct opening. The center of this sphere would be at the center of the duct opening, while the sphere radius would be the distance from the end of the duct to the point where emissions are captured. The air would be drawn through this imaginary sphere and into the duct hood. Now, the volume of air drawn through the sphere would be the product of the sphere surface area and the hood capture velocity, u_c:[29]

$$Q = u_c(4\pi x^2) \tag{1.16}$$

where

x = radius of imaginary sphere (ft)

Equation 1.16 applies to a duct whose diameter is small relative to the sphere radius. However, if the duct diameter is larger, the capture area will have to be reduced by the crosssectional area of the duct (D_d), or:

$$Q = u_c\left(4\pi x^2 - \frac{\pi D_{d^2}}{4}\right) \tag{1.17}$$

Similarly, if a flange were installed around the outside of the duct end, the surface area through which the air was drawn—and the volume flow rate—would be cut in half. That occurs because the flange would, in effect, block the flow of air from points behind it. Hence:

$$Q = u_c(2\ \pi x^2) \tag{1.18}$$

From these examples, it should be clear that the hood shape has a direct bearing on the gas flow rate drawn into it. But Equations 1.16 to 1.18 apply only to hoods with spherical flow patterns. For other hoods, other flow patterns apply—cylindrical, planal, etc. We can generalize this relationship between volumetric flow rate and hood design parameters as follows:

$$Q = f(u_i, x, Sh) \tag{1.19}$$

where

"*f(...)*" denotes "function of..."
"*Sh*" indicates hood shape factors
u_i = design velocity—capture, face, slot

Table 1.1 lists design equations for several commonly used hood shapes. As this table shows, Q is a function of x, the hood shape, and, in general, the capture velocity (u_c). In the case of a booth hood, the design velocity is the hood face velocity (u_f). For slotted side-draft and back-draft hoods, the slot velocity (u_s) is the design velocity. In practice, both the hood face and slot velocities are the same, as each measures the speed at which the gas passes through the hood inlet opening(s).

Table 1.1: Design Equations, Loss Factors, and Coefficients of Entry for Selected Hood Types[2]

Hood Type	Design Equation	Less Factor (k_h)	Coefficient of Entry (C_e)
Duct end (round)	$Q = 4\,x^2u_c$	0.93	0.72
Flanged duct end (round)	$Q = 2\,x^2u_c$	0.50	0.82
Free-standing slot hood	$Q = 2\,xLu_c$	1.78	0.55
Slot hood w/ sides, back	$Q = 0.5\,xLu_c$	1.78	N/A
Tapered hood	$Q = 2\,xu_c$	0.06[1]	0.97
Booth hood w/ tapered take-off duct (round)	$Q = u_fA_h$	0.25	0.89
Canopy hood	$Q = 1.4Pxu_c$	0.25	0.89
Canopy hood w/ insert	$Q = 1.4Pxu_c$	1.0	0.71
Dip tank hood (slotted)	$Q = 125A_t$	1.78	N/A
Paint booth hood	$Q = 100A_b$	0.25	N/A

[1] Both k_h and C_e pertain to round ducts and to hoods with a 45° taper. At other angles, k_h and C_e will differ.

N/A = Not applicable
Q = flow rate drawn into hood (ft³/min)
x = distance from hood to source (ft)
u_c = hood capture velocity (ft/min)
u_f = hood face velocity (ft/min)
u_s = hood slot velocity (ft/min)
A_h = hood vace area (ft²)
P = perimeter of source (ft)
L = width of hood slot (ft)
A_t = tank + drainboard surface area (ft²)
A_b = booth cross-sectional area (ft²)

When gas enters a hood, there is mechanical energy loss due to friction. This friction loss is calculated using Equations 1.1 and 1.2, assuming that the potential energy contribution from gravity, $\Delta z\,(g/g_c)$, and the work added to the system, W, are both zero. Thus:

$$vp_2 - vp_1 + \frac{u_2^2}{2g_c} - \frac{u_1^2}{2g_c} = -F \qquad (1.20)$$

Replacing these terms with the corresponding ones from Equations 2.7 and 2.12, we obtain:

$$SP_2 - SP_1 + VP_2 - VP_1 = -H_c = -k_hVP_2 \qquad (1.21)$$

where

SP_i = static gauge pressure at point i (in. w.c.)
VP_i = velocity pressure at point i (in. w.c.)
H_c = hood entry loss (in. w.c.)
k_h = hood loss factor (unitless)

In this equation, subscript 1 refers to a point just outside the hood face. Subscript 2 denotes the point in the duct, just downstream of the hood, where the duct static pressure, SP_2 or SP_h and the duct transport velocity, u_2 or u_t are measured. At point 1, the hood velocity pressure, VP_1, is essentially zero, as the air velocity there is negligible. Moreover, the static gauge pressure, SP_1, will be zero, as the absolute pressure at point 1 is assumed to be at one atmosphere, the reference pressure. After these simplifications are made, Equation 1.21 can be rearranged to solve for the hood loss factor (k_h):

$$k_h - \left(\frac{-SP_h}{VP_2}\right) - 1 \qquad (1.22)$$

At first glance, it appears that k_h could be negative, since VP is always positive. However, as the air entering the hood is under a vacuum created by a fan downstream, SP_h must be negative. Thus, the term "$-SP_h/VP_2$" must be positive. Finally, because the absolute value of SP_h is larger than VP_2, $k_h > 0$.

The hood loss factor varies according to the hood shape. It can range from 0.04 for bell mouth hoods to 1.78 for various slotted hoods. A parameter related to the hood loss factor is the coefficient of entry (c_e).[30] This is defined as:

$$c_e = \left[\frac{1}{(1 + k_h)}\right]^{1/2} \qquad (1.23)$$

c_e depends solely on the shape of the hood, and may be used to compute k_h and related parameters. Values of k_h and c_e are listed in Table 1.1.

Illustration: The static gauge pressure, SP_h, is -1.75 in. w.c. The duct transport velocity (u_t) is 3,500 ft/min. Calculate the loss factor and coefficient of entry for the hood. Assume standard temperature and pressure.

Solution: First, calculate the duct velocity pressure. By rearranging Equation 1.11 and substituting for u_t, we obtain:

$$VP = \left(\frac{u_t}{4{,}016}\right)^2 = \left(\frac{3{,}500}{4{,}016}\right)^2 = 0.76\ in.\ w.c.$$

Next, substitute for VP in Equation 1.22 and solve:

$$k_h = \left(\frac{-SP_h}{VP}\right) - 1 = -\left(\frac{-1.75}{0.76}\right) - 1 = 1.30$$

Finally, use this value and Equation 1.23 to calculate the coefficient of entry:

$$C_e = \left(\frac{1}{1 + 1.30}\right)^{1/2} = 0.66$$

Hood design velocities are listed in Table 1.2. Three kinds of velocities are shown: (1) capture (defined in Section 2.1), (2) face, and (3) slot. As stated in Section 1.2.1, the capture velocity is the air velocity induced by the hood to capture contaminants emitted at some distance from the hood inlet. The face velocity is the average velocity of the air passing through the hood inlet (face). A similar parameter is the slot velocity, which is the average air velocity through the hood slot openings, whose areas is only a fraction of the entire hood face area. Consequently, the slot velocity is usually much higher than the face velocity.[31]

Table 1.2: Hood design Velocities [2]

Operation/Hood Type	Velocity Type	Velocity Range(ft/min)
Tanks, degreasing	Capture	50-100
Drying oven	Face	75-125
Spray booth	Capture	100-200
Canopy hood	Capture	200-500
Grinding, abrasive blasting	Capture	500-2,000
Slot hood	Slot	2,000

Note that these velocities range from 50 to 100 ft/min (tank and degreasing hoods) to 2,000 ft/min, the recommended slot velocity for slotted side-draft/back-draft hoods. As a reference point, the velocity of air in industrial operations due to thermal mixing alone is 50 ft/min. Thus, hood design velocities must exceed this value if effective capture is to occur.[32]

Two other velocities are also discussed in the industrial hygiene literature, although they do not have as much bearing on hood design as the capture, face, or slot velocities. These are the plenum velocity and the transport velocity. Plenum velocity is the velocity of the gas stream as it passes through the tapered portion of a hood (plenum) between the hood opening and the duct connection. This plenum is a transition area between the hood opening and duct. Consequently, the plenum velocity is higher than the hood face velocity, but lower than the duct (transport) velocity. The transport velocity- the gas velocity through the duct- varies according to the waste gas composition. It is a crucial parameter in determining the duct diameter, the static pressure loss, and the sizes of the system fan and fan motor. (For more on transport velocity, see Section 1.3.3.)

1.3.2.2 Hood Sizing Procedure

As with many control devices and auxiliaries, there are several approaches to sizing hoods. Some of these approaches are quite complex, entailing a series of complex calculations that yield correspondingly accurate results. For instance, one hood sizing method in the literature involves first determining the hood dimensions (length and width for rectangular hoods; diameter, for circular). The next step is to estimate the amount of metal plate area (ft^2) required to fabricate a hood of these dimensions, via parametric curves. (No curves are provided for nonmetal hoods.) This plate area is input to an equation that includes a "pricing factor" and the per-pound price of metal. The cost of labor needed to fabricate this hood is estimated from equations similar to the plate-area relationships. Finally, the metal and labor costs are summed to obtain the total fabricated hood cost.[33]

This method no longer yields reasonably accurate hood cost. Since the labor cost data are outdated—1977 vintage—which makes them unescalatable. (The rule-of-thumb time limit for escalating costs is five years.) Even if the costs were up-to-date, the procedure is difficult to use, especially if calculations are made by hand.

A simpler sizing method—yet one sufficiently accurate for study estimating purposes—involves determining a single dimension, the hood face area (A_f). This area, identical to the hood inlet area, can be correlated against the fabricated hood cost to yield a relatively simple cost equation with a single independent variable. To calculate A_f, the following information is needed:

- Hood type
- Distance of the hood face from source (x)
- Capture (u_c), face (u_f), or slot velocity (u_s)
- Source dimensions (for some hood types).

As the equations in Table 1.1 indicate, these same parameters are the ones that are used to determine the volumetric flow rate (Q) through the hood and ductwork. With most control devices and auxiliaries being sized, Q is given. For hoods, however, Q usually must be calculated.

Illustration: A circular canopy hood is being used to capture emissions from a chromium electroplating tank. The hood face is 6 feet above the tank, an 8-foot diameter circular vessel. The capture velocity for this example is 200 ft/min. Assuming that the tank surroundings are at standard conditions, calculate the required volumetric flow rate drawn into the hood, the hood face area, and the hood face velocity.

Solution: Obtain the canopy hood equation from Table 1.1:

$$Q = 1.4 P x u_c \tag{1.24}$$

where

P = perimeter of tank (ft)
x = distance of hood above tank (ft)
u_c = capture velocity (ft/min)

Because the tank is circular, $P = \pi(8\text{ ft}) = 25.1$ ft.

Therefore:

$$Q = (1.4 ft)\ (6 ft)\left(200\ \frac{ft}{min}\right) = 42{,}200\ \frac{ft^3}{min}$$

For this type of canopy hood, the hood diameter is 40% greater than the tank diameter (hence, the "1.4" factor in Equation 1.24). Thus:

$$A_f = \left(\frac{\pi}{4}\right)([1.4]\ [8])^2 = 98.5\ ft^2$$

Finally, the hood face velocity (u_f) would be:

$$u_f = \frac{Q}{A_f} = \frac{42{,}200}{98.5} = 428\ \frac{ft}{min}$$

In this example, note that the hood face velocity is higher than the capture velocity. This is logical, given the fact that the hood inlet area is smaller than the area through which the tank fumes are being drawn. The face velocity for some hoods is even higher. For example, for slotted hoods it is at least 1,000 ft/min.[34] In fact, one vendor sizes the openings in his slotted hoods so as to achieve a slot velocity equal to the duct transport velocity.[35]

1.3.3 Ductwork Design Procedure

The design of ductwork can be an extremely complex undertaking. Determining the number, placement, and dimensions of ductwork components—straight duct, elbows, tees, dampers, etc.—can be tedious and time-consuming. However, for purposes of making study-level control system cost estimates, such involved design procedures are not necessary. Instead, a much simpler ductwork sizing method can be devised.

1.3.3.1 Two Ductwork Design Approaches

There are two commonly used methods for sizing and pricing ductwork. In the first, the total weight of duct is computed from the number and dimensions of the several components. Next, this weight is multiplied by a single price (in $/lb) to obtain the ductwork equipment cost. To determine the ductwork weight, one needs to know the diameter, length, and wall thickness of every component in the system. As stated above, obtaining these data can be a significant effort.

The second method is a variation of the first. In this technique, the ductwork components are sized and priced individually. The straight duct is typically priced as a function of length, diameter, wall thickness, and the material of construction. The elbows, tees, and other fittings are priced according to all of these factors, except for length. Other variables, such as the amount and type of insulation, also affect the price. Because it provides more detail and precision, the second method will be used in this section.

1.3.3.2 Ductwork Design Parameters

Again, the primary ductwork sizing variable are length, diameter, and wall thickness. Another parameter is the amount of insulation required, if any.

Length: The length of ductwork needed with an air pollution control system depends on such factors as the distance of the source from the control device and the number of directional changes required. Without having specific knowledge of the source layout, it is impossible to determine this length accurately. It could range from 20 to 2,000 feet or more. It is best to give the straight duct cost on a $/ft basis and let the reader provide the length. This length must be part of the specifications of the emission source at which the ductwork is installed.

Diameter: As discussed in Section 1.2.2., circular duct is preferred over rectangular, oval, or other duct shapes. For circular ducts, the cross-sectional area, A_d, is given by:

$$A_d = \frac{\pi D_d^{\,2}}{4} \qquad (1.25)$$

where

A_d = cross-sectional area of duct (ft^2)
D_d = duct diameter (ft)

The duct cross-sectional area is the quotient of the volumetric flow rate (Q) and the duct transport velocity (u_t):

$$A_d = \frac{Q}{u_t} \qquad (1.26)$$

Combining Equations 1.25 and 1.26 and solving for D_d:

$$D_d = 1.128 \left(\frac{Q}{u_t}\right)^{1/2} \qquad (1.27)$$

As Q is usually known, the key variable in Equation 1.27 is the duct transport velocity. The value of this variable must be chosen carefully. If the u_t selected is too low, the duct will be oversized and, more importantly, the velocity will not be high enough to convey any particulate matter in the waste gas stream to the control device. However, if u_t is too high, the static pressure drop (which is proportional to the square of u_t) will be excessive, as will be the corresponding fan power consumption.

Cost is also a consideration when determining the optimum duct diameter. The equipment cost increases with increasing duct diameter. However, the fan power cost changes inversely with diameter. Nonetheless, for study-estimating purposes, the optimum duct diameter does not have to be determined. It is sufficient to calculate the duct diameter merely by using the transport velocity values contained in this section.

The transport velocity typically varies from 2,000 to 6,000 ft/min, depending on the waste gas composition. The lower duct velocity would be adequate for a waste gas containing gaseous pollutants or very fine, light dusts, while the higher velocity would be needed to convey a stream

with a large quantity of metals or other heavy or moist materials. The velocities given in Table 1.3 may be used as general guidance[36]:

Table 1.3: Minimum Transport Velocity for Materials Conveyed

Material(s) Conveyed	Minimum Transport Velocity (u_t, ft/min)
Gases: very fine, light dusts	2,000
Fine, dry dusts and powders	3,000
Average industrial dusts	3,500
Coarse dusts	4,000-4,500
Heavy or moist dust loading	≥ 4,500

Table 1.4 supplements these values with recommended duct velocities for a variety of conveyed materials.

Wall thickness: The wall thickness of a duct depends on several factors—internal pressure, diameter, material of fabrication, and other structural parameters. Nonetheless, duct of a given diameter can be fabricated for a range of wall thicknesses, and vice-versa. For instance, 24-in. diameter 304 stainless steel "fully-welded longitudinal seam duct" is fabricated in thicknesses ranging from 22 to 14 gauge (0.0313 to 0.0781 in.). This same range of gauges is used with duct diameters ranging from 3 to 36 in.[37]

Note that the gauge number decreases with increasing wall thickness. This measure, which is traditionally used in the metal fabricating industries, is more convenient to deal with than the thickness expressed in inches, as the latter are usually small numbers less than 0.25. Moreover, the gauge number varies according to the metal used—carbon steel (galvanized or nongalvanized), stainless steel, or aluminum. Gauges for these metals are given in Table 1.5 for a wide range of nominal thicknesses. The gauge measure is not used with plastic duct, as the wall thickness is typically expressed in inches. In any event, the wall thickness usually does not need to be known to estimate duct cost, as this parameter is already accounted for in the cost equations. (See Section 1.4.)

Insulation: As discussed in Section 1.2.2., insulation can be either installed on the outer surface of ductwork or the ductwork itself can be fabricated with built-in insulation. In the first case, the amount of insulation required will depend on several heat transfer variables, such as: the temperature, velocity, composition, and other properties of the waste gas; the ambient temperature; the duct diameter, wall thickness, and thermal conductivity; and the desired surface ("skin") temperature. Determining these variables involves making a series of complex calculations that, while well-established, are beyond the scope of this chapter. Standard references as Perry's Chemical Engineers' Handbook and Plant Design and Economics for Chemical Engineers present these calculations, as do heat transfer bibliographies.[38, 39]

Table 1.4: Minimum Duct Velocities for Selected Materials[2]

Material	Minimum Transport Velocity (ft/min)
Aluminum dust (coarse)	4,000
Brass turnings	4,000
Cast iron boring dust	4,000
Clay dust	3,500
Coal dust (powdered)	4,000
Cocoa dust	3,000
Flour dust	3,000 - 5,000[1]
grain dust	2,500 - 3,000
Lead dust	4,000
Limestone dust	3,500
Magnesium dust (coarse)	3,000
Metal turnings	4,000-5,000
Plastics dust (buffing)	3,000
Rubber dust	2,500 (fine) - 4,000 (coarse)
Silica dust	3,500 - 4,500
Soap dust	3,000
Soapstone dust	3,000
Spray paint	2,000
Starch dust	3,000
Stone dust	3,500
Tobacco dust	3,500

[1] Transport velocity varies with foundry operation.

The second approach is to select pre-insulated ductwork. As mentioned previously, it can be equipped with any type and thickness of insulation. However, 1, 2, or 3 inches is typical. (Prices for these are presented in Section 1.4.)

1.3.3.3 Ductwork Pressure Drop

As mentioned in Section 1.3.1, ventilation system energy losses due to friction are traditionally computed as fractions of the velocity pressure, *VP*. In most cases, Equation 1.12 can be used to estimate these losses. Technically, though, these equations apply only to those regions in the ventilation system where there are no changes in the velocity pressure (i.e., where the duct diameter is constant). These regions would include straight duct, hoods, and such fittings as couplings

Table 1.5: Wall Thickness of Steel and Aluminum Duct [40]

Gauge Number	Nominal Thickness (inches)			
	Carbon Steel		Stainless Steel (304 or 316)	Aluminum 3003-H14[3]
	Galv[1]	Nongalv[2]		
28	0.0187	0.0149	0.0156	0.025
26	0.0217	0.0179	0.0188	0.032
24	0.0276	0.0239	0.0250	0.040
22	0.0336	0.0299	0.0313	0.050
20	0.0396	0.0359	0.0375	0.063
18	0.0516	0.0478	0.0500	0.080
16	0.0635	0.0598	0.0625	0.090
14	0.0785	0.0747	0.0781	–
12	0.1084	0.1046	0.1094	–
10	0.1382	0.1345	0.1406	–

[1] To provide equivalent strength and stiffness, the nominal thickness of aluminum is approximately 150% of the nominal thickness of galvanized carbon steel of the same gauge.
[2] Galvanized and paintable galvanized carbon steel.
[3] Nongalvanized carbon steel.

and simple elbows. But, with tees, wyes, and other divided flow fittings, the velocity—and velocity pressure—are not constant between the fitting inlet and outlet. The corresponding friction loss (F_b) is a function of both the upstream (inlet) and branch *VP*s, as the following equation indicates:[40]

$$F_b = VP_u (k_b - 1) + VP_b \qquad (1.28)$$

where

VP_u, VP_b = upstream and branch velocity pressures, respectively (in. w.c.)
k_b = branch loss coefficient

However, divided flow fittings generally are not used with simple pollution control ventilation systems, except in those cases where a tee fitting might be needed, say, for purposes of adding dilution air.[1]

As any fluid mechanics textbook would attest, the friction loss for ductwork is a complex function of several variables: duct diameter and length, transport velocity, and gas viscosity and density. Specifically, the Darcy-Weisbach and Colebrook Equations are typically used to make

[1] Divided flow fittings are needed with more complex control systems that collect waste gases from several emission points. The design of such ventilation systems is beyond the scope of this section, however.

this calculation, the latter being used to compute the Reynolds number.[41] Traditionally, the friction loss has been obtained from a nomograph or, more recently computer programs. A typical nomograph is found in Burton.[42] Also, to simplify the calculation, empirical equations have been derived for certain kinds of commercially-available ductwork. For instance, to estimate the friction loss per 100 ft (F_d/100 ft) at standard conditions for round, spiral, galvanized ductwork having 10 joints per 100 ft, use the following equation:[43]

$$\frac{F_d}{100\,ft} = 0.136\left(\frac{1}{D_d}\right)^{1.18}\left(\frac{u_t}{1{,}000}\right)^{1.8} \qquad (1.29)$$

where

D_d = duct diameter (ft), and: $0.25 \le D_d \le 5$

Clearly, this equation provides the total friction loss, not the loss factor (k). However, the reader may compute k for a given diameter (D_d) and flow rate (Q) by simply dividing the Equation 1.29 results by VP and multiplying by 100.

To estimate the friction loss for other duct materials, multiply the value from Equation 1.29 by a roughness correction factor, approximate values of which are:[44]

Table 1.6: Roughness Correction Factors for Various Duct Materials

Material	Roughness Correction Factor
Non-spiral-wound galvanized	0.9
Fiberglass (smooth finish)	0.8
ABS and PVC plastic	0.8
Concrete	1.4
Corrugated flex duct	2.3

Loss factors for fittings have also been compiled, based on experimental data. Mainly of interest are those for 90° elbows, arguably the most commonly used fitting in air pollution control systems. The "k_{90}" values for elbows vary according to the diameter and radius of curvature, which is expressed as a multiple of the elbow diameter. Typical ranges of these values are as follows:[45]

Table 1.7: Typical Ranges of Friction Loss Factor (k_{90})

Radius of Curvature	Friction Loss Factor (k_{90})
0.50	0.80
1.00	0.35
1.25	0.30 - 0.55
1.50	0.27 - 0.39
2.00	0.24 - 0.27
2.50	0.22 - 0.24

As these values indicate, the higher the radius of curvature, the lower the friction loss. This stands to reason, as the higher the radius of curvature, the more gradually the gas stream changes direction. For an elbow having of angle less than 90°, multiply the above k_{90} value by an adjustment factor ($\theta/90$), so that:

$$k_\theta = \left(\frac{\theta}{90}\right) k_{90} \qquad (1.30)$$

where

k_θ = loss factor for $\theta > 90°$

Illustration: A control device at a cosmetic factory is connected to a source by 250 feet of round spiral duct. The duct run includes three 90° elbows and two 45° elbows, each with a 1.50 radius of curvature. The volumetric flow rate (Q) of the waste gas (which contains entrained face powder) is 15,000 ft^3/min at standard conditions. Calculate the friction loss for the ductwork.

Solution: Because the material being conveyed in the ductwork (face powder) is light, an appropriate transport velocity (u_t) in this case is 2,000 ft/min. (See Table 1.4 above.) Upon substituting this value and the volumetric flow rate into Equation 1.27 we obtain the duct diameter (D_d):

$$D_d = 1.128 \left(\frac{15{,}000}{2{,}000}\right)^{1/2} = 3.09\ ft$$

Next, substitute the diameter and velocity into Equation 1.29 to compute the straight duct friction (static pressure) loss, F_d:

$$F_d = 0.136 \left(\frac{1}{3.09}\right)^{1.18} \left(\frac{2{,}000}{1{,}000}\right)^{1.8} \left(\frac{250}{100}\right) = 0.313\ in.\ w.c.$$

The 250/100 factor in this expression adjusts the friction loss from 100 feet (the basis of equation 10.29) to 250 feet (the length of the duct system in this illustration). The rest of the friction loss occurs through the five elbows (three 90°, two 45°), each with a 1.50 radius of curvature. These losses (F_c) are computed via Equation 1.12:

$$F_c = k_\theta VP \tag{1.31}$$

where

VP = $(2{,}000/4{,}016)^2$ (Equation 1.11, rearranged)
= 0.248 in. w.c.

For the 90° elbows, $k_\theta = k_{90} = 0.33$ (average of table range), and:

$$F_c = 3 \times 0.33\ (0.248) = 0.246 \text{ in. w.c.}$$

For the 45° elbows, $k_\theta = (45/90)k_{90} = 0.165$ (Equation 1.30), and:

$$F_c = 2 \times 0.165(0.248) = 0.0818 \text{ in. w.c.}$$

The total friction loss is, therefore:

$$F = 0.313 + 0.246 + 0.0818 = 0.641 \text{ in. w.c.}$$

From this illustration, two observations may be made: (1) the static pressure loss through the straight duct is not large, even at this length (250 ft.) and (2) the losses through the elbows—which total 0.328 in. w.c.—are larger than the straight duct loss. Though it may be tempting to neglect fittings losses for the sake of expediency, doing so can cause a significant underestimation of the ventilation system static pressure loss.

1.3.4 Stack Design Procedures

As with ductwork, the design of stacks involves a number of stream, structural, and site-specific parameters.[46,47] These include:

Waste gas variables: inlet volumetric flow rate, temperature, and composition;

Site-specific data: elevation above sea level, ambient temperature fluctuations, topographic and seismic data, meteorological records, and building elevations and layout;

Structural parameters: thickness of stack wall and liner, location of breaching opening, type of supports, load capacity of foundation, modulus of resistance, and natural vibration frequency.

Fortunately, for study cost-estimating purposes, the only two stack design parameters that need to be determined are: (1) the stack diameter and (2) the stack height. The other variables (e.g., wall thickness) are incorporated into the equipment cost correlations. The stack diameter is relatively easy to determine, as it depends primarily on waste stream conditions. The stack height is more difficult to arrive at, as it is influenced by several site-specific variables. Nonetheless, ample guidance has been developed to allow the estimator to determine an acceptably accurate stack height.

1.3.4.1 Calculating Stack Diameter

Because most stacks have circular cross-sections, the stack diameter (D_s, ft) can be calculated via the duct diameter formula (Equation 1.27):

$$D_s = 1.128 \left(\frac{Q_c}{u_c} \right)^{1/2} \tag{1.32}$$

where

u_c = stack exit velocity (ft/min)
Q_c = exit volumetric flow rate (actual ft^3/min)

It should be noted that the stack diameter in this formula is measured at the stack exit, not at the entrance. That is because, for structural reasons, the diameter at the bottom of the stack typically is larger than the top diameter. Also note that the stack exit velocity does not necessarily equal the duct transport velocity. Finally, Q_c may be different from the volumetric flow rate used to size the ductwork. Because the stack always follows the control device, the flow rate entering the device may not equal the flow rate entering the stack, either in standard or actual ft^3 /min terms. For instance, in a thermal incinerator, the outlet standard waste gas flow rate is almost always higher than the inlet flow rate due to the addition of supplemental fuel.

The stack exit velocity, u_c, affects the plume height, the distance that the plume rises above the top of the stack once it exits. In a well-designed stack, u_c should be 1.5 times the wind speed. Typically, design exit velocities of 3,000 to 4,000 ft/min are adequate.[48] This range corresponds to wind speeds of 34 to 45 mi/hr.

1.3.4.2 Calculating Stack Height

Estimating the stack height is more difficult than calculating the stack exit diameter. The stack height depends on several variables: the height of the source; the stack exit velocity; the stack and ambient temperatures; the height, shape, and arrangement of the nearby structures and terrain; and the composition of the stack outlet gas. Some of these variables are straightforward to determine, while others (such as the dimensions and layout of nearby structures) are difficult to determine without performing on-site modeling and monitoring studies.

The stack design height has two components: the height of the stack itself (H_s) and the plume rise height (H_{pr}). Together these components comprise the effective stack height (H_e). That is:

$$H_e = H_s + H_{pr} \tag{1.33}$$

However, the cost of the stack is a function of H_s alone. (See Section 1.4.) As discussed above, the plume rise is a function of the stack exit velocity. It also depends on the temperature differential between the stack gas and the ambient air. Specifically, a 1°F temperature difference corresponds to approximately a 2.5-ft. increase in H_{pr}.[49]

For those sources subject to State Implementation Plans (SIPs), the stack height (H_s) should be determined according to "good engineering practice" (GEP). GEP is defined as "the height necessary to insure that emissions from the stack do not result in excessive concentrations of any air pollutant in the immediate vicinity of the source as a result of atmospheric downwash, eddies, or wakes which may be created by the source itself, nearby structures, or nearby terrain obstacles."[50] In this respect, GEP establishes the maximum allowable stack height credit for purposes of calculating the ambient air quality impact of the emitting source. A source may build a stack to any height, but only a certain amount of stack height will be allowed in determining environmental impacts.[51]

For stacks constructed after January 12, 1979, the GEP stack height shall be the greater of: (1) 65 meters (213 ft); (2) the height demonstrated by an approved fluid model or field study that ensures that stack emissions do not cause excessive pollutant concentrations from atmospheric downwash, wakes, eddy effects., etc; or (3) the height determined by the following equation:[52]

$$H_s = H_b + 1.5L \tag{1.34}$$

where

H_s = GEP stack height, measured from the ground level elevation at the stack base (ft)

H_b = height of nearby structure(s) measured from this ground level elevation (ft)

L = lesser dimension (height or projected width of nearby structure(s))

1.3.4.3 Calculating Stack Draft

As discussed previously, waste gas flowing through hoods and ductwork loses static pressure due to friction. In the case of stacks, however, the gas stream can actually gain static pressure, as a result of stack draft, which is the draft created by the stack gas-ambient air temperature differential. Stack draft (SP_s, in. w.c.) can be calculated as follows:[53]

$$SP_s = 0.034\,(H_s - H_{br})\,\rho \times \left(\frac{1}{T_{amb}} - \frac{1}{T_{sa}}\right) \qquad (1.35)$$

where

H_{br} = height of stack breaching (inlet duct connection) above stack base (ft)

P = barometric pressure(in. w.c.)

T_{amb} = ambient temperature (°R)

T_{sa} = average stack gas temperature (°R)

Illustration: The waste gas from a thermal incinerator has an outlet flow rate and temperature of 21,700 actual ft^3/min. and 550°F, respectively. The maximum wind speed in the vicinity is 42 mi/hr, while the stack exit and ambient temperatures are 450°F and 70°F, in turn. The barometric pressure is 1 atm. (29.92 in. Hg). The incinerator is near a 35-ft tall brick building, while the "projected width" of an adjacent building is 40 ft. For a stack to disperse the incinerator offgas, calculate the required: (1) exit velocity, (2) diameter, (3) height, and (4) draft.

Solution:

Exit velocity: According to the above guideline, the velocity should be 1.5 times the wind speed, or:

$$u_c = 1.5 \times 42\ mph \times 88\ \frac{fpm}{mph} = 5{,}540\ \frac{ft}{min} \qquad (1.36)$$

Stack diameter: The exit volumetric flow rate is measured at the stack exit temperature, namely 450°F. However, the above flow rate was measured at 550°F, the incinerator outlet temperature. Correcting to the stack exit temperature, we obtain:

$$Q_c = \frac{21{,}700 \times (450 + 460)}{(550 + 460)} = 19{,}600 \ \frac{actual\ ft^3}{min} \qquad (1.37)$$

Substituting this value into Equation 1.32:

$$D_s = 1.128 \left(\frac{19{,}600}{5{,}540}\right)^{1/2} = 2.12\ ft$$

Stack height: As a first approximation, estimate the GEP stack height from Equation 1.33, where the variables H_b and L are 35 ft and 40 ft, respectively:

$$H_s = 35 + 1.5(40) = 95 \text{ ft.}$$

Clearly, this H_s is less than the GEP maximum height (213 ft), so it will be used in this example.

Stack draft: All of the inputs needed to compute the stack draft via Equation 1.35 are known except the stack breaching height, H_{br}. However, a minimum of 5 ft is recommended for this parameter.[54] This value will be used in this calculation. Also, the average stack temperature is:

$$T_{sa} = \frac{(450 + 550)}{2 + 460} = 960^\circ R \qquad (1.38)$$

Finally, the barometric pressure expressed in inches of water is:

$$\rho = 29.92\ in.\ Hg \times \frac{13.6\ in.\ water}{in.\ Hg} = 407\ in.\ w.c. \qquad (1.39)$$

Upon substitution, we obtain:

$$SP_s = (0.034)\ (118 - 5)\ (407) \left[\frac{1}{70 + 460} - \frac{1}{960}\right] = 1.32\ in.\ w.c.$$

1.4 Estimating Total Capital Investment

This section presents the information needed for estimating the total capital investment (TCI) for hoods, ductwork, and stacks. The TCI includes the equipment cost (EC) for the hood, ductwork, or stack; taxes; freight charges; instrumentation (if applicable); and direct and installation costs. All costs are presented in second quarter 1993 dollars, and are of "study" estimate accuracy (± 30 percent).[2] Moreover, the costs are for new facility installations; no retrofit costs are included.

The equipment costs are presented in Section 1.4.1, while the installation costs are shown in Section 1.4.2. In each of these sections, the three categories of equipment are covered in separate subsections.

1.4.1 Equipment Costs

Several vendors provided costs (prices) for each of the three equipment categories. Their responses reflected a range of sizes, designs, and materials of construction. These prices have been correlated against some easy-to-determine design (sizing) parameter via least-squares regression analysis. Each of these correlations pertains to a certain type of equipment (e.g., circular canopy hoods) within a specified size range of the parameter in question (e.g., 2 to 200 ft^2 inlet area). For that reason, a cost correlation should not be extrapolated outside the parameter range specified.

Some of the prices the vendors provided pertain to stock ("off-the-shelf") items, while other costs are for custom-fabricated equipment. Vendors tend to specialize in either stock or custom items. Most hoods and stacks are custom-made, either fabricated in the vendor's factory or erected on-site. Conversely, ductwork components usually are stock items, though larger pieces have to be custom-made. (Of course, there are exceptions to this.) Finally, all prices given in the following section are "free-on-board (F.O.B.) vendor," meaning that they include neither freight nor taxes.

1.4.1.1 Hood Costs

In all, four vendors provided prices for hoods.[55] These prices covered the following types of hoods:

- Canopy—circular
- Canopy—rectangular

[2] For information on escalating these prices to more current dollars, refer to the EPA report *Escalation Indexes for Air Pollution Control Costs* and updates thereto, all of which are installed on the OAQPS Technology Transfer Network (CTC Bulletin Board).

- Push-pull
- Slide-draft
- Back-draft (slotted)

Descriptions and design procedures for these hoods are given in Sections 1.2.1 and 1.3.2, respectively. As explained in Section 1.3.2, hood costs have been found to correlate well with the hood inlet or face area (A_f, ft^2). Furthermore, the functional form that best fits the cost-face area correlation (equation) is the "power function," or:

$$C_h = aA_f^{\,b} \tag{1.40}$$

where

C_h = hood cost ($)
a,b = equation regression parameters

The values of the equation parameters vary according to hood type and material of construction. These parameters are shown in Table 1.8.

Illustration: What would be the cost of the electroplating tank canopy hood sized for the illustration in Section 1.2.2. Assume that the hood is fabricated of FRP.

Solution: Recall that the face area (A_f) calculated for that hood was 98.5 ft^2. Because this is a circular canopy hood, the equation parameters from Table 1.8 are: $a = 123$ and $b = 0.575$. (Note that this hood area falls within the equation range of 2 to 200 ft^2.) Substituting these parameters into Equation 1.40, we obtain:

$$C_h = 123(98.5)^{0.575} = \$1{,}720.$$

1.4.1.2 Ductwork Costs

Several vendors provided ductwork prices, also for a range of sizes, materials, and designs. These prices covered the following equipment items:

Straight ductwork:

Circular
- Steel sheet (galvanized carbon, w/ & w/o insulation; 304 stainless;)
- Steel plate (coated carbon; 304 stainless)
- Plastic (FRP; PVC)

Square
- Steel (aluminized carbon; w/ & w/o insulation)

Table 1.8: Parameters for Hood Cost Equation[55]

Type of Hood	Fabrication Material	Equation Parameter a	b	Equation Range (A_f, ft^2)
Canopy-circular	FRP[1]	128	0.577	2-200
Canopy-rectangular	FRP	306	0.506	2-200
Push-pull	FRP	620	0.321	2-200
Side-draft	FRP	497	0.336	2-200
Backdraft (slotted)	PVC[2]	307	1.43	0.6-2.0[3]
Backdraft (slotted)	PVC[4]	797	0.504	1.1-2.1
Backdraft (slotted)	PP[5]	653	0.717	1.1-2.1
Backdraft (slotted)	FRP	956	0.519	1.1-2.1
Backdraft (slotted)	Galvanized steel	688	0.687	0.5-1.3

[1] Fiberglass-reinforced plastic
[2] Polyvinyl chloride
[3] For each slotted hood, ìequation rangeî denotes the range in the area of the slot openings, which is much less than the total hood face area
[4] Each hood is equipped with manual slot dampers and four rows of slots
[5] Polypropylene

Elbows (90°):

Steel (galvanized carbon, w/ & w/o insulation; 304 stainless)
Plastic (FRP; PVC)

Dampers:

Butterfly
- Steel (galvanized carbon, w/ & w/o insulation)
- Plastic (FRP; PVC, w/ & w/o actuators)

Louvered
- Steel (aluminized carbon w/ & w/o actuators)

Blast gate
- Steel (carbon)
- PVC

These prices were regressed against the diameter of the equipment item (straight duct, elbow, or damper). The regression correlations were of three forms: power function (primarily), exponential, and linear. Equation 1.40 depicts the power function, while the other forms are:

$$(Exponential) \qquad C_i = ae^{bD} \tag{1.41}$$

$$(Linear) \qquad C_i = a + bD \tag{1.42}$$

where

C_i = cost of equipment item in question
a,b = regression parameters
D = equipment diameter

The regression parameters are listed in Tables 1.9 to 1.11, along with the size applicability ranges for the respective correlations. (Note: The correlations should not be extrapolated outside these ranges.) The following paragraphs contain additional information about the price data and the correlations:

Straight duct: As indicated above, vendors provided prices for steel plate, steel sheet (spiral-wound and longitudinal seam), and plastic straight duct. The major difference between the two steel duct types lies in the wall thickness. Steel plate duct typically has wall thicknesses of 3/16 in. to 1/2 in., while steel sheet duct wall thicknesses usually range from 28 gauge to 10 gauge. As Table 1.5 shows, this range corresponds to wall thicknesses of 0.0149 in. to 0.1406 in., respectively, although the exact thicknesses will vary with the type of steel used (e.g., carbon vs. stainless). Also, as discussed in Section 1.3.3.2, each duct diameter can be fabricated with a range of wall thicknesses.

Most of the steel duct vendors supplied prices for a minimum and a maximum wall thickness for a given diameter. However, to simplify matters for cost estimators, these "low" and "high" prices first were averaged, and then the average prices were regressed against the diameters. This averaging was necessary, because those making study cost estimates usually do not have enough information available to predict duct wall thicknesses.

Prices for both circular and square insulated steel sheet duct were among the data received. The insulated circular steel duct is "double-wall, spiral-wound" in construction, wherein the insulation is installed between the inner and outer walls. Costs were provided for both 1-in. and 3-in. fiberglass insulation thicknesses. For the square duct, prices were given for a 4-in. thickness of mineral wool insulation applied to the outer surface of the duct. The correlation parameters in Table 1.9 reflect these specifications.

Prices for both carbon steel (galvanized, painted, or aluminized) and 304 stainless steel duct were received. The carbon steel duct is used in situations where "mild" steel is suitable, while the stainless steel duct is required whenever the gas stream contains high concentrations of corrosive substances.

Vendors gave prices for plastic (FRP and PVC) duct also (Table 1.11). However, for a given diameter this duct is fabricated in a single wall thickness, which varies from approximately 1/8 in. to 1/4 in. Consequently, the estimator is not required to select a wall thickness when costing plastic duct.

Elbows: Prices for steel sheet and plastic 90° elbows were also submitted. The steel sheet elbows were "gored" (sectioned) elbows fabricated from five pieces of sheet metal welded together. Like

Dampers: Prices were obtained for three types of dampers: butterfly, louvered, and blast gates. The galvanized carbon steel butterfly dampers were priced with and without 1-in. fiberglass insulation, while prices for the aluminized carbon steel louvered dampers were based on either manual or automatic control (via electric actuators). Similarly, the PVC butterfly dampers were manual or equipped with pneumatic actuators. Both the carbon steel and the PVC blast gates were manual. Correlation parameters for the steel and plastic dampers are shown in Tables 1.10 and 1.11, in turn.

Illustration: A fabric filter handling 16,500 ft^3/min of 200°F waste gas laden with noncorrosive cocoa dust is located 95 ft across from and 20 ft above, the emission source (a drying oven). Straight duct with four 90° elbows (all fabricated from spiral-wound, galvanized carbon steel sheet) and a butterfly damper (also galvanized CS) will be required to convey the gas from the source to the control device. Assume that the ductwork contains 1-in. thick insulation to prevent condensation. Estimate the cost of these items.

Solution: First, determine the diameter of the straight duct, elbows, and damper. From Table 1.4, the minimum transport velocity (u_t) for cocoa dust is 3,000 ft/min. Substituting this value and the gas volumetric flow rate into Equation 2.27, we obtain:

$$D_d = 1.128\left(\frac{16{,}500}{3{,}000}\right)^{1/2} = 2.65\,ft = 31.7\,in.$$

Next, obtain the costs of the ductwork items as follows:

Straight ductwork: From Table 1.9, select the equation parameters for galvanized circular spiral-wound duct (1-in. insulation) and substitute them and the diameter into the appropriate equation type (power function, Equation 1.40).

Table 1.9: Parameters for Straight Steel Ductwork Cost Equations[56]

Duct Type	Material	Instulation Thickness	Equation Type	Equation Parameter *a*	Equation Parameter *b*	Equation Range (*D*, in.)
Circular-spiral[1]	Sheet-gal CS[2]	None	Power Function	0.322	1.22	3-84
Circular-spiral	Sheet-304 SS[3]	None	Power Function	1.56	1.00	3-84
Circular-spiral	Sheet-galv CS	1	Power Function	1.55	0.936	3-82
Circular-spiral	Sheet-galv CS	3	Power Function	2.56	0.937	3-82
Circular-longitudinal[4]	Sheet-galv CS	None	Power Function	2.03	0.784	6-84
Circular-longitudinal	Sheet-304 SS	None	Power Function	2.98	0.930	6-84
Circular-longitudinal	Plate-coat CS[5]	None	Power Function	2.49	1.15	6-84
Circular-longitudinal	Plate-304 SS[6]	None	Power Function	6.29	1.23	6-84
Square	Sheet-alum CS[7]	None	Linear	0.254	2.21	18-48
Square	Sheet-alum CS	4	Linear	21.1	5.81	18-48

[1] Spiral-wound and welded circular duct
[2] Galvanized carbon steel sheet
[3] 304 stainless steel sheet
[4] Circular duct welded along the longitudinal seam
[5] Carbon steel plate with one coat of ìshop paintî
[6] 304 stainless steel plate
[7] Aluminized carbon steel sheet

However, a total of 115 ft (95 + 20) of duct is required, so:

$$Straight\ duct\ cost = \frac{\$39.4}{ft} \times 115\,ft = \$4{,}531$$

Elbows: From Table 1.10 correlation parameters for galvanized carbon steel, insulated elbows are 53.4 (a) and 0.0633 (b). However, the regression correlation form is exponential (Equation 1.41). Thus:

$$Elbow\ cost\ (\$) = 53.5\, e^{0.0633\,(31.7)} = \$397\ ea.$$

For four elbows, the cost is: $397 x 4 = $1,588.

Table 1.10: Parameters for Plastic Ductwork Cost Equations[56]

Ductwork	Material	Equation Type	Equation Parameter		Equation Range (*D*, in.)
			a	*b*	
Elbows[1]	Galv CS[2]	Exponential	30.4	0.0594	6-84
Elbows	304 SS	Exponential	74.2	0.0668	6-60
Elbows-insulated[3]	Galv CS	Exponential	53.4	0.0633	3-78
Elbows-butterfly[4]	Galv CS	Exponential	23.0	0.0567	4-40
Dampers-butterfly/insulated[5]	Galv CS	Exponential	45.5	0.0597	4-40
Dampers-louvered[6]	Alum CS[7]	Power Function	78.4	0.860	18-48
Dampers-louvered w/ actuators[8]	Alum CS	Power Function	208.	0.791	18-48
Dampers blast gates	Carbon Steel	Power	17.2	0.825	3-18

[1] Single-wall ìgoredî 90° elbows, uninsulated
[2] Galvanized carbon steel sheet
[3] Double-wall ìgoredî 90° elbows with 1-inch fiberglass insulation
[4] Single-wall ìopposed bladeî type manual butterfly dampers
[5] Double-wall ìopposed bladeî butterfly dampers with 1-inch fiberglass insulation
[6] Louvered dampers with 95-98% sealing
[7] ìAluminizedî carbon steel sheet
[8] Louvered dampers with electric actuators (automatic controls)

Damper: Also from Table 1.10, select the correlation parameters for galvanized carbon steel "dampers-butterfly/insulated" and substitute into Equation 1.41:

$$Damper\ cost\ (\$) = 45.5\, e^{0.0597\ (31.7)} = \$302$$

After summing the above three costs, we obtain:

$$Total\ ductwork\ cost = \$6{,}421 \gg \$6{,}420$$

1.4.1.3 Stack Costs

Prices for steel and PVC short stacks were obtained from four vendors.[57] The steel stack costs were for those fabricated from carbon and 304 stainless steels, both plate and sheet metal. As with ductwork, the difference between steel sheet and plate lies in the thickness. For these stacks, the sheet steel thickness ranged from 18 to 16 gauge (0.05 to 0.06 in., approximately). Steel plate thicknesses were considerably higher: 0.25 to 0.75 in, a fact that makes them more

resistant to wind and other loadings than stacks fabricated of steel sheet. This is especially true for taller stacks. The major drawback is that plate steel stacks are more costly than those fabricated from steel sheet.

Another feature that increases costs is insulation. As the correlation parameters show (Table 1.12), insulated stacks cost as much as three times more per foot than uninsulated. With or without insulation, a typical short (15-ft) steel stack consists of the following components:[58]

- Longitudinal seam duct (12-ft section)
- Reducer fitting (3-ft)
- Drip pan
- Support plate (1/4-in, welded to stack)
- Rectangular tap (for connecting to fan discharge)
- Ring (for attaching guy wires)

Table 1.11: Parameters for Plastic Ductwork Cost Equations[56]

Duct Type	Material	Equation Type	Equation Parameter		Equation Range (*D*, in.)
			a	*b*	
Straight duct	PVC[1]	Power Function	0.547	1.37	6-48
Straight duct	FRP[2]	Exponential	11.8	0.0542	4-60
Elbows-90°	PVC	Power Function	3.02	1.49	6-48
Elbows-90°	FRP	Exponential	34.9	0.0841	4-36
Dampers-butterfly	PVC	Power Function	10.6	1.25	4-48
Dampers-butterfly	FRP	Power Function	35.9	0.708	4-36
Dampers-butterfly w/ actuators[3]	PVC	Exponential	299.	0.0439	4-48
Dampers-blast gate	PVC	Power Function	8.14	1.10	4-48

[1] Polyvinyl chloride
[2] Fiberglass-reinforced plastic
[3] Butterfly dampers with pneumatic actuators (automatic controls). all other dampers listed in this table are manually controlled.

Taller stacks may require additional components, such as ladders and platforms, guy wires or other supports, and aircraft warning lights. (See Section 1.2.3.) Table 1.12 lists the parameters and applicable ranges of the stack cost correlations. The correlations cover short PVC stacks, and taller stacks fabricated from plate steel (carbon and 304 stainless types) and sheet steel (insulated and uninsulated). Except for three double-wall sheet steel designs, these stacks are of single-wall construction. Note that all of the correlations are power functions. Also note that the equations apply to various ranges of stack height. In all but one of these equations the cost is expressed in $/ft of stack height. The exception is the cost equation for insulated carbon steel sheet stacks of heights ranging from 30 to 75 feet. In this equation the cost is expressed in $.

This last cost equation is different in another respect. The other six equations in Table 1.12 correlate stack cost ($/ft) with stack diameter (D_s, in.). However, this seventh equation correlates stack cost with stack surface area (S_s, ft^2), a variable that incorporates both the stack diameter and the stack height ($H_{s,ft}$.). The surface area is calculated via the following equation:

$$S_s = \left(\frac{\pi}{12}\right) D_s H_s \tag{1.43}$$

Table 1.12: Parameters for Stack Cost Equations[54]

Material	Equation Parameters[1]		Equation Range	
	a	*b*	D_s(in)[2]	H^s (ft)[3]
PVC[4]	0.393	1.61	12-36	≤10
Plat-coated CS[5]	3.74	1.16	6-84	20-100
Plate-304 SS[6]	12.0	1.20	6-84	20-100
Sheet-galv CS[7]	2.41	1.15	8-36	≤75
Sheet-304 SS[8]	4.90	1.18	8-36	≤75
Sheet-insul CS/DW[9]	143.	0.402	18-48	≤15
Sheet-uninsul CS/DW[10]	10.0	1.03	18-48	≤15
Sheet-insul CS/DW[11]	142.	0.794	24-48	30-75

[1] All cost equations are power functions. (See Equation 1.40) Except where noted, costs are expressed in terms of $/ft of stack height.
[2] Stack diameter range to which each equation applies.
[3] Stack height range to which each equation applies.
[4] Polyvinyl chloride
[5] Carbon steel plate with one coat of ìshop paint.î
[6] 304 stainless steel plate
[7] Galvanized carbon steel sheet
[8] 304 stainless steel sheet
[9] Aluminized carbon steel sheet covered with 4 inches of fiberglass insulation (double-wall construction).
[10] Uninsulated aluminized carbon steel sheet (double-wall construction).
[11] Costs for these stacks are expressed in $, and are correlated with the stack surface area. (S_s, ft^2).

where

1/12 = stack diameter (D_s) conversion factor

Illustration: Estimate the cost of the stack sized in Section 1.3.4.3.

Solution: Recall that the stack dimensions were: $H_s = 95$ ft and $D_s = 2.12$ ft = 25.4 in. Both dimensions fall within the ranges of the cost correlations for steel plate stacks. Because the previous illustration did not indicate whether the waste gas was corrosive, we will estimate the prices for both carbon steel and 304 stainless steel plate stacks.

Upon substituting the equation parameters and stack dimensions into Equation 1.40, we obtain:

$$Price\ (carbon\ steel) = 3.74\ (25.4)^{1.16} \left(\frac{\$}{ft}\right) \times 95\,ft = \$15{,}100$$

$$Price\ (304\ stainless) = 12.0\ (25.4)^{1.20} \left(\frac{\$}{ft}\right) \times 95\,ft = \$55{,}300$$

Notice that the price of the stainless steel stack is nearly four times that of the carbon steel stack. In view of this difference, the estimator needs to obtain more information on the waste gas stream properties, so that he/she can select the most suitable stack fabrication material. Clearly, it would be a poor use of funds to install a stainless steel stack where one is not needed.

1.4.2 Taxes, Freight, and Instrumentation Costs

Taxes (sales, etc.) and freight charges apply to hoods, ductwork, and stacks, as they do to the control devices that these auxiliaries support. As discussed in Section 1, Chapter 2, these costs vary, respectively, according to the location of the ventilation system and the site's distance from the vendor. Typical values are 3% (taxes) and 5% (freight) of the total equipment cost.

Unlike the control devices, ventilation systems generally are not instrumented. The exception would be an electric or pneumatic actuator for a butterfly or louvered damper. In such a case, however, the cost of the instrument (actuator and auxiliaries) would be included in the damper price. Thus, no supplementary instrumentation cost is included.

1.4.3 Purchased Equipment Cost

With ventilation systems, the purchased equipment cost (PEC_t) is the sum of the equipment, taxes, and freight costs. Incorporating the typical values listed in Section 1.4.2, we obtain:

$$PEC_t = EC_t + 0.03EC_t + 0.05EC_t = 1.08\ (EC_t) \qquad (1.44)$$

where

EC_t = total cost of hood(s), ductwork, and stack(s)

1.4.4 Installation Costs

When making a cost estimate for an air pollution control system according to the procedure in this Manual, the estimator first determines the cost of the control device, then estimates the costs of such auxiliaries as the hood, ductwork, stack, fan and motor, and other items. To these items he/she adds the costs of instrumentation, taxes, and freight, to obtain the PEC. Finally, the estimator multiplies the PEC by the installation factor appropriate to the control device (e.g., 2.20 for gas absorbers) to obtain the total capital investment. In these cases, the installation factor incorporates all direct and indirect costs needed to install and start up the control system equipment, including, of course, the hood, ductwork, and stack.

For this reason, it usually is unnecessary to estimate the installation cost of the ventilation system separately. However, there may be occasions where a hood, a stack, or ductwork has to be installed alone, either as replacement equipment or to augment the existing ventilation system. In those instances, the estimator may want to estimate the cost of installing this item.

As might be imagined, these installation costs vary considerably, according to geographic location, size and layout of the facility, equipment design, and sundry other variables. Nonetheless, some of the vendors (and a peer reviewer[59]) provided factors for hoods and ductwork, which, when multiplied by their respective purchased equipment costs, will yield approximate installation costs. These are:

- Hoods: 50 to 100%
- Ductwork: 25 to 50%

If one or both of these factors is used, the total capital investment (TCI) of the hood and/or ductwork would be:

$$TCI = (1 + IF_{h/d}) \times PEC_{h/d} \qquad (1.45)$$

where

$IF_{h/d}$ = installation factor for hood(h)/ductwork(d)

$PEC_{h/d}$ = purchased equipment cost of hood (h)/ductwork (d)

1.5 Estimating Total Annual Cost

1.5.1 Direct Annual Costs

Ventilation systems incur few, if any, direct annual costs, as they function to support control devices. There are no costs for operating or supervisory labor, operating materials, or waste treatment/disposal allocated to ventilation systems. Maintenance costs would also be minimal, except for such minor expenses as painting, insulation repair, or calibration of automatic damper controls. The only utilities cost would be the incremental electricity needed for the waste gas stream to overcome the static pressure loss in the hood, ducting, and stack.[3] The incremental electricity cost (C_c, \$/yr) can be calculated as follows:

$$C_c = (1.175 \times 10^{-4}) \frac{p_c Q F_d \theta}{\epsilon} \qquad (1.46)$$

where

p_c	=	electricity price (\$/kwh)
Q	=	waste gas flow rate (actual ft^3/min)
F	=	static pressure drop through ventilation system (in. w.c.)
θ	=	operating factor (hr/yr)
ϵ	=	combined fan-motor efficiency

Illustration: In the cosmetic factory ventilation system, what would be the cost of the electricity consumed by the fan needed to convey the gas through the ductwork? Assume an electricity price of \$0.075/kwh, a combined fan-motor efficiency of 0.6, and an 8,000-hr/yr operating factor.

Solution: Recall that the pressure drop and gas flow rate for this illustration were 0.313 in. w.c. and 15,000 actual ft^3/min, respectively. Upon substituting these values and the other parameters into equation 10.40, we obtain:

$$C_c = (1.175 \times 10^{-4})\ (0.075)\ (15{,}000)\ (0.313)\ (8{,}000)/0.6 = \$552/\text{yr}.$$

1.5.2 Indirect Annual Costs

The indirect annual costs for ventilation systems include property taxes, insurance, general and administrative (G&A), and capital recovery costs. (Overhead—a fifth indirect annual cost—is not considered, because it is factored from the sum of the operating, supervisory, maintenance

[3] Technically, this direct annual cost should be allocated to the ventilation system fan, not to the hood, ductwork, and stack. The fan power cost equation will be included in the Manual section on fans. However, as the fans section has yet to be written, this equation has been provided as a temporary convenience to Manual users.

labor and maintenance materials costs, which is negligible.) When a ventilation system is part of a control system, these costs are included in the control system indirect annual cost. However, if the ventilation equipment has been sized and costed separately, these costs can be computed from the total capital investment (*TCI*) via standard factors, as follows:

The "*CRF*" term in the capital recovery equation is the capital recovery factor, which is a function of the economic life of the ventilation system and the interest rate charged to the total capital investment. (See Section 1, Chapter 2 of this Manual for more discussion of the *CRF* and the formula used for computing it.)

For a ventilation system, the economic life varies from at least 5 to 10 years to 15 to 20 years or more.[60,61] In general, the ventilation equipment should last as long as the control system it supports. As discussed in Section 1, Chapter 2, the interest rate to use in the *CRF* computation should be a "pre-tax, marginal (real) rate of return" that is appropriate for the investor. However, for those cost analyses related to governmental regulations, an appropriate "social" interest (discount) rate should be used. For these kinds of analyses, the Office of Management and Budget (OMB) directs that a real annual interest rate of 7% be used.[62] (This replaces the 10% rate OMB previously had mandated.)

Table 1.13: Computation Equations for Indirect Annual Costs

Indirect Annual Cost	Computation Equation
Property Taxes	0.01 x TCI
Insurance	0.01 x TCI
General and Administrative	0.02 x TCI
Capital Recovery	CFR x TCI

1.5.3 Total Annual Cost

The total annual cost (*TAC*) is calculated by adding the direct (*DC*) and indirect (*IC*) annual costs:

$$TAC = DC + IC \tag{1.47}$$

1.6 Acknowledgements

Several firms and individuals provided very useful technical and cost information to this chapter. Foremost among these was Todd N. Stine of United McGill Corporation (Raleigh, NC), who submitted current prices for a variety of ductwork items, as well as a comprehensive product catalog and engineering design manual. In addition, Mr. Stine patiently replied to the author's many questions, providing supplemental data when requested. Samir Karandikar of EPCON Industrial Systems (The Woodlands, TX) and Gregory P. Michaels of Piping Technology & Products (Houston, TX) also were very helpful in submitting data and responding to inquiries.

The author also would like to thank the following firms for their valuable contributions:

- Air Plastics, Inc. (Mason, OH)
- General Resource Corporation (Hopkins, MN)
- Harrington Industrial Plastics, Inc. (Chino, CA)
- Intellect Systems & Marketing, Inc. (Bohemia, NY)
- Wer-Coy Metal Fabrication Co. (Warren, MI)

In addition, several individuals reviewed the draft chapter and provided valuable suggestions, supplemental information, or both. The EPA peer reviewers, all located at Research Triangle Park, NC, were:

- James C. Berry (OAQPS/ESD)
- Peter A. Eckhoff (OAQPS/TSD)
- Norman Kaplan (ORD/AEERL)
- James H. Maysilles (OAQPS/ESD)
- Larry Sorrels (OAQPS/ESD)

Finally, Howard Goodfellow of Goodfellow Consultants, Inc. (Mississauga, Ontario, Canada) also reviewed this Section and supplied helpful comments.

References

[1] Goodfellow, H.D. "Ancillary Equipment for Local Exhaust Ventilation Systems". In: Air Pollution Engineering Manual. New York, NY: Van Nostrand Reinhold/Air and Waste Management Association. 1992, pp. 155-205.

[2] Burton, D. Jeff. Industrial Ventilation Work Book. Salt Lake City: DJBA, Inc. 1989.

[3] The Measurement Solution: Using a Temporary Total Enclosure for Capture Efficiency Testing. Research Triangle Park, NC: U.S. Environmental Protection Agency. August 1991 (EPA-450/4-91-020), pp. 3,11.

[4] The Measurement Solution, pp. 11-29.

[5] Heinsohn, Robert Jennings. Industrial Ventilation: Engineering Principles. New York, NY: John Wiley & Sons, Inc. 1991.

[6] Telephone conversation between William M. Vatavuk, U.S. Environmental Protection Agency (Research Triangle Park, NC) and Todd N. Stine, United McGill Company (Raleigh, NC), May 24, 1993.

[7] Thermoplastic Duct (PVC) Construction Manual, 2nd Edition 1994. Vienna, VA: Sheet Metal and Air Conditioning Contractors' National Association, Inc. (SMACNA). May 1987, pp. 61-85.

[8] Thermoplastic Duct Construction Manual, p. 64.

[9] Burton, p. 6-7.

[10] Dust Control System Accessories Price List. Huntington Park, CA: Murphy-Rodgers, Inc. July 1992.

[11] Price and Data Catalog: Standard Ductwork Components. Warren, MI: Wer-Coy Metal Fabrication Co. 1992-93.

[12] Letters from Samir Karandikar, EPCON Industrial Systems (Woodlands, TX) to William M. Vatavuk, U.S. Environmental Protection Agency (Research Triangle Park, NC). May 21 and June 9, 1993.

[13] "Double Wall Insulated Duct and Fittings." In: Sheet Metal Division Catalog. Groveport, OH: United McGill Corporation. 1990.

[14] "Single-Wall Round and Flat Oval Duct and Fittings." In: Sheet Metal Division Catalog. Groveport, OH: United McGill Corporation. 1990.

[15] HVAC Duct Construction Standards: Metal and Flexible. Vienna, VA: Sheet Metal and Air Conditioning Contractors' National Association, Inc. (SMACNA). 1985, pp. 2-15 to 2-17.

[16] Wherry, T.C. and Peebles, Jerry R., "Process Control". In: Perry's Chemical Engineers' Handbook, Sixth Edition. New York, NY: McGraw-Hill, Inc. 1984.

[17] Product catalog. Rio, WI: Gaskets, Inc. 1994.

[18] HVAC Duct Construction Standards, pp. 4-2 to 4-3.

[19] HVAC Duct Construction Standards, pp. 4-2 to 4-7.

[20] Letter from Howard D. Goodfellow, Goodfellow Consultants (Mississauga, Ontario, Canada) to William M. Vatavuk, U.S. Environmental Protection Agency (Research Triangle Park, NC). February 23, 1994.

[21] Guide for Steel Stack Design and Construction. Vienna, VA: Sheet Metal and Air Conditioning Contractors' National Association, Inc. (SMACNA). 1983.

[22] Goodfellow, pp. 192-193.

[23] Goodfellow, p. 193.

[24] Peters, Max S. and Timmerhaus, Klaus D. Plant Design and Economics for Chemical Engineers, Third Edition. New York: McGraw-Hill, Inc., 1980, pp. 508-510.

[25] Burton, pp. 2-10 to 2-11.

[26] Burton, p. 2-11.

[27] Burton, pp. 4-5 to 4-8.

[28] Burton, p. 5-12.

[29] Burton, pp. 5-15 to 5-16.

[30] Burton, p. 5-5.

[31] Burton, pp. G-2, G-5.

[32] Burton, p. 5-18.

[33] Vatavuk, William M. and Neveril, Robert B., "Estimating Costs of Air-Pollution Control Systems, Part III: Estimating the Size and Cost of Pollutant Capture Hoods," Chemical Engineering, December 1, 1980, pp. 111 to 115.

[34] Telephone conversation between William M. Vatavuk, U.S. Environmental Protection Agency (Research Triangle Park, NC) and Dennis Woll, Air Plastics, Inc. (OH), August 10, 1993.

[35] Telephone conversation between William M. Vatavuk, U.S. Environmental Protection Agency (Research Triangle Park, NC) and Pat Caputo, Intellect Systems & Marketing, Inc. (Bohemia, NY), October 22, 1993.

[36] Burton, "Chart 9".

[37] Letter from Todd N. Stine, United McGill Corporation (Raleigh, NC) to William M. Vatavuk, U.S. Environmental Protection Agency (Research Triangle Park, NC). June 10, 1993.

[38] Green, Don W. and Maloney, James O. Perry's Chemical Engineers' Handbook, Sixth Edition. New York, NY: McGraw-Hill, Inc. 1984.

[39] Peters, Max S. and Timmerhaus, Klaus D. Plant Design and Economics for Chemical Engineers, Fourth Edition. New York, NY: McGraw-Hill, Inc. 1991.

[40] Engineering Design Reference Manual for Supply Air Handling Systems. Groveport, OH: United McGill Corporation. 1992, pp. 34.

[41] Engineering Design Reference Manual, p. 8.

[42] Burton, "Chart 5".

[43] Engineering Design Reference Manual, p.7.

[44] Burton, p. 6-6.

[45] Burton, "Chart 13".

[46] Goodfellow, p. 193.

[47] Guide for Steel Stack Design and Construction, pp. 39 to 50.

[48] Goodfellow, p. 193.

[49] Carlton-Jones, Dennis and Schneider, H.B., "Tall Chimneys," Chemical Engineering, October 14, 1968, p. 167.

[50] Guideline for Determination of Good Engineering Practice: Stack Height (Technical Support Document for Stack Height Regulations) (Revised). Research Triangle Park, NC: U.S. Environmental Protection Agency. June 1985 (NTIS PB-85-225241), p.1.

[51] Guideline for Determination of Good Engineering Practice, pp. 50-51.

[52] Guideline for Determination of Good Engineering Practice, pp. 1-2.

[53] Goodfellow, p. 194.

[54] Guide for Steel Stack Design and Construction, p. 4.

[55] Hood cost data request responses from four hood vendors to William M. Vatavuk, U.S. Environmental Protection Agency (Research Triangle Park, NC). June-July 1993.

[56] Ductwork cost data request responses from six vendors to William M. Vatavuk, U.S. Environmental Protection Agency (Research Triangle Park, NC). May-July 1993.

[57] Stack cost data request responses from four vendors to William M. Vatavuk, U.S. Environmental Protection Agency (Research Triangle Park, NC). May-July 1993.

[58] Op. cit., Stine-Vatavuk letter, June 10, 1993.

[59] Goodfellow-Vatavuk letter.

[60] Goodfellow-Vatavuk letter.

[61] Telephone conversation between William M. Vatavuk, U.S. Environmental Protection Agency (Research Triangle Park, NC) and Todd N. Stine, United McGill Company (Raleigh, NC), December 10, 1993.

[62] Darman, Richard. Guidelines and Discount Rates for Benefit-Cost Analysis of Federal Programs (OMB Circular No. A-94 Revised). Washington, DC: Office of Management and Budget. October 29, 1992.

EPA/452/B-02-001

Section 2

Generic Equipment and Devices

EPA/452/B-02-001

Chapter 3

Permanent Total Enclosures (PTEs)

Chandrika Prasad
Douglas A. Toothman
Michael E. Lukey
Pacific Environmental Services, Inc.
Herndon, VA 20170

Daniel Charles Mussatti, Senior Economist
U.S. Environmental Protection Agency
Office of Air Quality Planning and Standards
Air Quality Strategies and Standards Division
Innovative Strategies and Economics Group
MD-5: C339-01
Research Triangle Park, NC 27711

September 2002

Contents

3.1 Introduction 3-3

3.2 PTE Criteria 3-4

3.3 PTE Design Parameters 3-5
- 3.3.1 Compliance with EPA Method 204 3-5
- 3.3.2 OSHA Standards 3-7
  - 3.3.2.1 Health Considerations 3-7
  - 3.3.2.2 Safety Considerations 3-11
  - 3.3.2.3 Worker Comfort 3-12
- 3.3.3 Process Configuration 3-13
- 3.3.4 Access to PTE 3-14
- 3.3.5 Size of PTE 3-14
- 3.3.6 Air-Conditioning System 3-16
- 3.3.7 Makeup Air System 3-16
- 3.3.8 Makeup Air Fan 3-16
- 3.3.9 Example Problem 3-17

3.4 Estimating Total Capital Investment 3-23
- 3.4.1 Purchased Equipment Cost 3-23
- 3.4.2 Installation Cost 3-27

3.5 Estimating Total Annual Cost 3-31
- 3.5.1 Direct Annual Cost 3-32
- 3.5.2 Indirect Annual Cost 3-32

3.6 Acknowledgments 3-34

References 3-34

3.1 Introduction

Permanent total enclosures (PTEs) are permanently installed structures that completely surround a source (s) of emissions. PTEs capture all emissions and contain them for discharge to an abatement device such as an incinerator or absorber. PTEs must meet each of the U.S. Environmental Protection Agency's (EPA's) five point criteria listed in Table 3.1.

PTEs are unique because they accommodate production personnel within its structure during operation. Consequently, they have an Occupational Safety and Health Administration (OSHA) regulated system of air flow control for supplying fresh air to the space enclosed. By definition they have a capturing efficiency of 100 percent and do not need to conduct a capture efficiency test. Capture efficiency is a component of overall efficiency, which can be expressed as:

$$OCE = CE \times DE \qquad (3.1)$$

where OCE = overall control efficiency,
CE = capture efficiency, and
DE = destruction or recovery efficiency.

Since capturing efficiency is 100%, the overall efficiency will be equal to the control device destruction efficiency.

In addition to avoiding the need for a capture efficiency test, companies may choose to employ PTEs rather than other capture systems because:

1. A high overall control efficiency is required due to regulations or new source review involving best available control technology (BACT). For example, Subpart KK[1], the National Emissions Standards for Hazardous Air Pollutants (NESHAPs) for the printing industry, requires an overall control efficiency of 95 percent for organic hazardous air pollutants (HAPs) for packaging flexographic and rotogravure presses using only add-on control. BACT for rotogravure presses has been established at 98+ percent overall control efficiency for volatile organic compounds (VOCs). Using PTE assures the source it has fully met (and exceeded) its regulatory requirement.

2. Implementation of more stringent standards for measuring the capture efficiency for an emission source. EPA has established rigorous data quality objectives associated with testing techniques for determining capture efficiency specified in Reference Method 204[2]. Installing a PTE can avoid the need for secondary control in the future due to tightening standards.

3. Continuous compliance requirements under Title V[3], the new Compliance Assurance

Monitoring rule[4], New Source Performance Standards [5], and NESHAPs [6]. Using PTE simplifies meeting this requirement for capture efficiency.

Any process or operation whose emissions are not totally captured is a candidate for a PTE. Industries that have used PTEs as part of control systems [7] are:

- Flexographic printing
- Rotogravure printing
- Coating (paper, film, fabric, plastic, and metal)
- Laminating
- Screen printing
- Can coating
- Plastic card coating

Due to the increasing use of PTEs, EPA has developed a methodology for estimating PTE costs. This methodology is presented in Chapter 2 of this Manual. The purpose of this chapter is to provide a quick means to generate study cost estimates for PTEs.

3.2 PTE Criteria

The EPA's five-point criteria given in EPA Method 204 is reproduced in Table 3.1.

Table 3.1: The EPA Method 204: Criteria for a Permanent Total Enclosure[2]

No.	Description	Requirement
1	Location of openings	Any natural draft opening (NDO) shall be at least four equivalent opening diameters from each VOC emitting point unless otherwise specified by the administrator.
2	Areas of openings	The total area of all NDOs shall not exceed 5 percent of the surface area of the enclosureís four walls, floor, and ceiling.
3	Flow rate into enclosure	The average facial velocity (FV) of air through all NDOs shall be at least 3,600 m/hr [200 ft/min (this equates to a negative pressure difference of 0.007†in. of water or 0.013 mm Hg)]. The direction of air flow through all NDOs must be into the enclosure.
4	Access doors/windows	All access doors and windows whose areas are not included in item 2 and are not included in the calculation in item 3 shall be closed during routine operation of the process.
5	Emission capture	All VOC emissions must be captured and contained for discharge through a control device.

3.3 PTE Design Parameters

PTE design takes into account the following factors [8]:

- Compliance with EPA Method 204
- OSHA Standards
  - Health Considerations
  - Safety Considerations
  - Worker Comfort
- Process Configuration
- Access to PTE
- Size of PTE
- Air-Conditioning System
- Makeup Air System

These factors are described in greater detail below.

3.3.1 Compliance with EPA Method 204

Criterion No. 1 :All NDOs must be at least four equivalent diameters from each emission point.

An NDO is a any opening in the PTE that remains open during operation and is not connected to a duct in which a fan is installed. The dimensions of an NDO and its distance from the to the nearest point of emission are measured to ensure compliance with Criterion No. 1.

The equivalent diameter is calculated using the formula:

$$D = \sqrt{\frac{4A}{\pi}} \tag{3.2}$$

where D = equivalent diameter (in.),
A = area of the NDO (sq. in.) and,
π = 3.1416.

Criterion No. 2: The total area of the NDOs must be less than 5 percent of the enclosure surface area.

Total NDOs and enclosure areas are calculated, including walls, ceiling, and floor of the enclosure.

$$A_{NDO} = \sum_{i=1}^{N} A_i \quad (3.3)$$

$$A_{NDO} < .05\ A_e \quad (3.4)$$

$$A_e = L \times H \quad (3.5)$$

where A_{NDO} =1% of TCI Total NDO area,
Ai = Individual NDO area,
A_e = Total enclosure area,
L = Total enclousure length, and
H = Enclosure height.

Criterion No. 3: The average face velocity (FV) of air thru all NDOs shall be at least 200 fpm and the direction of flow into the enclosure.

The volumetric flow rate of each gas stream exiting and entering the enclosure are measured or otherwise determined and the facial velocity is calculated using the following equation:

$$FV = \frac{Q_o - Q_i}{A_{NDO}} \quad (3.6)$$

where FV = facial velocity (fpm),
Q_o = the total volumetric flow from all gas streams exiting the enclosure through an exhaust duct or hood (acfm),
Q_i = the total volumetric flow from all gas streams entering the enclosure through a forced makeup air duct; zero if no forced makeup air is provided to the enclosure (acfm), and
A_{NDO} = total area of all NDOs (sq ft).

The FV should be at least 200 fpm (3,600 m/hr) for compliance. Q_o is always greater than Q_i, the difference being made up by the air entering the NDOs and louvers.

The direction of air flow through all NDOs is measured or verified to be inward by measuring the pressure difference between the inside and outside of the PTE . The low pressure leg of the device is connected to tubing that terminates inside the enclosure. The high pressure leg opens to the outside of the enclosure. The outside pressure the PTE should be at least 0.007 in. of H_2O

(0.013 mm of Hg) higher than the inside pressure for compliance. If FV is less than 500 fpm, the continuous inward flow of air is verified using streamers, smoke tubes, or tracer gases. If FV is greater than 500 fpm, the direction of air flow through the NDOs is considered to be inward at all times without verification.

Criterion No. 4: All access doors and windows whose areas are not accounted for in Criterion No. 2 and are not included in the calculation for Criterion No. 3 are kept closed during normal operation of the source(s).

Criterion No. 5: All VOCs emitted within the PTE are delivered to an air pollution control device in order to meet this criterion.

3.3.2 OSHA Standards

Regulation of occupational health and safety in the workplace is the responsibility of the Occupational Safety and Health Administration (OSHA), an organization within the Department of Labor of the Federal government. The following section discusses how OSHA requirements affect the design of PTEs.

3.3.2.1 Health Considerations

OSHA adopted permissible exposure levels (PELs) as the best existing standards for worker exposure for a large number of substances. PELs are expressed in terms of time-weighted average (TWA-generally 8-hour), short-term exposure level (STEL), and ceiling concentration (C). OSHA standards are published in 29 CFR Part 1910 Subpart Z [9]. As new information becomes available, PEL values may be changed or new substances may be added to the existing list.

The amount of ventilation air required to maintain VOC concentrations below PELs within an enclosure can be estimated using the following relationship:

$$Q_i = \frac{\sum_{i=1}^{n} K E_i}{\left(\frac{60 \text{ min}}{\text{hr}}\right) PEL_i} \qquad (3.7)$$

where Q_i = ventilation air flow rate (scfm) required to stay below the PEL for VOC i,
i = individual VOC,
Σ = summation over all processes emitting VOC i,
K = fraction of process emissions not immediately captured by the PTE exhaust gas stream,
E_i = process uncontrolled emission rate (lb/hr) for VOC i, and
PEL_i = permissible exposure level (lb/ft^3 at standard conditions — 70°F and 1 atm) for VOC i.

The PELs are generally given in units of milligrams per cubic meter (mg/m^3) and/or parts per million by volume (ppmv). The PEL in lb/ft^3 is obtained by multiplying the PEL in mg/m^3 by 62.43×10^{-9}. The PEL in ppmv is converted to lb/ft^3 at standard conditions by multiplying by a factor of 2.6×10^{-9} M where M is the molecular weight. The volumetric flow rate, scfm, is converted to actual conditions (actual temperature and pressure) using ideal gas laws. The procedure is repeated for every VOC, and the highest value of Q_i is considered the design flow rate, Q.

New equipment is designed to maximize VOC capture, approximately 5 to 10 percent (K = 0.05 to 0.10) of total process emissions escape. For older, poorly maintained equipment (built before 1980), this factor is as high as 30 percent (K = 0.30).

Assuming complete mixing of VOCs within the enclosure, the average VOC concentration in the enclosure is calculated by the following equation:

$$C_{iavg} = \frac{\sum_{i=1}^{n} K E_i}{Q\left(\frac{60 \text{ min}}{\text{hr}}\right)} \qquad (3.8)$$

where C_{iavg} = average concentration (lb/ft^3) for VOC,
Σ = summation over all processes emitting VOC i,
K = fraction of process emissions escaping into the enclosure,
E_i = process emission rate (lb/hr) for VOC i, and
Q = design flow rate (actual ft^3/min) from the enclosure.

In practice, the mixing of VOCs within an enclosure is rarely complete. Mixing is a function of the performance characteristics of the ventilation system which depend upon a number of variables such as:

- Temperature of the delivery supply air
- Temperature within the PTE

- Amount and locations of supply and exhaust air
- Locations of objects within the PTE
- Shape and size of the PTE
- Presence or absence of heat sources within the PTE
- Injection velocity of the supply air

Thermal stratification due to plant equipment results in the buildup of VOC concentrations in certain areas within the enclosure. To accurately determine such stratification, the engineer needs detailed information about the sources, enclosure, and ventilation system. Proper and detailed design of the ventilation system, accounting for the amount and location of incoming air, source locations, location of the exhaust points, amount of exhausted air, etc.

Local concentrations may vary considerably by factors from 1 for well-designed ventilation systems to 10 for poorly designed systems[10] in comparison to the average concentration (see Figure 3.1). Thus,

$$C_{imax} = K_1 \, C_{iavg} \tag{3.9}$$

where C_{imax} = maximum concentration (lb/ft^3) for VOCi,
K_1 = 1 to 10 (depending upon the degree of mixing/circulation), and
C_{iavg} = average concentration (lb/ft^3) for VOC .

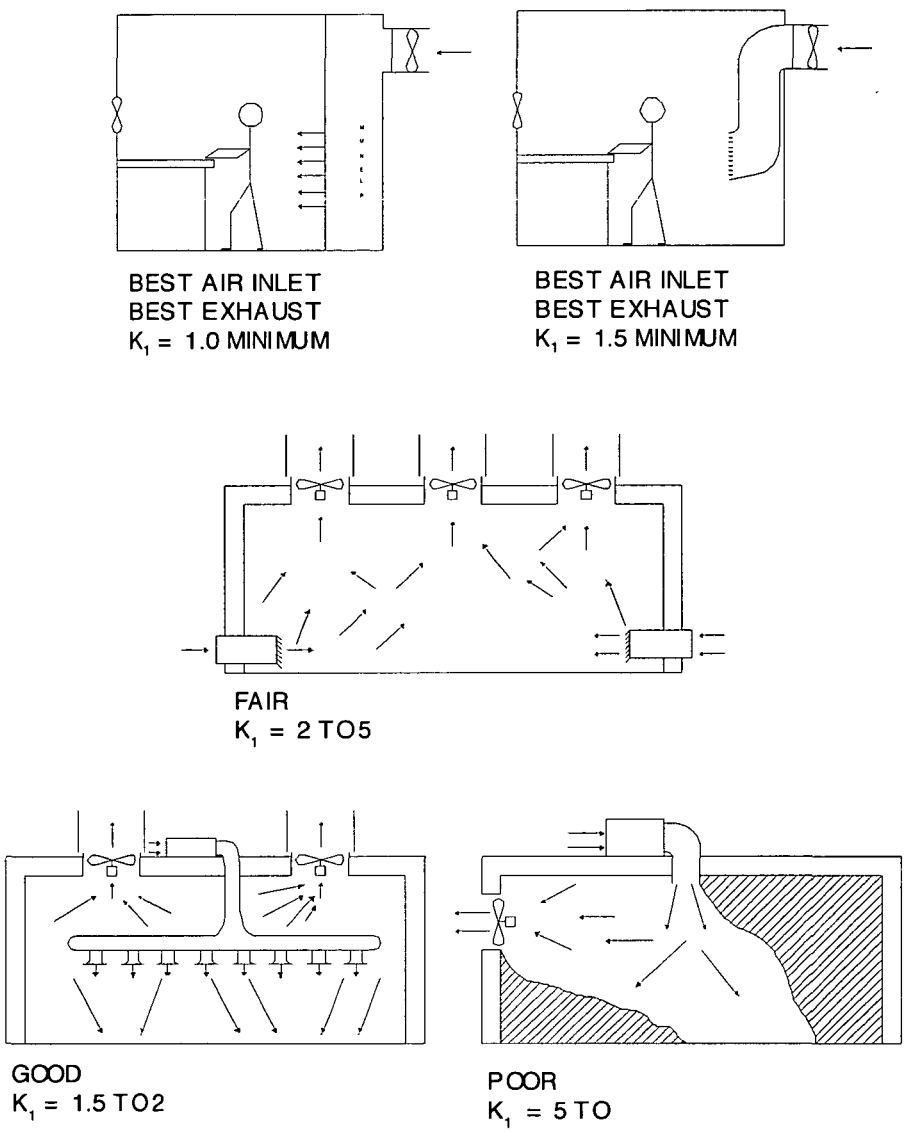


Figure 3.1: Typical K_1 Factors Based on Inlet and Exhaust Locations

Theoretically, C_{imax} should not exceed PEL_i. However, the average concentration, C_{iavg}, within an enclosure is a useful value for comparison to the PEL because a typical operator is constantly moving within the PTE and is not expected to remain at locations with high concentrations for more than a few minutes during the day.

3.3.2.2 Safety Considerations

Combustible gases pose a risk of fire or explosion to personnel and facility. The lowest level at which a gas supports combustion is called the lower flammable limit (LFL) or lower explosive limit (LEL). Below this level the gas is too lean to support combustion. There is a corresponding upper flammable limit (UFL), above which the concentration is too rich to support combustion. Different gases combust at different concentrations.

For fire safety, OSHA requires the concentration of a flammable vapor or mist in a large PTEs, such as an entire building or manufacturing area, not to exceed 25 percent of the LFL. For small PTEs, such as those enclosing a single piece of equipment, concentrations are limited to 10 percent of the LFL.

LFLs rarely govern when evaluating flammable vapor or mist concentrations in an enclosure because PELs are more restrictive. Table 3.2 displays the LFL, safety level, and PEL for many commonly used industrial solvents.

Table 3.2: LFLs, Safety Levels, and PELs for Common Industrial Solvents

Solvent	LFL(ppmv)	10% LFL(ppmv)	PEL(ppmv)
Acetone	25,000	2,500	1,000
Benzene	12,000	1,200	1
Ethyl alcohol	33,000	3,300	1,000
Ethyl acetate	20,000	2,000	400
Hexane	11,000	1,100	500
Isopropyl alcohol	20,000	2,000	400
Isopropyl acetate	18,000	1,800	250
Methyl alcohol	60,000	6,000	200
Methyl ethyl ketone	18,000	1,800	200
Methyl methacrylate	17,000	1,700	100
n-propyl acetate	17,000	1,700	250
n-propyl alcohol	22,000	2,200	200
Styrene	9,000	900	100
Toluene	11,000	1,100	200
Xylene	9,000	900	100

The LFL concentration determines safe levels in enclosed spaces such as baking and drying ovens and ductwork to protect against fires and explosions. Concentrations within the enclosure can be calculated using procedures given in Section 3.4.2.1. To estimate the concentration in a duct, use the following equation.

$$C_{di} = \frac{E_i}{Q\left(\frac{60\ \text{min}}{\text{hr}}\right)} \qquad (3.10)$$

where C_{di} = concentration in the duct (lb/ft^3) under actual conditions for VOC i,
E_i = emission rate through the duct (lb/hr) for VOC i, and
Q = flow rate through the duct (actual ft^3/min).

For safety, the concentration within the duct should not exceed 25 percent of the LFL.

To safeguard against concentrations reaching unsafe levels, PTEs and associated ducts should be equipped with instruments to monitor concentrations. Additional equipment such as alarms can be provided to sound automatically when concentrations reach unsafe levels. The enclosure may also need a water sprinkler system or fire suppression system. Emergency training for the workers may also be required.

Three widely used fire suppression systems are water sprinklers, carbon dioxide (CO 2), and FM200 (a gas developed by Great Lakes Chemical Corporation as a replacement for halon gas). Water sprinkler systems may not be sufficient for special environments, high risk areas, isolated locations, or unusual hazards. For these cases, FM200 and CO2 systems are required.

3.3.2.3 Worker Comfort

Good ventilation is necessary to ensure worker comfort and provide healthful working conditions. The amount of ventilation required is expressed in terms of room air changes per hour (RACs/hr), calculated using the following equation:

$$\frac{\text{RACs}}{\text{hr}} = \frac{Q\left(\frac{60\ \text{min}}{\text{hr}}\right)}{V} \qquad (3.11)$$

where Q = ventilation air required (actual ft^3/min), and
V = volume of enclosure (ft^3) excluding space occupied by equipment.

Generally, it takes 10 to 15 RACs/hr to provide adequate worker comfort within an enclosure. However, the RACs/hr are compared with the amount of ventilation (dilution) air required to safeguard against potential health hazards and fire and explosive conditions.

3.3.3 Process Configuration

Process configuration and the location of emission sources influence PTE design. If sources are located close to each other, a single PTE can be designed to enclose all the sources. In some cases, the entire building or the room can be converted into a PTE with just a change in the ceiling height to satisfy RAC requirements. If the sources are separated by relatively large distances, it may be more appropriate to build several PTEs. In some cases, a PTE is built around only the emitting portion of the source (such as the printing head of a press). For multiple sources within a PTE, there are situations when some of the sources do not need to be controlled to comply with regulations or permit conditions because compliant materials are used. In such cases, a PTE within a PTE allows sources using compliant materials to be vented directly to the atmosphere. If access to the emission point is not required, a small unmanned PTE can be built around it. Such PTEs are constructed as close to the emission point as possible. Although small in size, they require engineering ingenuity for proper design around a complex emission point. Several examples of PTEs are shown in Figures 3.2 through 3.5.

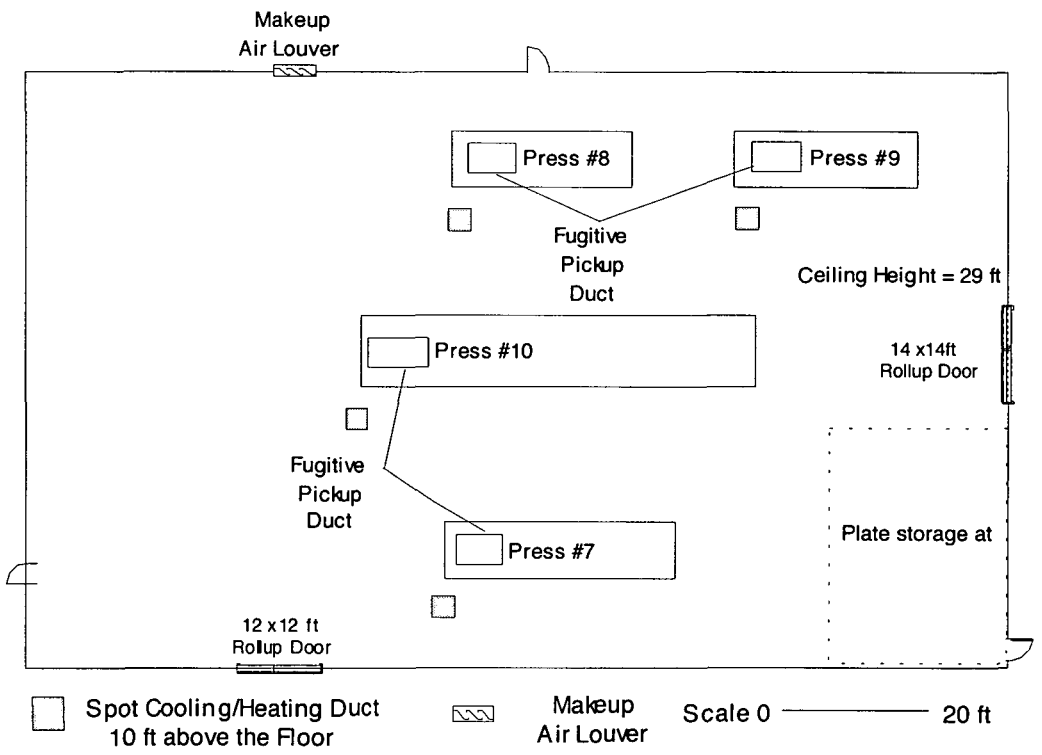


Figure 3.2: Manugacturing Area as a PTE

3.3.4 Access to PTE

Access is needed for material flow into and out of the enclosure. A variety of doors ranging from simple personnel doors to automatic rollup or sliding doors are used. For visual inspection of the process in the PTE, several glass windows are typically recommended.

3.3.5 Size of PTE

The size of a PTE depends largely on two factors: location of the sources and capacity of the existing or proposed air pollution control equipment. Large PTEs require large exhaust flow rates (hence an air pollution control device with a large design flow rate) to provide adequate RACs/hr for worker comfort. If the exhaust flow rate of the control device is relatively small, a smaller PTE is adequate.

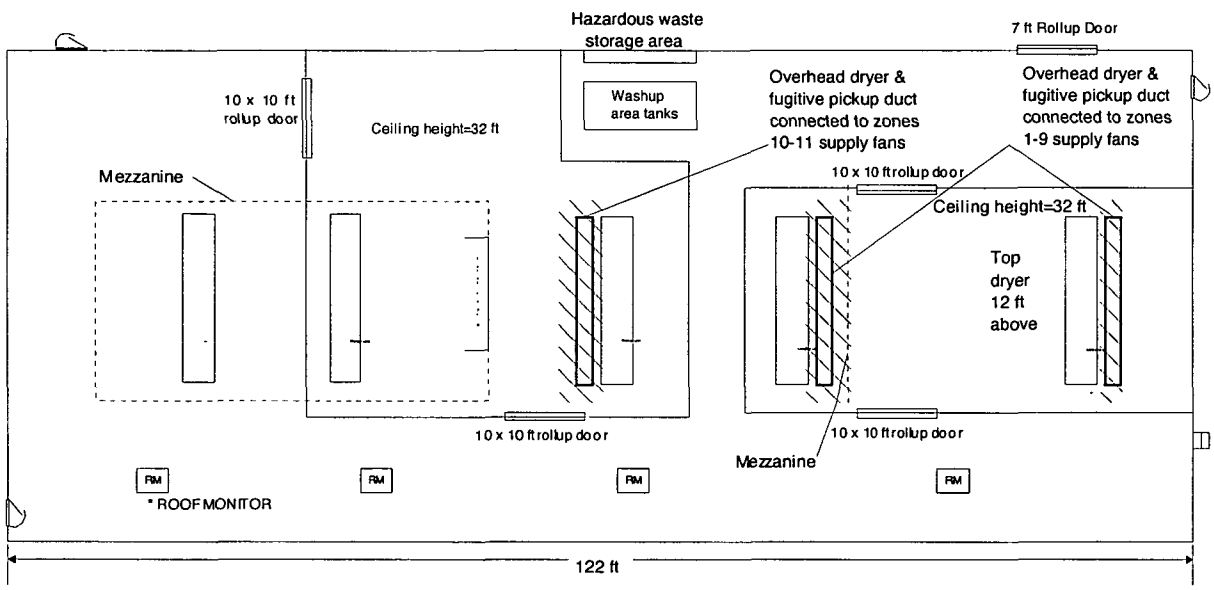


Figure 3.3: PTE Around Several Sources

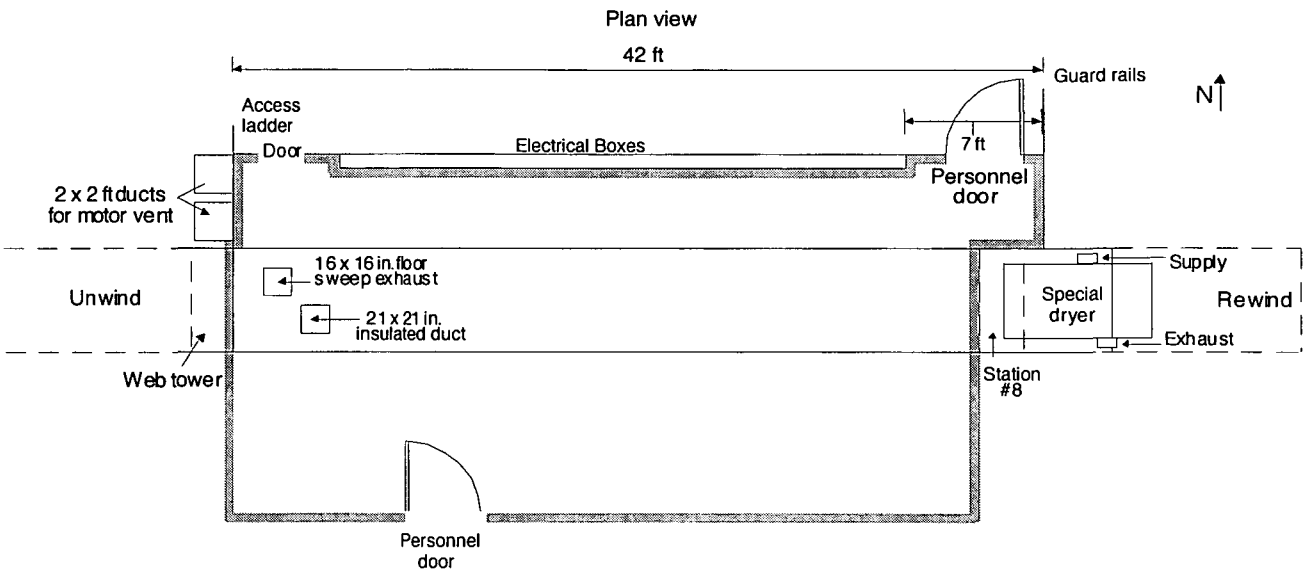


Figure 3.4: PTE Around a Single Source

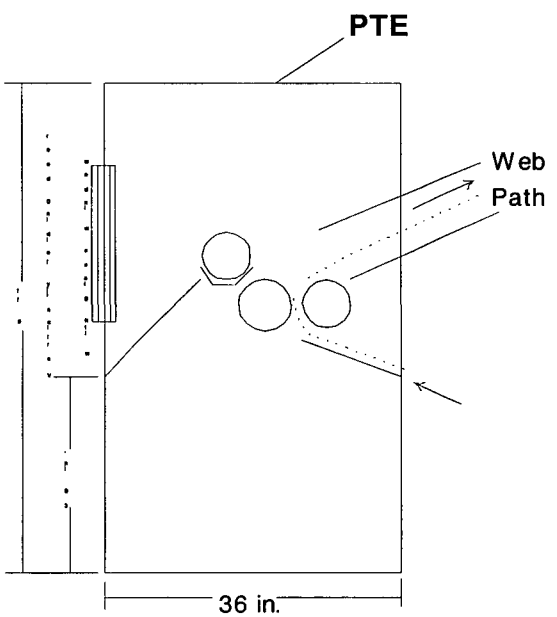


Figure 3.5: Unmanned PTE (Around One Station of a Coater)

3.3.6 Air-Conditioning System

If the installation of a PTE results in heat buildup, some air conditioning can be added. All air conditioning should be of a closed loop design (return air is taken from the interior of the PTE, passes through coils, and is delivered back to the PTE) to avoid violating the PTE criteria. The necessary design criteria for air conditioning are available in the appropriate American Society of Heating, Refrigerating and Air-Conditioning Engineers (ASHRAE) handbooks.

3.3.7 Makeup Air System

Whenever air is exhausted from a room or enclosure, supply air must enter the enclosure to take its place. For low exhaust rates, air entering through NDOs, cracks, and incomplete seals may be adequate. Air flow can be increased by installing louvers in the walls of the enclosure. For large exhaust volumes, additional air must be provided by a fan to make up for the difference between the air exhausted and the air entering through the NDOs and cracks. In some cases makeup air is used for spot cooling where air is introduced directly at the work station. The amount of makeup air required is given by the following equation:

$$Q_m = Q_e - Q_{NDO} \tag{3.12}$$

where Q_m = amount of makeup air (acfm),
Q_e = exhaust flow rate from the PTE (acfm), and
Q_{NDO} = amount of air entering the PTE through NDOs (acfm).

3.3.8 Makeup Air Fan

This section only considers the fan required for makeup air. Exhaust fans that transport gases from the PTE to the control device are part of the control system and not considered in this chapter. The performance of a fan is characterized by its "fan curve" which presents quantitatively the relationship between the volume of air flow, the pressure at which this flow is delivered, the speed of rotation, the power required, and the efficiency. The basic information required to select a fan is the actual volumetric flow rate and the fan static pressure (FSP). Other factors that influence the selection are stream characteristics, drive arrangement and mounting, operating temperatures, inlet size and location, and efficiency. The FSP is defined as follows:

$$FSP = SP_o - SP_i - VP_i \tag{3.13}$$

where FSP = (in. w.c.),
SP_o = static pressure at outlet (in. w.c.),
SP_i = static pressure at inlet (in. w.c.), and
VP_i = velocity pressure at inlet (in. w.c.).

Manufacturers provide multi-rating tables for fan selection in Figure 3.6.[11] For every

volumetric flow rate and fan static pressure, the required fan speed (RPM) and the fan power (BHP) are given. If values in the table fall between desired values, interpolation is acceptable. The multi-rating tables are based on standard conditions of 70°F and 29.92 in. Hg pressure. For a given flow rate and static pressure, several fan selections are possible and it is not unusual to find four or more fan sizes that provide the required flow rate at a given pressure drop. Usually, the fans in the middle of a rating table are the most efficient (about 75-80 percent efficiency). If the design operating parameters are near the top or the bottom of the table, select a smaller or larger fan.

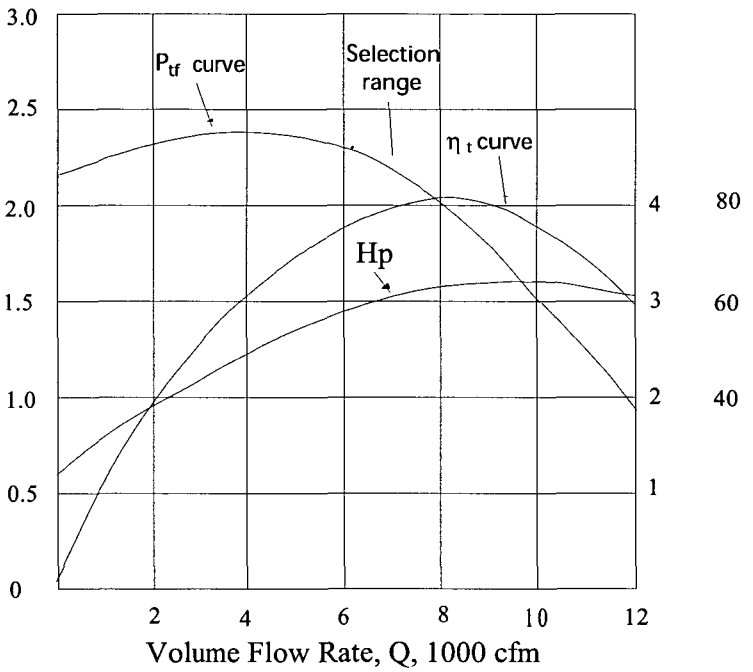


Figure 3.6: Conventional Fan Performance curve used by most manufactures

3.3.9 Example Problem

To illustrate the design process for a PTE, this section provides an example problem and demonstrates how each design parameter meets the EPA five-point criteria previously discussed in Section 3.2.

A high volume specialty packaging products company located in an ozone attainment area

has operated an eight-color rotogravure printing press (Press #1) for 6 years, using only high solvent inks to meet customer demands for quality. To meet the state control technology requirement, the company installed a 15,000 scfm thermal incinerator. The existing incinerator has been tested and demonstrated a destruction efficiency of 95 percent. Due to increased demand for its products, the company now plans to install another eight-color rotogravure printing press (Press #2). In order to meet the best available control technology (BACT) requirement under the Prevention of Significant Deterioration (PSD) regulations, the company proposes to install a 15,000 scfm catalytic incinerator and a permanent total enclosure for the new press. The maximum as-applied ink usage for each press is 400 lb/hr and consists of 12 percent ethyl alcohol, 52 percent ethyl acetate, 4.8 percent toluene, 3.2 percent hexane, and 20 percent solids by weight. The existing configuration of the press room is shown in Figure 3.7.

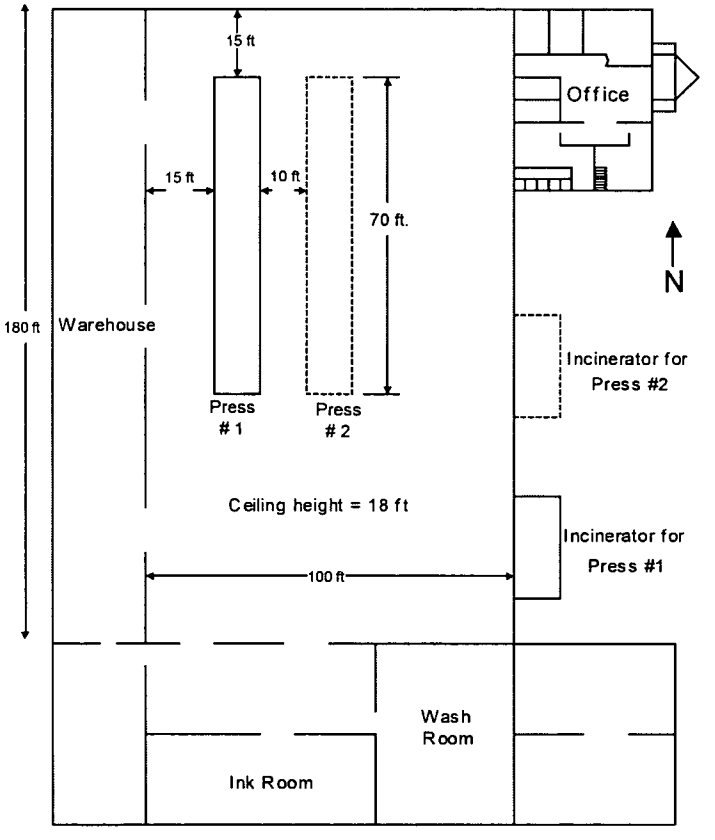


Figure 3.7: Example Plant Layout

The company needs to demonstrate an overall control efficiency of 97 percent for Press #2 as required by the PSD permit. Because the facility is subject to Subpart [KK1] of the MACT standards, an overall control efficiency of 95 percent will have to be demonstrated for Press #1.

To comply with an overall control efficiency of 95 percent, the company must demonstrate a capture efficiency of 100 percent (95/0.95) for Press #1. Assuming the destruction efficiency of the catalytic incinerator to be at least 98 percent, the company must demonstrate a capture efficiency of 99 percent (97/0.98) for Press #2. The designer decided a single PTE around both presses creates fewer obstructions to the work flow and cost less to install.

With a destruction efficiency of 95 percent for the Press #1 incinerator (as tested) and 98 percent for the Press #2 incinerator (as guaranteed), the company can easily demonstrate compliance with the required overall control efficiency.

Size of the PTE

The press room is 100 by 180 ft and the ceiling is 18 ft, resulting in a press room volume of 324,000 cu ft. The exhaust flow rate from the press room is 30,000 scfm (sum of the exhaust flow rates of the two incinerators). Converting the existing press room into a PTE requires the least amount of construction and causes the least amount of disturbance to the occupied space. However, that size enclosure will provide only 5.6 RACs/hr while 10 to 15 air changes provide adequate worker comfort. For this reason, the company decided to build a smaller PTE around the presses.

The existing configuration of the press room is such that a smaller PTE can be built easily by erecting only two additional walls: one on the east side of Press #2 and the other on the south side of the two presses.

In order to provide adequate space for material movement at both ends of the presses, the design places the south wall of the PTE, 15 ft from the nearest end of the presses. On the east side of Press #2, a slightly larger space (20 ft wide)is selected. The overall dimensions of the PTE are 65 ft wide, 100 ft long, and 18 ft high as shown in Figure 3.8, with a volume of 117,000 cu ft. With an exhaust flow rate of 30,000 scfm, this provides 15 RACs/hr.

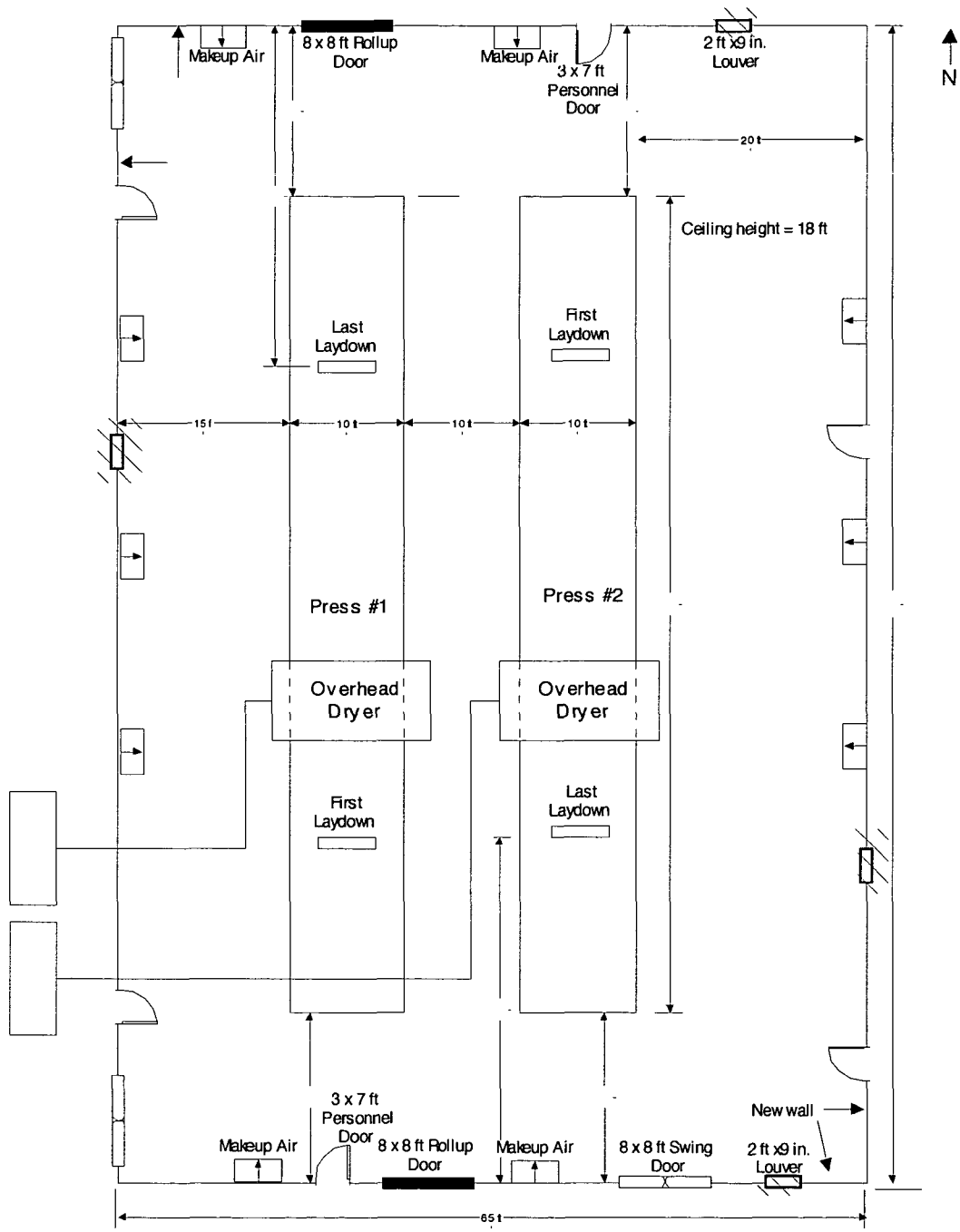


Figure 3.8: Example Permanant Total Enclosure

PTE Wall Material

Typical materials include sheet metal, modular panels, concrete masonry units (CMUs), and drywall. The type of material used depends on the existing equipment and the extent of construction obstructions. In this example, the height of the enclosure is relatively low (18 ft) and the total length of the wall to be constructed is only 165 ft. In addition, there is plenty of room for movement of equipment/material, and there are no obstructions over the presses. Therefore, any appropriate material can be used, depending upon user preference. Due to their ease of construction, portability, and appearance, the engineers selected modular panels for constructing the walls.

Access to the PTE (Doors and Windows)

Most production materials move to the PTE from the warehouse, ink room, and washup areas. Therefore, the design includes two 8 x 8 ft rollup door for the north and south walls of the PTE, (selected to minimize waiting time for material transfers of material in and out of the PTE) three 8 x 8 ft swing doors, two on the west wall, and one on the south wall, and four standard size (3 x 7 ft) personnel doors, one on each wall of the PTE. Windows are not used. In sum, the following access is provided:

Item	Number
Swing doors (8 x 8 ft)	3
Rollup doors (8 x 8 ft)	2
Personnel doors (3 x 7 ft)	4
Windows	None

Louvers

Because of the number of doors specified and the expected frequent opening of the doors on the west and south walls of the PTE, some air movement into the enclosure may occur. However, in order to provide better mixing and ventilation within the PTE, the designers decided to install one large (2ft x 9 in.) louver on each wall of the PTE to minimize the amount of makeup air required. Steel louvers were selected because of their greater durability.

Makeup Air

From the data given in Table 3.3, the total area of the NDOs, not including the louvers, is 198 sq in. (1.38 sq ft). By EPA standards, the average facial velocity must be at least 200 fpm. Furthermore, if the velocity is greater than 500 fpm, the direction of air flow through the NDOs is considered to be inward at all times without verification. The designers used a facial velocity of 600 fpm to provide a margin of safety. With a facial velocity of 600 fpm, the air flow through the NDOs (excluding the louvers) is 728 scfm which is about 3.4 percent of the total air flow (30,000 scfm). By providing louvers (one on each wall of the PTE), the total NDO area is increased to

1,062 sq in. (7.38 sq ft) changing the total air flow through the NDOs to 4,428 scfm. The makeup air required is 25,572 (30,000 - 4,428) scfm. The intake for the makeup air is on the roof of the building and ducted into the PTE at several locations to enhance mixing.

The makeup air required for the example PTE is at least 25,572 scfm. The required static pressure (system resistance) is determined by summing pressure losses through the system components. Assuming that the fan is located in the center on the roof of the PTE and four ducts are installed to deliver makeup air through the four walls of the PTE, the total length of the duct is 400 ft. Each duct carries about 6,393 cfm (a quarter of 25,572 scfm). The number of 90 degree bends required is 12. The radius of curvature of the bends (elbows) is 1.5. The duct diameter is calculated using equations in Section 2, Chapter 1 "Hood, Ducts and Stacks". The flow through each duct is 6,390 scfm. Since the material being conveyed is fresh air, a transport velocity of 2,000 fpm is selected (See Section 2, Chapter 1 "Hood, Ducts and Stacks"). The duct diameter (D_d) is:

$$D_d = 1.128\ (6{,}390/2{,}000)^{½} = 2 \text{ ft}$$

Friction losses are:

F_d $= 0.136\ (½)^{1.18}\ (2{,}000/1{,}000)^{1.8} (400/100) = 0.80$ in. w.c. (straight duct)

VP $= (2{,}000/4{,}016)^2 = 0.25$ in. w.c. (Velocity pressure)

F_c $= 12 \text{x } 0.33 \text{x } 0.25 = 0.99$ in. w.c. (elbows)

Total friction loss $= 1.79$ in. w.c ≈ 1.75 w.c.

Since the inlet to the makeup fan will be open to the atmosphere, the velocity pressure at its inlet will be negligible, the designers selected a fan to provide at least 25,572 scfm (70°F and 1 atm) at a static pressure of at least 1.75 in. w.c.

For a flow rate of 25,572 cfm and static pressure of 1.75 w.c., the wheel diameter of the fan selected is 36.5 in. The required horsepower for the fan selected is 11.0 Hp. Most of the fan manufacturers provide fan motors and starters to match the fan load.

Duct

The example system requires 400 ft of makeup air duct, with a diameter of 2.04 ft. From a number of designs and materials available, the designers choose: duct fabricated from spiral-wound, galvanized carbon steel sheet, four galvanized carbon steel butterfly dampers and twelve 90 degree elbows.

Other Considerations

Other factors in the design of PTEs are air conditioners, safety equipment, hoods, lighting and instrumentation. Although important, they are not part of the scope of this chapter.

3.4 Estimating Total Capital Investment

This section presents the procedures and data necessary for estimating capital costs for PTEs. Total capital investment, TCI, includes purchased equipment cost and direct and indirect installation cost. Most costs in this chapter are presented in 1st quarter 1997 dollars and represent a national average.

3.4.1 Purchased Equipment Cost

Walls

Materials typically used are concrete masonry units (CMUs), drywall, sheet metal, and modular panels. The cost of these materials is commonly given in terms of $/sq ft.

Table 3.3: Cost for different construction materials[12,18].

Wall Material	Purchased cost (1997 $/sq ft)
CMUs	1.49
Drywall	0.59
Sheet metal	1.69
Modular panels	9.76
Lexan (thermoplastic polymer) -Ω in. thick	8.00
Lexan (thermoplastic polymer) -æ in. thick	12.00

The total area of the wall is 2,970 sq ft. Although modular panels are expensive, they were selected because of their appearance (finished on both sides), strength, and flexibility in modifying or relocating. At a unit cost of $9.76/sq ft, the material cost for walls is:

Wall material cost =$9.76/sq ft x 2,970 sq ft = $28,987

Doors

Doors are made in several standard sizes. Their prices are given by units.

Table 3.4: Cost of different door types[12].

Door type	Purchased cost (1997 $ each)
Strip curtain door, 3 x 7 ft	275
Strip curtain door, 8 x 8 ft	575
Personnel door, steel, 3 x 7 ft	575
Sliding door, strip curtain, 8 x 8 ft	780
Sliding door, steel 8 x 8 ft	1,240
Bump door, steel, 3 x 7 ft	1,290
Bump door, steel, 8 x 8 ft	1,830
Rollup door, low speed, 8 x 8 ft	4,255
Rollup door, high speed, 8 x 8 ft	10,165

The estimated cost of the nine doors needed in the design is:

Personnel doors purchased cost	4 doors x $575 each	= $ 2,300
Swing doors purchased cost	3 doors x $1,830 each	= $ 5,490
Rollup doors purchased cost	2 doors x $10,165 each	= $20,330

Auxiliary Equipment

Louvers

Prices for louvers correlate well with the size of the louvers. The following equations can be used to estimate the national average cost of steel and aluminum louvers:

$$\text{Steel louvers[13]:} \quad C = 42 + 35\,A \quad (1.5 \le A \le 7.75) \tag{3.14}$$

$$\text{Aluminum louvers[13]:} \quad C = 47 + 39\,A \quad (1.5 \le A \le 7.75) \tag{3.15}$$

where C = cost for each louver in 1997 dollars and
A = surface area of each louver in sq ft.

Four steel louvers (2 ft x 9 in.) are required for the example PTE. The cross sectional area of each louver is 1.5 sq ft. Using equation (3.13), the louvers' purchased cost is:

Purchased cost per louver	42 + (35 x 1.5)	= $94.5 each
Purchased cost for 4 louvers	4 x $94.5 each	= $378

LFL Monitors

Table 3.5: Cost of LFL monitors [12,16,17]

Item	Price(1997 $ each)
%LFL monitor using flame ionization detectors	10,845
%LFL monitor using catalytic bead	3,325

Because of their lower cost, catalytic bead type LFL monitors were selected.

Cost of 2 catalytic bead monitors = 2 x $3,325 = $6,650

Safety Equipment

Table 3.6: Cost of miscellaneous safety equipment[13,17]

Item	Price(1997 $ each)
Smoke detector (ceiling type)	75
Smoke detector (fixed temperature)	28
Alarm bell	70
Alarm siren	131
Alarm signal	50
Flame detector	2,925

Two ceiling type smoke detectors were selected.

Cost of smoke detectors (ceiling type) = 2 x $75 each = $150

Ductwork

Procedures for designing and estimating costs for ductwork systems are given in Section 2, Chapter 1 "Hoods, Ducts and Stacks" of this Manual. The total ductwork cost is comprised of the cost of its components: straight duct, elbows and dampers

Table 3.7: Cost of Ductwork

Item	Quanity	Cost
Straight duct cost ($/ft)	$=1.71(24)^{0.936}$	= $33.49/ft
Total cost for 400 ft	=$33.49/ft x 400 ft	= $13,395
Elbow cost ($)	$=58.9\ e^{0.0633(24)}$	= $269 each
Cost for 12 elbows	=12 x $269	= $3,229
Damper cost ($)	$=50.2\ e^{0.0597(24)}$	= $210 each
Cost for four dampers	=4 x $210	= $840
Total cost for ductwork	**= $13,395 +3,229 + 840 = $17,464**	

Fans, Motors, and Starters for Makeup Air

The fan cost equation is presented as follows:

$$C = 56.3\ D^{1.2} \quad (12.25 \leq D \leq 36.5) \tag{3.16}$$

where C = cost in 1997 dollars, and
D = fan wheel diameter (in.).

The wheel diameter of the makeup air fan is 36.5 in. By substituting in Equation 11.14, the total cost of the fan, belt-driven motor, and starter is $4,219.

Instrumentation

One of the five-point criteria for a PTE is to maintain a negative pressure of 0.007 in. w.c. in the PTE. This requires an extremely sensitive and reliable pressure monitor. In addition to the monitor, most vendors recommend a pressure surge damper (to dampen sudden pressure changes). The prices for the equipment are given below.

Table 3.8: Cost of instrumentation equipment [14]

Item	Price(1997 $ each)
Differential pressure monitor	487
Surge damper	22
Alarm	20
Total cost	**$529**

Freight and Taxes

Freight charges depend upon the distance between the site and vendor. Sales taxes depend upon the location of the site and the vendor. National average values for freight and taxes are 5 percent and 3 percent of the total equipment cost.

Table 3.9: Total Purchase Equipment Cost (PEC)

Item		Cost ($)
Basic and Auxiliary Equipment		
Walls		28,987
Doors		29,270
Louvers		378
LFL monitors		6,650
Smoke detectors		150
Makeup air ductwork		17,465
Fan, motor, starter		4,219
Total equipment cost (TEC)		**87,120**
Instrumentation Equipment		529
Freight charges	0.05 x 87,120	4,356
Taxes	0.03 x 87,120	2,614
Total Purchased Equipment Cost (PEC)		**94,619**

3.4.2 Installation Cost

Direct Installation Cost

The direct installation cost consists of installation costs for the basic equipment, auxiliary equipment, and instrumentation.

Walls

Major factors affecting the installation cost for walls are the existing equipment and extent of obstructions. The national average costs of installation for walls assuming moderate obstructions are given in Table 3.10 (multiply these costs by a factor of 1.5 for severe obstruction to construction [12,15]):

Table 3.10: Cost of Wall Installation Based on Material[12,18]

Wall material	Direct installation cost (1997 $/sq ft)
CMU	3.10
Drywall	2.90
Sheet metal	12.91
Modular panels	7.97
Lexan (thermoplastic polymer) -Ω in. thick	2.90*
Lexan (thermoplastic polymer) -æ in. thick	2.90*
*Assumed same as drywall[12,18]	

Installation cost of modular panel walls with an area of 2,970 sq ft:

$$= \$7.97/\text{sq ft} \times 2{,}970 \text{ sq ft} = \$23{,}671$$

Doors

Table 3.11: Cost of door installation based on type [12]

Door type	Direct installation cost(1997 $ each)		
Strip curtain door, 3 x7 ft	240		
Strip curtain door, 8 x 8 ft	285		
Personnel door, steel, 3 x 7 ft	415		
Sliding door, strip curtain, 8 x 8 ft	890		
Sliding door, steel, 8 x 8 ft	1,745		
Bump door, steel, 3 x 7 ft	730		
Bump door, steel, 8 x 8 ft	2,575		
Rollup door, low speed, 8 x 8 ft	3,045		
Rollup door, high speed, 8 x 8 ft	3,910		
Personnel door installation cost	4 doors x $415 each	=	$1,660
Swing doors installation cost	3 doors x $2,575 each	=	$7,725
Rollup door installation cost	2 doors x $3,910 each	=	$7,820
Total		=	**$18,035**

Auxiliary Equipment

Louvers

The labor cost for installing louvers corresponds to the size of the louver. The following equations can be used to estimate the national average cost of installation for steel and aluminum louvers:

Steel louvers[13]: $C = 8 + 1.7\ A$ each $(1.5 \leq A \leq 7.75)$ (3.17)

Aluminum louvers[13]: $C = 9 + 1.9\ A$ each $(1.5 \leq A \leq 7.75)$ (3.18)

where C = cost in 1997 dollars, and
A = louver surface area in sq ft.

The area of the selected louvers is 1.5 sq. ft, therefore:

Louver installation cost	8 + (1.7 x 1.5)	=	$10.55 each
Installation cost for 4 louvers	4 x $10.55 each	=	$42

LFL Monitors

Table 3.12: Installation cost for LFL monitors

Item	Installation cost(1997 $ each)
%LFL monitor using flame ionization detectors	2,700
%LFL monitor using catalytic beads	1,000

Total Installation cost of two catalytic bead monitors = 2 x $1,000 each = $2,000

Ducts

As discussed in Section 2, Chapter 1, the installation cost for ductwork varies from 25 to 50 percent of the material cost. Assuming an average of 37.5 percent, the installation cost for the makeup air ductwork is estimated as:

Makeup air duct installation cost	=	37.5% x material cost	
	=	0.375 x $17,464	=$ 6,549

Fans, Motors, and Starters for Makeup Air

Installation costs for fans, motors, and starters are given by the following equations:

Fans[16]:	$C = 51.89\ D - 380.9$	$(10 \leq D \leq 20)$	(3.19)
Motors[17]:	$C = 43 + 2.16\ H$	$(2 \leq H \leq 100)$	(3.20)
Starters[18]:	$C = 78.68\ Ln(H) - 15$	$(2 \leq H \leq 100)$	(3.21)

where C = cost in 1997 dollars,
D = fan wheel diameter (in.),
H = fan motor and starter horsepower, and
Ln(H) = natural log of horsepower.

Since the installation cost equation applies only to diameters of 10 to 20 in. The installation cost for a 36.5 in. fan can be calculated on the basis of two fans with a wheel diameter of 18.25 in. each. This yields a fan installation cost of $1,132. The installation cost of a single fan 36.5 in. in diameter can also be calculated based on extrapolation of Equation 3.19. This yields:

Fan installation cost = (51.89 x 36.5) - 380.9 = $1,513.

Equation 3.20 yields:

Fan motor installation cost = 43 +(2.16 x 11) = $ 67

Equation 3.21 yields:

Motor starter installation cost = 78.68 Ln(11.0) -15 = $174

Total fan, motor, and starter cost = $1,513+67+174 = $1,754

Instrumentation

Table 3.13: Installation cost for instrumentation components [13,15]

Item	Installation cost(1997 $ each)
Differential pressure monitor	200
Surge damper	20
Alarm	60
Total	280

Indirect Installation Cost

Indirect installation costs are generally estimated from a series of factors applied to the purchased equipment cost. For PTEs, these costs are not dependent on the purchased equipment cost and national average indirect costs related to installation are used.

Table 3.14: Indirect Installation Costs [12,15]

Engineering	$ 5,000
Contractors	15,000
Compliance Test	2,500
Total indirect	$22,500

The total capital investment (TCI) is the sum of the purchased equipment cost, direct installation cost, and indirect installation cost.

Table 3.15: Total Capital Investment

Item	Cost ($)
Purchased Equipment	94,619
Direct Installation Cost	
Walls	23,671
Doors	18,035
Louvers	42
LFL monitors	2,000
Smoke detectors	78
Makeup air	6,549
Ductwork	
Fan, motor, starter	1,754
Differential pressure	200
Monitor	
Surge damper	20
Alarm	60
Indirect Installation Cost	22,500
Total Capital Investment	**169,528**

3.5 Estimating Total Annual Cost

The total annual cost is the sum of direct and indirect annual costs and the recovery credit. Recovery credits represent the value of materials or energy recovered by the control system. Recovery credits are usually associated with control equipment not applicable to PTEs.

3.5.1 Direct Annual Cost

There are no costs for operating, supervisory labor, operating materials, or waste disposal allocated to a PTE. Maintenance costs will be minimal, except for such minor expenses as painting, repairs, or calibration of instruments. The operating cost is the one for only utilities electricity used to operate the auxiliary equipment such as supply fans for makeup air and air conditioning if needed.

The national average electricity cost for operating the supply fan is estimated as follows:

$$C_e = \frac{(1.175 \times 10^{-4}) P_e Q \Delta P S \Theta}{\eta} \qquad (3.22)$$

where
- C_e = electricity cost ($/yr),
- 1.175×10^{-4} = a dimensionless conversion factor,
- P_e = electricity price ($/kWh),
- Q = exhaust flow rate (acfm),
- ΔP = static pressure drop through the makeup air system (in. w.c.),
- S = specific gravity with respect to air (=1),
- Θ = operating hours per year and
- η = combined fan-motor efficiency.

The electricity cost is calculated as follows:

Makeup air flow rate	=	26,200 acfm
Static pressure drop	=	1.75 in. w.c.
Electricity price	=	$0.06/kWh
Operating hours	=	8,760 hr/yr (maximum possible in a year)
Overall efficiency	=	0.75

Substituting these values yields a direct annual cost of $3,775 per year.

3.5.2 Indirect Annual Cost

The indirect annual costs for a PTE include property taxes, insurance, general and administrative charges, overhead, and capital recovery costs. These costs can be estimated from the total capital investment (TCI) using standard factors from this Manual as given below:

Table 3.16: Indirect Annual Cost Factors

Item	Factor
Property taxes	1% of TCI
Insurance	1% of TCI
General & administrative	2% of TCI
Capital Recovery	Capital Recovery Factor x TCI

The TCI is $169,528. Overhead is not considered because it is based on the sum of the operating, supervisory, and maintenance labor and materials costs, which are negligible for a PTE. For the example PTE, the cost for the first three items is:

Property taxes	=	0.01 x $169,528	=	1,695
Insurance	=	0.01 x $169,528	=	1,695
General and administrative	=	0.02 x $169,528	=	3,391
Total			=	$6,781

The capital recovery factor (CRF) is a function of the economic life of the equipment and the interest charged on the total capital investment previously discussed in this Manual:

$$CRF = I(1+I)^n/[(1+I)^n-1]$$

where I = annual interest rate in fraction (i.e., 7% = 0.07) and
n = economic life in years.

For a PTE, the economic life is the same as the life of the building which might be 20-30 years or of the particular equipment enclosed by the PTE which might be less. The interest rate value recommended by the Office of Management and Budget (OMB) is 7 percent. (This replaces the 10 percent rate previously recommended by OMB.) An economic life of 30 years and an interest rate of 7 percent yields a CRF of 0.080586.

Capital recovery	=	0.080586 x TCI	
	=	0.080586 x 169,528	= $13,662
Total annual indirect cost	=	$6,781 +$13,662	= $20,443.

Total annual cost is calculated as follows:

Total direct cost	=	3,775
Total indirect cost	=	20,443
TOTAL	=	$24,218

3.6 Acknowledgments

The authors would like to acknowledge the contributions of the following firms and individuals who provided the necessary data in writing this chapter.

- Robert D. Holton of Herr Industrial, Inc. (Lancaster, PA)
- M & R Construction (Richmond, VA)
- Sierra Monitor Corportation (Milpitas, CA)
- Modus Instruments, Inc. (Northboro, MA)
- Robinson Industries, Inc.(Zelionople, PA)
- Greenheck Corportation (Schofield, WI)
- Guardian Fire Protection Services, Inc. (Gaithesburg, MD)
- Overhead Door Company of Washington, DC (Beltsville, MD)
- Dwyer Instruments, Inc. (Michigan City, IN)
- Mark Elder & Associates, P.C (Norfolk, VA)

References

[1] 40 Code of Federal Regulations Part 63 Subpart KK

[2] 40 Code of Federal Regulations Part 51 Appendix M

[3] 40 Code of Federal Regulations Part 70

[4] 40 Code of Federal Regulations Part 64

[5] 40 Code of Federal Regulations Part 60

[6] 40 Code of Federal Regulations Parts 61 and 63

[7] Lukey, Michael E. "Total Permanent Enclosures for the Surface Coating Industry," Air and Waste Management Association Specialty Meeting, Paper No. 92-51-07, May 1, 1992.

[8] Lukey, Michael E. "Designing Effective and Safe Total Permanent Enclosures," Air and Waste Management Association Specialty Meeting, Paper No. A1098, February 15, 1993.

[9] 29 CFR Part 1910 Subpart Z

[10] Worrall, Mike. "Practical Considerations for Permanent Total Enclosures" in: the Proceedings of the Specialty Conference on Emerging Solutions to VOC and Air Toxic Control, February 26-28, 1997, San Diego, CA.

[11] ASHRAE. 1992. 1992 ASHRAE Handbook HVAC Systems and Equipment. Atlanta, GA.

[12] Letters and data from Robert D. Holton, Herr Industrial, Inc. (Lancaster, PA) to Douglas A. Toothman, Pacific Environmental Services, Inc. (Herndon, VA). November 18, 1997 and December 24, 1997.

[13] Means, R.S. Facilities Construction Cost Data (Kingston, MA), 1997.

[14] Price and Data Catalog: Room Pressure Monitor. Modus Instruments, Inc. 1994 and telephone conversation between Modus Instruments, Inc. (Northboro, MA) and Chandrika Prasad, Pacific Environmental Services, Inc. (Herndon, VA). December 29, 1997.

[15] Personal experiences of senior staff members, Pacific Environmental Services, Inc., (Herndon, VA) Air Quality Planning and Stack Sampling groups.

[16] Letters and data from Stephen R. Ferree, Sierra Monitor Corporation (Milpitas, CA) to Chandrika Prasad, Pacific Environmental Services, Inc. (Herndon, VA). September 23, 1997 and December 30, 1997.

[17] Telecommunication between Fred Henderson of Commonwealth Controls (Annapolis, MD) and Chandrika Prasad, Pacific Environmental Services, Inc. (Herndon, VA). June 12, 1998.

[18] Read Plastics, Rockville, MD, September 8, 1998.

[19] Telecommunication between Austin Chadwick of E. and K. Controls (Leesburg, VA) and Chandrika Prasad, Pacific Environmental Services, Inc. (Herndon, VA). June 12, 1998.

TECHNICAL REPORT DATA

(Please read Instructions on reverse before completing)

1. REPORT NO.	2.	3. RECIPIENT'S ACCESSION NO.
452/B-02-001		

4. TITLE AND SUBTITLE	5. REPORT DATE
The EPA Air Pollution Control Cost Manual	January, 2002
	6. PERFORMING ORGANIZATION CODE

7. AUTHOR(S)	8. PERFORMING ORGANIZATION REPORT NO.
Daniel Charles Mussatti	

9. PERFORMING ORGANIZATION NAME AND ADDRESS	10. PROGRAM ELEMENT NO.
U.S. Environmental Protection Agency Office of Air Quality Planning and Standards Air Quality Standards and Strategies Division Innovative Strategies and Economics Group Research Triangle Park, NC 27711	11. CONTRACT/GRANT NO.

12. SPONSORING AGENCY NAME AND ADDRESS	13. TYPE OF REPORT AND PERIOD COVERED
Director Office of Air Quality Planning and Standards Office of Air and Radiation U.S. Environmental Protection Agency Research Triangle Park, NC 27711	Final
	14. SPONSORING AGENCY CODE EPA/200/04

15. SUPPLEMENTARY NOTES

Updates and revises EPA 453/b-96-001, OAQPS Control Cost Manual, fifth edition (in English only)

16. ABSTRACT

In Spanish, this document provides a detailed methodology for the proper sizing and costing of numerous air pollution control devices for planning and permitting purposes. Includes costing for volatile organic compounds (VOCs); particulate matter (PM); oxides of nitrogen (NOx); SO2, SO3, and other acid gasses; and hazardous air pollutants (HAPs).

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Economics Cost Engineering cost Sizing Estimation Design	Air Pollution control Incinerators Absorbers Adsorbers Filters Condensers Electrostatic Precipitators Scrubbers	

18. DISTRIBUTION STATEMENT	19. SECURITY CLASS *(Report)*	21. NO. OF PAGES
Release Unlimited	Unclassified	1,400
	20. SECURITY CLASS *(Page)* Unclassified	22. PRICE

EPA/452/B-02-001

Section 2

Generic Equipment and Devices

EPA/452/B-02-001

Chapter 4

Monitors

Daniel C. Mussatti
Innovative Strategies and Economics Group
Air Quality Strategies and Standards Division
Office of Air Quality Planning and Standards
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

Margaret Groeber
SAIC
2260 Park Ave, Suite 402
Cincinnati, OH 45206

Dan Maloney
D&E Technical
1008 W. William Street
Champaign, IL 61821

Walter Koucky and
Paula M. Hemmer
E. H. Pechan & Associates, Inc.
3622 Lyckan Parkway, Suite 2002
Durham, NC 27707

October 2000

Contents

4.1 Introduction 4-3

4.2 Continuous Emissions Monitoring Systems 4-4
- 4.2.1 Sampling Systems 4-5
  - 4.2.1.1 Extractive CEMS 4-6
  - 4.2.1.2 In Situ CEMS 4-8
- 4.2.2 Monitors and Gas Analyzers 4-10
- 4.2.3 Data Acquisition System 4-13

4.3 Parametric Monitoring 4-14
- 4.3.1 Particulate Matter (PM) 4-16
- 4.3.2 Sulfur Dioxide 4-18
- 4.3.3 Carbon Monoxide 4-19
- 4.3.4 Nitrous Oxides 4-21
- 4.3.5 Opacity 4-23
- 4.3.6 VOCs 4-23
- 4.3.7 DAS 4-26

4.4 Estimating Capital and Annual Costs for CEMS 4-26
- 4.4.1 Development of Cost Equations 4-27
- 4.4.3 Total Annual Costs 4-33

4.5 Sample Calculation 4-34

4.6 Acknowledgements 4-36

References 4-36

4.1 Introduction

Emissions monitoring is an increasingly important part of air pollution control. Air pollution legislation often takes the form of emissions limits or guidelines which an industrial process must meet. Monitoring demonstrates compliance with regulatory or permit limits. In addition, monitoring provides information regarding gaseous pollutants and particulate matter released into the atmosphere that can be used for compiling emissions inventory data, permitting new and existing facilities, and performing audits. Industrial facilities can use emissions monitoring to assess and monitor process control and efficiency, to determine pollution control device efficiency, and to monitor health and safety within the plant. Participation in emissions trading programs generally requires emissions monitoring.

The term monitor refers to a wide variety of instrumentation used to measure the concentration of both gaseous compounds, particulate matter and physical properties such as opacity in a waste gas stream. There are many different types of monitors commercially available for emissions monitoring. Monitors generally require additional equipment for sample collection, calibration of instruments, and data acquisition and processing. Monitors must be able to provide accurate reproducible data.

The 1990 Clean Air Act required enhanced and periodic monitoring for specific pollutants at various stationary sources. These requirements were codified in the Compliance Assurance Monitoring (CAM) Rule. Emissions units with air pollution control equipment at sources regulated under Title V are required to have CAM. CAM requires a modification to the Title V permit to include a program to establish monitoring adequate to demonstrate compliance with applicable regulations. Title V recordkeeping and reporting requirements apply to CAM affected units. States have flexibility in establishing adequate CAM approaches.

Under the CAM Rule, there are two viable monitoring options for monitoring source compliance with permits or regulations. The first option is **continuous emissions monitoring** (CEM), which is a direct measurement of pollutant concentration from a duct or stack on a continuous or periodic basis. The second option is **parametric monitoring**, which involves indirect measurement of emissions by monitoring key parameters related to the operating status of air pollution control equipment or process equipment. Parametric monitoring requires demonstration that the process or control parameters being monitored correlate to measured pollutant emission levels.

CEM is required for large sources or sources that have monitoring requirements under New Source Review (NSR), New Source Performance Standard (NSPS), National Emissions Standards for Hazardous Air Pollutants (NESHAPS), or other State requirements. CEM is required under some of the EPA regulations for either continual compliance determinations or determination of exceedances of the standards. [1] Parametric monitoring is more frequently used at small emission sources. As a result of the CAM Rule, parametric monitoring is becoming increas-

ingly important. Use of parametric monitoring can provide more flexible and less expensive options for demonstrating compliance for regulated sources.

Selecting the proper monitoring equipment or parametric method involves more than basic cost and performance comparisons. Operational conditions vary from facility to facility for a given source category, making the choice of monitoring equipment unique to each installation. The choice of monitoring system depends on the following [Clarke, 1998] considerations:

- physical/chemical properties of the pollutant and waste gas stream,
- regulatory or permitting limits and any associated reporting requirements,
- location and method of collecting, processing, and disposing of samples,
- calibration and accuracy requirements,
- quality assurance and quality control requirements,
- maintenance requirements, and
- facility safety and management.

This chapter describes cost estimation methods for monitoring equipment used to determine compliance status under the Clean Air Act.

4.2 Continuous Emissions Monitoring Systems

A continuous emission monitoring system(s) (CEMS) is an integrated system that demonstrates source compliance by collecting samples directly from the duct or stack discharging pollutants to the atmosphere. A CEMS consists of all the equipment necessary for the determination of a gas or particulate matter concentration or emission rate. This includes three basic components:

- the sampling and conditioning system,
- the gas analyzers and/or monitors, and
- data acquisition system (DAS) and controller system.

A CEMS can be designed to monitor a single pollutant or multiple pollutants and waste gas stream parameters. Gaseous compounds, particulate matter, opacity, and volumetric flow rate are typically monitored by CEMS. Figure 4.1 depicts a typical CEMS layout for multiple parameter monitoring.

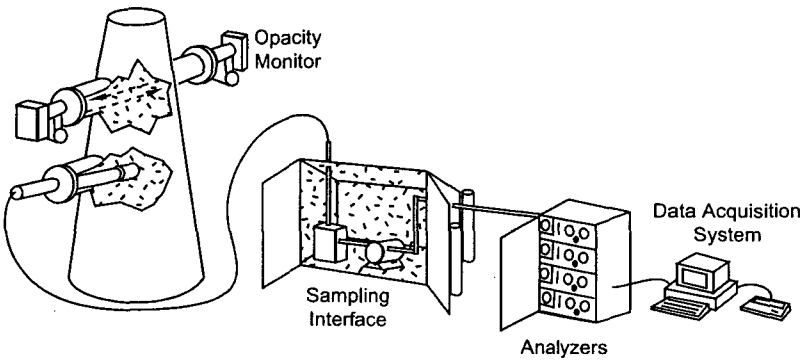


Figure 4.1: Typical CEMS for Multiple Parameter Monitoring

Proper placement of sampling ports in the waste gas stream and proper equipment selection for the components are all critical for the collection of accurate and reproducible information from a CEMS. For this reason, the design of a CEMS is usually based on vendor experience and, therefore, vendor-specific. Most systems are provided on a "turn key" basis where the vendor supplies, installs, and and tests all necessary equipment [18].

EPA has published standard methods for installing, operating and testing CEMS. EPA rules specify the reference methods that are used to substantiate the accuracy and precision of the CEMS. The EPA also maintains performance specifications used for evaluating the acceptability of the CEMS after installation. Finally, the rules provide quality assurance and control procedures to evaluate the quality of data produced by CEMS once in operation [18]. The data produced under these standard or reference methods are direct and enforceable measurements of emissions.

4.2.1 Sampling Systems

CEMS are divided into two major categories, extractive and in situ. **In situ** CEMS typically have monitors and/or analzyers located directly in the stack or duct. **Extractive** CEMS capture a sample from the duct or stack, condition the sample by removing impurities and water, and transport the sample to an analyzer in a remote, environmentally protected area. Some monitoring system designs may employ both types of systems. The two systems are discussed in greater detail in the next sections.

All sampling systems need programmable logic controllers (PLCs) to link the sampling equipment to both monitors and DAS. PLCs are generally modular in design and used widely throughout industry. Typical functions of PLCs are:

- Logic timing
- Counting
- Data transfer
- Triggering automatic functions
- Providing analog to digital signal conversion
- Registering alarms
- Data logging
- Perform mathematical calculations or calibration functions

In CEMS applications, PCLs manage sampling and calibration by controlling solenoid valves that send either waste gas or calibration gas to the monitor. This information is also sent to the DAS to prevent calibration data from inadvertently being used as sample data. PLCs typically control functions such as zero and span checks, alarms for excess emissions or system malfunctions and interfacing with the DAS.

4.2.1.1 Extractive CEMS

In an extractive CEMS, the system extracts a sample at a specified site in the waste gas stream and then transports it to a monitor in an environmentally protected area. This type of system protects the monitoring instrumentation from the high temperatures, high velocities, high pressures, particulate matter, corrosive substances, and water vapor in the waste gas stream.

A sample is transported from the sampling probe location to the analyzer or monitor. In general, the sample requires some form of conditioning prior to analysis. Conditioning can include, filtering of particulate mater, removal of water vapor, cooling of the sample, and dilution of the sample. Extractive systems are generally classified based on the type of conditioning: hot-wet, cool-dry, or dilution. **Hot-wet systems** maintain the sample at high temperature and do not remove water vapor. **Cool-dry systems** lower the sample temperature and remove water vapor. **Dilution systems** sample at low flow rates or dilute the sample prior to analysis which results in lower water vapor and particulate matter content. Conditioning may be performed at the port or at the analyzer. Depending on the type of system, extractive CEMS sampling and conditioning equipment can include sampling probe/port, sampling transfer lines, line heaters, a pump, a filter, a condenser or dryer, and chillers. The choice of sampling system type is application-specific [18.] Figure 4.2 shows a typical extractive system with a cool-dry sampling system.

Extractive analyzers are typically less expensive and easier to maintain and repair than in situ analyzers. This is primarily due to their location in an environmentally controlled room at ground level, rather than at the source. Due to their location they do not require additional environmental protection. In addition, the analyzers are easily accessible to technicians for maintenance and repair. Having an environmentally controlled room also allows the calibration gasses and systems to be located in the same area, which simplifies calibrations.

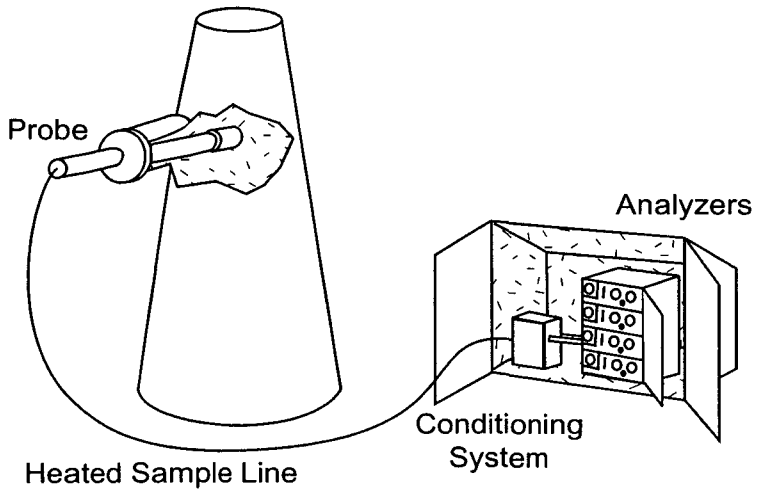


Figure 4.2: Example of an Extractive CEMS with Cool-Dry Sample

However, the advantages of extractive CEMS can be offset by the requirements of the sampling system. Initial costs of sampling systems can be quite high, and sampling and conditioning equipment requires routine maintenance. Other sample handling problems include:

- Probes and lines clogging with contamination,
- Heated lines failing in cold climates causing water to freeze and block lines,
- Probe filter causing loss of pollutant as it passes through the probe media (scrubing),
- Dilution probe causing temperature, pressure, gas density effects, and water droplet evaporation when dilution air is added to the sample gas,
- Water entrainment,
- Leaks in the tubing or elsewhere in the system,
- Adsorption of pollutant to the wall, filter, tubing or other components, and
- Absorption of pollutant to the water which is removed by a conditioning systems.

Other important factors in selection and design of monitoring systems include:

- Regulatory requirements;
- Data availability (% time monitor supplies data)
- Volume of waste gas must which must be collected and conditioned [18].

There are a number of commercially available CEMS monitors and gas analyzers available, including several multi-pollutant analyzers. This manual provides costs for the following types of extractive CEMS given in Table 4.1:

Table 4.1: Pollutant Monitoring Capability for Commercially Available Extractive CEMS

Gaseous Compound Analyzers	Monitors
NO_x	Opacity
SO_2	PM
$COCO_2$	Flow Rate
O_2	
THC	
HCl	

4.2.1.2 In Situ CEMS

In situ CEMS are systems where the analyzer is physically located in the stack or duct. The effluent gas is measured in situ as it flows through a sampling location placed in the stack or duct. Two types of in situ measurements are possible: point (in-stack) and path (cross-stack). **Point** measurements take place at the precise point where the sampling cell is located. **Path** measurements are taken across a given path in the emissions stream. Most path measurements are taken by sending a signal across the stack and reflecting it back to a detector near the source of the signal. The emissions crossing that path are then averaged over a given period of time. Figure 4.3 depicts a typical in situ CEMS.

In situ monitors require durable construction and are generally enclosed in sturdy, sealed cabinets to protect them from extreme temperatures, moisture and corrosive gases. As a result, in situ monitors are generally more expensive than comparable extractive monitors.

The primary advantage of in situ monitors is the location of the monitor in close proximity to the sampling probe, which minimizes the loss of contaminate from leaks, absorption, and adsorption, and also eliminates the need for a complex and costly sampling and conditioning system.

Although in situ analyzers were developed to avoid maintenance and availability problems associated with the sampling systems used in extractive monitoring, some problems remain. Service, maintenance and replacement of in situ analyzers is more difficult than with extractive units due to their locations. Because the concentration of pollutants (especially particulate matter) in a stack is not uniform, placement of the in situ analyzer (like placement of the extractive analyzer's probe) is a critical consideration. The sampling probe can become contaminated or plugged. Although the problem of gas sample transportation to the monitor has been eliminated by in situ placement, the need to take calibration gas to the in situ analyzer has taken its place.

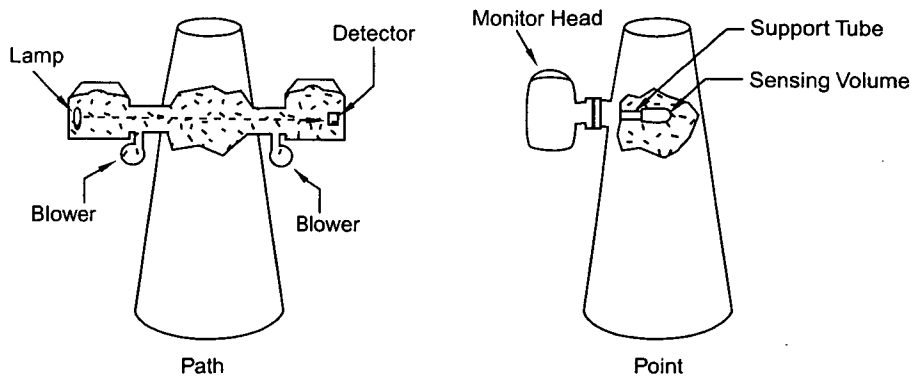


Figure 4.3: Example of an In Situ CEMS with Path and Point Sampling

There are a number of commercially available CEMS monitors and gas analyzers available, including several multi-pollutant analyzers. This manual provides costs for the following types of in situ CEMS given in Table 4.2:

4.2.2 Monitors and Gas Analyzers

A monitor is a device that senses or measures a physical/chemical property of a given substance such as light absorption. The sensor or measuring device generates an electrical output signal. Strip charts and/or computer data acquisition systems record the output signal, which correlates to a pollutant concentration or other parameter (e.g. flow rate) through an equation, graph, or more complicated mathematical relationship. CEMS convert results into units of the applicable emission standard and provide a record (typically a printed chart and/or an electronic data file). Many integrated systems also include a calibration system for gas analyzers that automatically performs and records the required calibrations on a periodic basis (e.g., daily).

Table 4.2: Pollutant Monitoring Capability for Commercially Available In Situ CEMS

Gaseous Analyzers Compound	Monitors
SO_2	Opacity
CO	PM
O_2	Flow Rate
SO_2/NO_x	
$SO_2/NO_x/O_2$	
CO/CO_2	

Older generation gas analyzers produced only a relative measurement (e.g., percent of full scale) that needed to be compared against the calibration gases before the stack concentration could be calculated. Many current generation analyzers can control and integrate calculation data allowing them to read actual stack concentrations on the front of the instrument. These analyzers may also perform data acquisition functions and communicate directly with computers that produce reports. The configurations and installation requirements vary widely between different analyzers and applications.

Critical factors in selecting the type of analyzer or monitor for a particular application include gas concentration, stack and ambient temperatures and the presence of contaminants that could damage or interfere with the sampling or analyzer systems. Other issues such as data availability requirements may influence analyzer selection or drive the need for two analyzers with one in a backup capacity. These issues impact equipment selection and can substantially impact capital, operating, and maintenance costs. As manufactures overcome past limitations, monitors and gas analyzers are becoming more versatile. The selection of a monitor and the cost analysis should be performed on a site-specific basis.

A technical discussion on the types of monitors and gas analyzers that are commercially available for extractive and in situ systems is beyond the scope of this document. Reference [18] provides a detailed technical discussion of gas analyzers and monitors for various types of CEMS and the pollutants and parameters that can be monitored. Table 4.3 and Table 4.4 summarize the various types of monitors that are currently available for extractive and in situ systems including both point and path type monitors. A discussion on selected monitors is provided below.

Fourier Transformation Infared Spectroscopy

Fourier Transformation Infrared Spectroscopy (FTIR) detects compounds based on the absorption of infrared light at critical wavelengths. The amount of absorption is dependant on the molecular bonds present in the waste gas compounds. This absorption creates a unique "fingerprint", or chemical signature, that can be analyzed to determine the compounds present and their concentrations. Current FTIR CEMs can accurately monitor up to six gaseous compounds (SO_2, NO_x, CO, HCl, CO_2, and O_2) various hazardous air pollutants, and volatile organic compounds simultaneously. Figure 4.4 illustrates a simplified schematic of an FTIR Analyzer. [4]

Current FTIR systems are primarily extractive sampling instruments and have similar installation requirements to extractive CEMS. Although FTIR instruments tend to be more expensive than other analyzers, the ability to monitor multiple pollutants with one instrument improves its cost effectiveness. As FTIR CEMS are a relatively new technology, there is little information on their long-term performance. Due to the precision of the instrument, maintenance requirements are high. Maintenance of a FTIR CEMS requires the following:

Table 4.3: Extractive CEMS Gas Analyzers [18]

Absorption Spectroscopic Methods (Infrared/Ultraviolet)	Luminescence Methods	Electro Analysis Methods	Paramagnetic Methods
Spectrophotometry	Fluorescence	Polargraphy	Thermomagnetic
Differential Absorption	Chemiluminescen	Potentiometry	Magnetodynamic
Gas Filter Correlation	Flame Photomete	Electrocatalysis	Magnetopuematic
Fouier Transform Infrared		Amperomatic	
		Conductimetric	

Table 4.4: In-Situ CEMS Gas Analyzers [18]

Gas Analyzers		PM Monitors	
Point	Path	Point	Path
Second Derivative	Differential Absorption	Light Back Scattering Absorption	
Polargraphy	Gas Filter Correlation	Ion Charge	Light Scattering and Absorption
Potentiometry		Nuclear Radiation	
Electrocatalysis			Attenuation

- Technical maintenance personnel
- High priced parts
- Lengthy calibration
- Short frequency

Opacity Monitor

Opacity monitors are in situ path devices based on the principle of transmissometry; the measurement of the transmission of light through a fluid. A light source of known frequency is generated by one of the following devices: LED, incandescent light, or laser. The opacity monitor then detects the decrease in light transmission across the stack due to particulate matter. Light absorption and scattering due to particulate matter in the gas stream is detected at a specified optical wavelength that minimizes absorption by other material in the stack gas. Interference caused by high levels of NO_2 and water droplets can reduce accuracy. An opacity monitor consists of a light source for generating the light, a transmissometer for accurately measuring the transmission of light, an internal reference system for calibration, and a data acquisition system for data collection. [3]

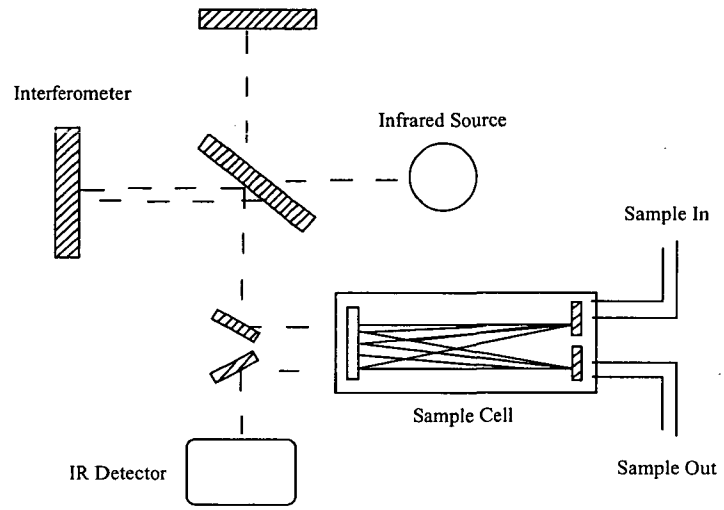


Figure 4.4: Simplified Schematic of an FTIR Analyzer [4]

Particulate Matter Monitor

The standard EPA reference methods for measuring PM are based on flowing a measured volume of waste gas across a particle filter and capturing the PM. The filter is weighed before and after exposure to determine the weight of PM in the measured volume of air. This technique is known as gravimetric measurent.

Particulate matter (PM) monitors are a relatively new technology, and, therefore, make use of newer techniques. Typical approaches include light scattering measurement, transmissometry (see opacity monitors), and other optical and electrostatic techniques. The method that comes closest to the gravimetric method is beta attenuation, where a strip of filter media is exposed to a known volume of the gas stream. The filter media then goes through a beta ray source and detector that measures the attenuation (absorbtion) of the beta source by the PM on the filter. This method is subject to variation due to high beta attenuation of heavy metal in the PM.

CEMS cannot replicate the EPA method, and, therefore, rely on surrogate measures of PM concentration, such as the optical or electrostatic characteristics of the PM in their path. For processes where the PM and other stack characteristics are constant, a calibrated instrument can provide reasonable accuracy. In application such as hazardous waste incinerators, where the gas stream can vary substantially, the potential for inaccuracy increases.

PM monitoring is an advancing technology, and changes in techniques and instrumentation are likely to occur quickly. These changes result in changes in instrument costs. Although a general cost has been provided for PM monitors, this cost is less reliable than the costs of better established technologies such as extractive gas monitors. If more reliable cost inforation is required, a cost estimate should be obtained from a vendor.

4.2.3 Data Acquisition System

Data acquisition systems (DAS) consist of a computer, monitor, printer and software that interface with the monitoring system and provide reports, data storage, and screen displays. Analyzers produce an output signal in volts or milliamps that represents a fraction of the full scale reading established using calibration gases. This output signal typically goes to a strip chart recorder that uses colored pens and paper graph charts to record the analyzers readings. This reading must be interpreted based on the calibration value; for example, if a calibration gas of 10 ppm produced a signal of 10 volts, then a reading of 4 volts corresponds to a concentration of 4 ppm. While many CEMs still include strip chart recorders as back-up systems, most CEMs rely on DAS for data processing and management.

DAS typically include analog to digital conversion boards that take the voltage or milliamperage signal from the analyzer and convert it into digital information that can be understood by a computer. Newer generation monitors have the ability to include calibration information and directly report concentrations; they are also capable of storing data and communication directly to computers with digital information. The computer can also provide controlling functions for the monitors such as performing calibrations, if not provided by a PLC system.

Reporting requirements can have a significant effect on the design of a DAS, and the reporting frequency and averaging time for the monitoring results can impact capacity and cost. However, the growing power of personal computers has improved the functionality and lessened the upper-end costs for DAS, (Table 4.16 in Appendix A shows a range of cost between $16,000 and $20,000). Proprietary software typically comes from the DAS vendor. This software manages data and produces quality assured reports for use by plant personnel and regulatory authorities. Examples of DAS computer program functions include [7]:

- Allowing the operator to interface with the CEMS;
- Averaging data, calculating emissions estimates, and creating reports;
- Providing electronic and hard copies of logs and reports;
- Interfacing with other computer systems.

4.3 Parametric Monitoring

Parametric monitoring differs from CEMS in that emissions are not monitored directly. Parametric monitoring is the monitoring of key, emissions-correlated easurables (such as pressure). Operating parameters are monitored by thermocouples, differential pressure gauges, or

other instrumentation. For example, a thermal treatment device designed to control VOCs demonstrates compliance with a VOC destruction efficiency of greater that 90% as long as a temperature of 1800 °F is maintained by the device. This correlation of temperature to emissions reduction is established thorough periodic monitoring (e.g., annual compliance testing). Parametric monitoring allows the use of temperature monitoring in place of VOC monitoring for this device once the correlation of temperature to VOC destruction has been established.

The use of parametric monitoring can provide more flexible and less expensive options than CEMS for demonstrating compliance of regulated sources. EPA's view of the use of parametric monitoring is expressed in the May 1, 1998 "EPA Draft Final Periodic Monitoring Guidance" document.

Parametric monitoring provides a reasonable assurance of compliance, but the CAM Rule should be consulted for guidance on the type of parametric monitoring that might satisfy periodic monitoring [3]. An additional source of information that includes additional monitoring parameters beyond those used in the CAM Rule is the "Ohio EPA's Operation and Maintenance (O&M) Guidelines for Air Pollution Control Equipment".

When parametric monitoring is used for continuous compliance monitoring, the equipment requirements can be similar to CEMS. Although the gas sampling systems used by emissions monitors are not likely to be components of parametric monitoring systems, some type of calibration and data acquisition systems are likely to be required. The type of process, control equipment and pollutant to be monitored determine the selection of a parametric monitoring system. Data reduction, record-keeping, and reporting are performed independently of sampling in a parametric system, however, they are inherent to regulatory compliance demonstration. For many sources, a combination of parametric monitoring and a data acquisition system is sufficient to comply with CAM. Some forms of parametric monitoring may use the same types of data acquisition, record keeping and reporting as CEMS.

> "When using parametric data to satisfy the periodic monitoring requirement, the permit should specify a range which will assure that the source is in compliance with the underlying requirement. Wherever possible, the proposed range should be supported by documentation indicating a correlation between the parameter(s) and compliance with the emission limit, although it is not required that the range be set such that an excursion from the range will indicate noncompliance with the associated limit. The permit should also include some means of periodically verifying this correlation.
>
> For example, the permit may require periodic stack testing to verify direct compliance with the applicable requirement. At the same time, the test data could be used to set the parameter ranges that will be used to determine compliance between tests.
>
> The permit should also specify what happens when a parameter exceeds the established range. For example, the permit should specify whether excursion from the established range is considered a violation or whether it will instead trigger corrective action and/or additional monitoring or testing requirements to determine the compliance status of the source."

Most monitoring is required to demonstrate compliance with applicable emissions limitations for specific pollutants. Although multiple pollutants may at some times be correlated to the same parameter, in most cases, the parametric monitoring method depends on the pollutant of interest. Sample collection, analysis, and data reduction methods are specific to the type of contaminant or process measurement being monitored.

A brief discussion and examples of parametric monitoring are given in the following sections for a variety of pollutants including PM, SO_2, CO, NO_x, VOCs, and opacity. The costs for parametric monitoring of a single unit is also presented at the end of each section in Tables 4.5 - 4.10. [18]

Cost estimates for parametric monitoring were taken from the supporting information related to the regulatory impact analysis for the CAM rulemaking. The cost estimates contained in this section are not sensitive to the size of the equipment. In general, they represent medium sized units that do not already have applicable monitoring requirements under NSPS or other federal programs. These costs represent monitoring for one control device such as a single thermal unit or baghouse. The costs reported are generic in nature, while the true cost will depend on a number of factors (size for instance). Larger units may have multiple control devices and would require multiple parametric monitoring devices. In addition, larger units typically already have monitoring systems in place. Many of these units are required to upgrade their monitoring under the CAM Rule. Rather than relying solely on the cost estimates provided by this document, an "expert" on the design and choice of parametric monitoring equipment should be consulted to determine the true cost.

4.3.1 Particulate Matter (PM)

The two principal methods of controlling PM emissions currently in use by U.S. industry are electrostatic precipitators (ESPs) and fabric filters, also called baghouses. Parametric monitoring has been used for many years to monitor ESP performance. Items such as gas volume and velocity, temperature, moisture, rapping (cleaning) frequency, and the electrostatic field's voltage and current applied are indicators that can be monitored to assure continued ESP performance. ESPs are typically used by larger sources that may already be subject to NSPS or other monitoring requirements. ESPs are not typically viewed as cost effective control devices for smaller sources.

The CAM Rule used parametric monitoring of a baghouse as its basis for establishing PM parametric monitoring costs. Fabric filtration can be applied to a wide range of sources, from small shot-blast units to large steel mills. This section uses monitoring of the pressure drop across the baghouse as an example of parametric monitoring. A baghouse operates much like a vacuum cleaner with a fan either blowing dirty air through (positive pressure) the filter or drawing air into (negative pressure) the filter. In either case, it takes substantial air pressure to force the air through the filter. The pressure drop is a measurement of this difference in pressure between the clean and

dirty sides of the filter. Static pressure gauges can be installed at the inlet and outlet of the fabric filter to determine the unit's pressure drop. As the fabric becomes clogged with dust there is more resistance to air flow, resulting in an increased pressure drop.

Typically, a baghouse is cleaned in sections, with jets of counter-flowing air used to blow captured dust off the filter and into a hopper. In many installations, the baghouse will follow a routine cycle with the pressure drop increasing as the bag becomes coated with dust, and dropping back to a baseline value after it is cleaned. Pressure drop measurements are used to determine if the fabric is being properly cleaned and that the baghouse is operating as designed. Abnormally high values may indicate that the filter media is becoming "blinded" by materials, such as organic aerosols, that cannot be removed. This is a potential indication of a failure to capture and control the process PM. Abnormally low values may indicate holes in the filter media or mechanical failure of baghouse components. Table 4.5 provides cost estimations for parametric monitoring of PM emissions using pressure drop across the filter fabric.

As with other types of CAM, monitoring of pressure drop is a useful indicator of baghouse performance, but does not guarantee compliance with emission standards. Any parametric monitoring program for fabric filtration control equipment should be considered part of an overall compliance program that includes routine inspections and maintenance logs that help to predict and eliminate equipment problems before they occur. Routine monitoring of the key operating parameters will improve the performance of a fabric filter and increase its effective service life. Establishing an effective operation and maintenance program is an important component of predicting baghouse failures.

There are several other methods for monitoring. PM visual opacity monitoring by certified smoke readers is one method. Other methods include use of PM CEMS which are now on the market. However, PM CEMS are still considered a new technology. These methods are generally more expensive than parametric monitoring of PM.

Another type of PM control that is typically applied to organic aerosols is thermal treatment. Although this is primarily a VOC control technique, it is effective for the control of high molecular weight organic compounds that can condense to form PM. Combustion temperatures are measured by thermocouples installed in thermal treatment units. Temperature measurements can be used to evaluate combustion practices and, if maintained within designated operating ranges, would provide a reasonable level of confidence for compliance with a PM emission limitation. Temperature monitoring of a thermal treatment device does not require installation of additional equipmentexcept possibly for a DAS [9].

4.3.2 Sulfur Dioxide

The two principal methods of control of sulfur dioxide (SO_2) in use in the U.S. today are wet gas scrubbing and spray dryers. Spray dryers are becoming more prevelant on new and start-up installations, but wet gas screbbers are still more widely used overall.

Table 4.5: Cost Summary for Parametric Monitoring of Particulate Emissions Using Pressure Drop Across Fabric Filter.

Item	Total Cost, $
Capital and other initial costs[a]	**1,07 0**
Planning[b]	240
Equipment selection[c]	2,050
Install and calibrate sstem[d]	630
Total capital Investment (TCI)	3,990
Annual Costs, $ /yr	
Operation and maintenance[e]	270
Recordkeeping[f]	2,015
Property taxes, insurance, and administrative[g]	160
Capital recovery (CRF =)[h]	380
Total Annual Cost, $/yr	2,825

[a] Engineer, 32hrs @$30/hr + managerial review 2hrs @$50/hr + $10 telephone charges
[b] Engineer and purchasing agent 4 hrs @$30/hr
[c] Equipment manufacturer cost
[d] In house and contractor combined labor cost of $360
[e] 10% of purchased equipment cost + In house and contractor labor cost of $65
[f] 5 min. per shift 3 shifts per day x (365 daysyr) @$17.50/hr of operator time for managerí s review @$50/hr, 10% of operator time for clerical support @$10/hr and $100 for supplies.
[g] Based on 4% of TCI
[h] CRF = 0.0944 x TCI based on 20 year life and 7% interest.

The CAM Rule used wet scrubbers (gas-absorbers) to determine its RIA SO_2 monitoring costs. Wet scrubbers use a variety of techniques including packing materials, perforated trays, and sprayers to force close contact between the dirty gas and the gas scrubbing liquid (liquor) flowing through the scrubber. One SO_2 parameter used to indirectly monitor emissions is the pressure drop across a wet scrubber measured by a differential pressure gauge or manometer. Similar to our discussion of the baghouse, abnormally high and abnormally low pressure drops can indicate operational problems. Abnormally low pressure drops indicate that the dirty gas is probably not being forced into adequate contact with the scrubber liquor and that SO_2 is probably being released without adequate treatment. Abnormally high pressure drops are likely to indicate mechanical problems such as flooding (excessive liquor) or clogging (contamination of the packing material). These problems indicate failure to adequately capture and control SO_2. The CAM techniques used in this example are generally applicable to other pollutants beyond SO_2. [10]

Monitoring the pressure drop in a gas scrubber is less expensive than using SO_2 CEMS, but it only gives an indication of scrubber operation and is not necessarily an indication of compliance with applicable regulations. For a true indication of compliance, parametric monitoring should be used. Table 4.6 provides cost estimates for parametric monitoring of a wet scrubber using pressure drop.

One of the simplest forms of parametric monitoring is monitoring of fuel sulfur content and fuel useage. Fuel sulfur content is typically available as a maximum specification from the fuel vendor. It can also be sampled on-site and provided as a weight percent sulfur. The molecular weight of SO_2 is twice that of elemental sulfur. Therefore, by monitoring the rate of fuel use, the SO_2 emissions rate can be easily calculated by assuming complete combustion of all fuel sulfur to SO_2. Fuel purchase records may be adequate to monitor fuel use. If this information is not adequate to demonstrate compliance with the applicable standard, fuel monitoring devices can be purchased.

Liquid and gaseous fuels can be monitored using a totalizer, which measures gallons or cubic feet of gas used. Totalizers are available with electronic signals for use with DAS. Solid fuel monitoring could be accomplished by weighing of fuel. Another approach would be to measure the heat output of the equipment. For example, boiler steam output monitored and converted to heat input. A relationship between the fuel required and steam produced for a particular fuel can easily be established for most industrial boilers. Once this relationship is established, steam production could be used as a surrogate for monitoring fuel use.

4.3.3 Carbon Monoxide

Carbon monoxide results from incomplete combustion of carbon based fuels. Some types of combustion equipment, such as incinerators may produce relatively high levels of carbon monoxide. Thermal treatment of the off-gas may be used to burn carbon monoxide and other products of incomplete combustion. Most industrial combustion equipment, including stationary turbines and other stationary engines, produce relatively small amounts of carbon monoxide. For these sources, combustion optimization is the typical control method. Control of key engine operating parameters, such as fuel, air, and engine load, optimizes combustion and lets the engine operate at a low and compliant emissions level. Oxides of nitrogen, VOCs, and other pollutants can also be effectively limited through combustion optimization.

Some industrial combustion equipment requires a fairly narrow set of operating parameters. For this type of equipment, periodic testing can establish an emissions pattern that correlates to optimum operating conditions. The operating conditions that correlate to violations of emissions limitations can be monitored using parametric monitoring techniques. The critical aspect of this type of monitoring is to establish the relationship between operating conditions and emissions. During a periodic compliance test, the key parameters, such as operating temperature, excess air and load can be monitored concurrently with CO. By establishing a correlation between these parameters and CO emissions rates for the range of operating conditions, algorithms can be developed to predict emissions.

These algorithms can be programmed into a DAS. The DAS can then monitor operations and determine if any of the conditions that produce excess emissions occur. Portable combustion analyzers are an acceptable monitoring option for CO sources and can be used to measure excess air or O2, air flow, and temperature.

Table 4.6: Cost Summary for Pressure Drop Across Wet Scrubber

Item	Total Cost, $
Capital and other initial costs	
Planning[a]	4,890
Equipment selection[b]	0
Support facilities[c]	2,000
Purchased equipment cost[d]	3,260
Install and check DAS[e]	5,680
Data collection text[f]	16,140
Total Capital Investment (TCI)	31,970
Annual Costs, $/yr	
Operation and maintenance[g]	900
Annual RATA[h]	10,930
Recordkeeping and reporting[i]	2,020
Property taxes, insurance, and administrative[g]	1,280
Capital recovery[h]	3,020
Total Annual Cost, $/yr	26,650

[a] Based on $ 4,250 labor to review regulations, define monitoring requirements and develop CAM plan plus $640 in supplies.
[b] Cost of selecting PC-based data acquisitions system included in planning costs.
[c] Cost of installing sampling ports in stack.
[d] Cost based on Pentium class PC, monitor, printer, and operating software.
[e] PC installation and interconnection for sensor signals, equipment calibrations and start-up services.
[f] Cost for data collection testing is based on the cost for initial RATA testing on a CEM.
[g] Based on 10% of purchased equipment cost + 10% of installation labor cost.
[h] Cost for data collection testing is based on the cost for annual RATA testing on a CEM.
[i] 5 min. per shift 3 shifts per day x(365 days/yr) @$17.50/hr for operators. Add 2.5% of operator time for engineerí s review @$30/hr, 2.5% of operator time for managerí s review @$50/hr, 10% of operator time for clerical support @$10/hr and $100 for supplies.
[j] QA planning, training, and equipment inventory estimated to be 50% of CEM cost.
[k] Based on 4% of TCI
[l] CRF = 0.0944 x TCI based on 20 year life and 7% interest.

This type approach can be a cost effective manner of parametric monitoring, particularly when several identical units are operated by a company. The costs of developing parametric monitoring techniques for additional identical units should be substantially less than for the first unit.

Cost estimates for the initial development of this type of parametric monitoring of CO on an individual combustion unit are contained in Table 4.7. In this example, portable analyzers are

used for a short period of time to establish a relationship between operating conditions and emissions. The purchased equipment cost is lower than using a CO CEM, however, a data acquisition system is required. For most combustion equipment that operates within a predictable range, this method offers greater assurance of compliance than pressure drop monitoring described in the previous two examples. Some industrial combustion equipment operates at near steady-state conditions, and simpler parametric monitoring may be adequate. Many industrial boilers already monitor operating parameters such as load and combustion airflow using strip chart recorders. Some units may be able to demonstrate that their existing monitoring is adequate to maintain compliance.

4.3.4 Nitrous Oxides

NO_x emissions from industrial combustion equipment can be monitored in the same manner as CO emissions discussed above. NO_x emissions will vary with load and will typically increase as the load increases. To limit NO_x generation, load, combustion zone temperature and excess air need to be minimized. Although the algorithm that describes the relationship between NO_x and operating conditions is obviously going to be different than the one developed for CO, the basic approach is identical. Stationary turbines produce more NO_x than CO and may operate much closer to regulatory limits for NO_x. The parametric monitoring approach may need additional periodic direct testing of NO_x emissions if the margin of compliance is small.[12] Cost estimates for the initial development of parametric monitoring of NO_x on an individual combustion unit are contained in Table 4.8.

4.3.5 Opacity

Opacity regulations are intended to support compliance with PM emissions limitations. Opacity standards can be thought of as surrogate or parametric approaches to determining PM compliance. Opacity can be measured using an opacity monitor or through the use of EPA Methods 9 and 22. It is possible that parametric approaches, such as those discussed for CO and NO_x that rely on correlating operating status of the equipment to emissions rates, can be used. However, for most processes, high opacity is not a typical operation and probably cannot easily be correlated to typical operating parameters.

The CAM Rule proposed EPA Method 9 as a method of establishing compliance with opacity regulations and can also be considered a method or supporting method of verifying compliance with PM emissions limits. Using EPA Method 9, opacity is measured by a certified smoke reader who visually observes the opacity or optical density of the plume. The readers eyes are "calibrated" by undergoing recertification every six months. This method is useful for plants with control devices that normally produce no visible emissions, but when controls fail, visible emissions occur. For example, consider a printing press with a drying oven that produces visible smoke. To eliminate the smoke, thermal combustion control equipment is installed. For this process, any visible emissions are likely to indicate operating problems with the control equipment.

Figure 4.7: Cost Summary for Parametric Monitorying of CO Emissions Using Temperature and other Combution Operating Parameters

Item	Total Cost, $
Capital and other initial costs	
Planning[a]	4,890
Equipment selection[b]	0
Support facilities[c]	2,000
Purchased equipment cost[d]	3,260
Install and check DAS[e]	5,680
Data collection text[f]	16,140
Total Capital Investment (TCI)	31,970
Annual Costs, $/yr	
Operation and maintenance[g]	900
Annual RATA[h]	10,930
Recordkeeping and reporting[i]	2,020
Property taxes, insurance, and administrative[g]	1,280
Capital recovery[h]	3,020
Total Annual Cost, $/yr	26,650

[a] Based on $ 4,250 labor to review regulations, define monitoring requirements and develop CAM plan plus $640 in supplies.
[b] Cost of selecting PC-based data acquisitions system included in planning costs.
[c] Cost of installing sampling ports in stack.
[d] Cost based on Pentium class PC, monitor, printer, and operating software.
[e] PC installation and interconnection for sensor signals, equipment calibrations and start-up services.
[f] Cost for data collection testing is based on the cost for initial RATA testing on a CEM.
[g] Based on 10% of purchased equipment cost + 10% of installation labor cost.
[h] Cost for data collection testing is based on the cost for annual RATA testing on a CEM.
[i] 5 min. per shift 3 shifts per day x(365 days/yr) @$17.50/hr for operators. Add 2.5% of operator time for engineeri s review @$30/hr, 2.5% of operator time for manageri s review @$50/hr, 10% of operator time for clerical support @$10/hr and $100 for supplies.
[j] QA planning, training, and equipment inventory estimated to be 50% of CEM cost.
[k] Based on 4% of TCI
[l] CRF = 0.0944 x TCI based on 20 year life and 7% interest.

Table 4.8: Cost Summary for Parametric Monitoring of NO_x Emissions Using Temperature and other Combustion Parameters

Item	Total Cost, $
Capital and other initial costs	
Planning[a]	4,890
Equipment selection	0
Support facilities[c]	2,000
Purchased equipment cost[d]	3,260
Install and check DAS[e]	5,680
Data collection text[f]	16,140
Total Capital Investment (TCI)	31,970
Annual Costs, $/yr	
Operation and maintenance[g]	900
Annual RATA[h]	10,930
Recordkeeping and reporting[i]	2,020
Property taxes, insurance, and administrative[g]	1,280
Capital recovery[h]	3,020
Total Annual Cost, $/yr	26,650

[a] $4,250 labor to review regulations, define monitorying requirements and develop CAM plan plus $640 in supplies.
[b] Cost of selecting PC-based data acquisitions system included in planning costs.
[c] Cost of installing sampling ports in stack.
[d] Cost based on Pentium class PC, monitor, printer, and operating software.
[e] PC installation and interconnection for sensor signals, equipment calibrations and start-up services.
[f] Cost for data collection testing is based on the cost for initial RATA testing on a CEM.
[g] Based on 10% of purchased equipment cost + 10% of installation labor cost.
[h] Cost for data collection testing is based on the cost for annual RATA testing on a CEM.
[i] 5 min. per shift 3 shifts per day x(365 days/yr) @$17.50/hr for operators. Add 2.5% of operator time for engineerí s review @$30/hr, 2.5% of operator time for managerí s review @$50/hr, 10% of operator time for clerical support @$10/hr and $100 for supplies.
[j] QA planning, training, and equipment inventory estimated to be 50% of CEM cost.
[k] Based on 4% of TCI
[l] CRF = 0.0944 x TCI based on 20 year life and 7% interest.

Most air pollution emissions points are also subject to opacity regulations. For opacity regulations, Methods 9 and 22 are enforceable reference methods and not parametric methods. Opacity reading is less expensive than direct emissions monitoring using CEMs. PM CEMs are now on the market, but are a relatively new technology (see Section 4.2.2). However, opacity reading does have its drawbacks. The presence of water vapor in the stack, the color of smoke emitted, and the position of the sun can substantially influence apparent opacity. In spite of these complicating factors, opacity reading remains in wide use because of the lack of alternative methods for easily determining PM emissions.

Cost estimates for parametric monitoring of opacity using visual opacity readings on an individual unit are contained in Table 4.9.

4.3.6 VOCs

The use of temperature monitoring to assure thermal destruction of organic particles is primarily applied to assuming VOC destruction. Periodic testing, such as a compliance test, establishes the performance of the thermal treatment (e.g., 98% destruction of VOC) at the minimum operating temperature achieved during the test. Provided this temperature is maintained and the type and amount of VOC feed to the thermal unit do not change substantially, the performance of the unit is demonstrated.

In order to evaluate control costs for the CAM Rule, EPA developed a parametric monitoring approach for carbon adsorbers, which are frequently used to abate VOC emissions. Periodic or continuous direct measurement of outlet VOC concentration is one type of parametric monitoring applied to VOC adsorbent control devices. The purpose of this monitoring is to detect "breakthrough" of VOC through the carbon, which occurs when the carbon becomes saturated with VOCs and can longer remove them from the gas stream. VOCs then pass through the carbon uncontrolled. The adsorbtion capacity of the carbon and the VOC concentration in the gas stream help determine an appropriate monitoring approach.

Larger systems typically regenerate the carbon onsite, often many times a day. As a result, the potential for breakthrough is high in these systems, so many other parameters are typically monitored to maintain safety and performance. Measuring the inlet gas temperature and the temperature of the carbon bed can detect potential fires. Monitoring of a pressure drop across the carbon adsorber is an indicator of proper gas flow, carbon bed plugging, or carbon bed channeling. Static pressure gauges, magnehelic gauges, or manometers can be installed at the inlet and outlet to determine pressure drop. Continuous VOC monitoring may also be appropriate, for these systems. If a low resolution VOC monitor is used, VOC monitoring becoms a parametric method rather than a CEM method. The monitor used could be less sensitive and expensive than a VOC CEM since it is only required to detect the VOC concentrations after the carbon absorber has risen to a level that indicates breakthrough [13].

Table 4.9: Cost Summary for Parametric Monitoring of Opacity Using the Visible Emissions Method

Item	Total Cost, $
Capital and other initial costs	
Planning[a]	1,070
Course selection[b]	240
Training Course[c]	550
Certification[d]	1,100
Total Capital Investment (TCI)	2,960
Annual Costs, $/yr	
15 minute daily observation[e]	1,700
Semiannual certification[f]	1,100
Recordkeeping and reporting[g]	2,015
Property taxes, insurance, and administrative	120
Capital recovery	280
Total Annual Cost, $/yr	5,215

[a] Engineer, 32 hrs @$30/hr + managerial review 2hrs @$50/hr + $10 telephone charges.
[b] Engineer and purchasing agent 4 hrs @$30/hr.
[c] One-day training course for two plant operators @$17/hr + $200 to contractor + $50 other costs.
[d] Two days for two operators to pass certification tests @$17/hr.
[e] 15 min. per day opacity observation for operator @$17/hr
[f] 5 min. per shift 3 shifts per day x(365 days/yr) @$17.50/hr for operators. Add 2.5% of operator time for engineeri s review @$30/hr, 2.5% of operator time for manageri s review @$50/hr, 10% of operator time for clerical support @$10/hr and $100 for supplies.
[g] Based on 4% of TCI
[h] CRF = 0.0944 x TCI based on 20 year life and 7% interest.

Smaller systems may not regenerate the carbon onsite. Periodic replacement of the carbon or the entire system are common practices. The system can be as simple as a 55 gallon drum filled with carbon and a hose that can be connected to a source of VOCs (such as a small storage tank). Multiple drums can be stored onsite and switched out when the carbon becomes saturated with VOCs. A recycling vendor can then recycle the used drums, leaving fresh drums as replacements. For these systems, periodic testing with sample tubes may be adequate for detecting when the carbon is saturated and drum replacement is required. This periodic testing can be used to establish a reasonable replacement schedule. Cost estimates for parametric monitoring of VOCs using carbon absorption capacity on an individual unit are contained in Table 4.10.

4.3.7 DAS

The type of recordkeeping used to demonstrate compliance should be reasonably consistent with the size, complexity and regulatory requirements of the source and the source's potential for excess emissions. In the cost summaries presented in the previous sections, a DAS price was only included in the cost estimates for CO and NO_x parametric monitoring. For other examples, such as monitoring the pressure drop across a baghouse, simple manual methods can be adequate; recordkeeping can consist of an operator manually logging the pressure drop once per shift. However, larger sources, or sources with more stringent regulatory requirements, may necessitate the use of a DAS.

The data acquisition systems involved with parametric monitoring do not differ greatly from DAS for CEMS. The need to acquire an electronic signal, then process, store, check, and summarize the signal as a reporting parameter is identical. Some special signal conditioning may

Table 4.10: Cost Summary for Parametric Monitoring of VOCs Using Carbon Absorption Capacity

Item	Cost, $
Capital and other initial costs	
Planning[a]	1,070
Equipment selection[b]	240
Purchased equipment cost[c]	620
Install and calibrate system[d]	630
Total Capital Investment (TCI)	2,960
Annual costs, $/yr	
Operation and maintenance[e]	130
Recordkeeping[ff]	9,795
Property taxes, insurance, and administrative	100
Capital recovery	240
Annual Cost, $/yr	10,265

[a] Engineer, 32 hrs @$30/hr + managerial review 2hrs @$50/hr + $10 telephone charges
[b] Engineer and purchasing agent 4 hrs @$30/hr
[c] Equipment manufacturer/supplier cost
[d] In house and contractor combined labor cost of $630
[e] 10% of purchased equipment cost + In house and contractor combined labor cost of $65
[f] 5 min. per shift 3 shifts per day x(365 days/yr) @$17.50/hr for operators. Add 2.5% of operator time for engineeri s review @$30/hr, 2.5% of operator time for manageri s review @$50/hr, 10% of operator time for clerical support @$10/hr and $100 for supplies.
[g] Based on 4% of TCI
[h] CRF = 0.0944 x TCI based on 20 year life and 7% interest.

be required, however, most DAS are equipped or easily upgraded to handle signals such as temperatures provided by different types of thermocouples. In the CO and NO_x examples, a DAS and computer are used to develop correlations between process parameters and observed emission profiles. In this example, the DAS is essential in acquiring process operating information that is correlated by the computer to an emissions profile.

4.4 Estimating Capital and Annual Costs for CEMS

The U.S. Environmental Protection Agency (EPA) developed a computer software program for estimating the cost of CEMS titled Continuous Emission Monitoring System Cost Model, Version 3.0 (CEMS Cost Model). The CEMS cost estimation methods in this chapter represent a simplified version of this model appropriate for use with the spreadsheets used throughout this manual. With the exception of rounding errors, the costs estimates produced from this method match the values obtained with the CEMS Cost Model.

This approach represents an adequate estimation method for permit engineers verifying equipment costs during permit analysis or for engineers performing initial costs of equipment at typical installations. Total Capital Investment (TCI) and Total Annual Cost (TAC) can be estimated for numerous CEMS configurations, without going to the more complex CEMS Cost Model. The equations provided in this section do not cover all of the scenarios and monitor types and equipment combinations that are available in the CEMS Cost Model.

This methodology estimates study-level costs for a single CEMS to monitor emissions from one source at a facility. The value obtained for a single CEMS should not be multiplied by the number of CEMS required for a multiple source facility since this overestimates the cost of multiple CEMS. A more detailed approach would require consideration of additional factors that impact the accuracy, reliability and cost of installing and maintaining a monitoring system. Detailed cost estimates should rely on the more complete CEM Cost Model along with vendor or other expert analysis of application specific requirements.

4.4.1 Development of Cost Equations

The cost equations for TCI and TAC in this section were developed from the CEMS Cost Model using multiple linear regression techniques. Factors that impacted capital costs, annual costs, personnel cost factors, and equipment cost factors, functioned as variables in the regression analysis. These factors are assigned default values from CEMS Cost Model data.

This manual assumes the necessary personnel to install a CEMS includes a corporate environmental engineer (CEE), two plant technicians, a CEMS consultant, and test personnel. The cost factors associated with these personnel include wages, overhead, travel time, travel fare, per

diem, and fees. The TCI and TAC equations are derived assuming the values given in Table 4.15 located in Appendix A. These assumptions must be considered when determining the applicability of the cost equations. The default values from the CEMS Cost Model for personnel cost factors are supplied in Table 4.11. The data in Table 4.11are fully loaded hourly rates for each employee type. The default values can be modified if location specific or vendor specific information is available (e.g., local labor rates).

The equipment cost factors include the cost of the CEMS monitors and analyzers and auxiliary equipment. The monitor and analyzer costs are specific to the CEMS configuration (Extractive, In Situ, and FTIR) and the pollutant(s) or parameter(s) monitored. Auxiliary costs include the sampling system, DAS equipment, shelter for equipment, and controls. It also includes equipment, such as access ladders and platforms, and both system fabrication and installation. The TCI and TAC equations are derived assuming the values given in 4.16 located in Appendix A. The default values from the CEMS Cost Model for the equipment cost factor is supplied in Table 4.12. The equipment costs presented in Table 4.12 are averages of costs provided by several vendors for development of the CEMS Cost Model. These default values can be modified if vendor specific information is available.

Table 4.11: Default Personnel Hourly Rates and Cost Factors

Cost Item	CEE	Plant Technicial	Plant Technician II	CEMS Consultant	Test Personal
Wage rate, $/hr w/o OH	30.00	18.00	27.00	27.00	16.00
Overhead (OH), % of wage rate	40	40	40	200	200
Fee, % profit	N/A	N/A	N/A	10	10
Hourly Rate[1]	42.00	25.20	37.8	89.1	52.8

[1] Loaded hourly rate, $/hr (wage rate with OH & Fee)

Muli-variable linear regression was performed using the default cost factors to produce regression constants for various CEMS sampling configurations and pollutant monitors. There are unique regression constants for both the TCI and TAC cost equations, which act as "correction factors" for the default values of the cost factors. The set of constants to be utilized in the cost equations is determined by the CEMS design. Design options which are accounted for include:

- Device Type - the CEMS sampling configuration (Extractive, In Situ, and FTIR),
- Parameter Monitored - single pollutant, multiple pollutants, opacity, and flow,
- Pre-control sample - additional sampling location prior to the pollution control device, and

Table 4.12: Default Analyzer and Monitor Equipment Costs for CEMS ($)

Pollutant or Parameter	Extractive	In-situ	FTIR
Gaseous Compound Analyzers			
NO_x	10,440	N/A	N/A
SO_2	12,500	35,000	N/A
CO	8,490	28,000	N/A
CO_2	7,890	N/A	N/A
O_2	5,860	6,600	N/A
THC	10,200	N/A	N/A
HCl	12,390	N/A	N/A
SO_2/NO_x	N/A	37,000	N/A
$SO_2/NO_x/O_2$	N/A	45,000	N/A
CO/CO_2	N/A	34,000	N/A
Monitors[a]			
Opacity	25,000	25,000	25,000
PM	37,700	37,700	37,700
Flow	18,000	18,000	18,000
FTIR analyzer	N/A	N/A	100,000[b]

[a] All CEMS use identical opacity, PM, and flow monitors.

[b] Add $8,000 for capability to monitor before control as well as after control.

- New Installation - installation on a new facility versus retrofit on an existing facility.

The user must first select between an Extractive, In Situ, or FTIR installation, thenselect the pollutant(s) or parameter(s) to be monitored. The equations assume one CEMS sampling location installed downstream ofthe pollution control device. The cost for an additional sampling location prior to the control can be included using the Pre-control sample parameter. The equations assume retrofit installation of the CEMS on an existing facility and correct for the cost of installation on a new facility using the New installation parameter. The regression constant sets are located in Table 4.13 for capital costs and Table 4.14 for annual costs.

Table 4.13: Coefficients for Calculating Total Capital Investment (TCI) for CEMS

Parameter Measured	Pre-Control Sample	Installation	*k1*	*k2* (hrs)	*k3* (hrs)	*k4* (hrs)	*k5* (hrs)	*k6* (hrs)	*k7* (hrs)
Device Type	**Extractive**								
NO_x		X	$88,366	332.5	152.5	0	109.9	90.7	1
NO_x	X	X	$150,130	368.5	248.1	0	120.8	135.0	2
NO_x			$88,634	342.7	167.7	0	109.9	90.7	1
NO_x	X		$150,606	383.1	282.1	0	120.8	135.0	2
HCl	X		$88,866	332.5	152.5	0	109.9	95.7	1
HCl	X	X	$150,630	368.5	248.1	0	120.8	140.0	2
HCl			$89,134	342.7	167.7	0	109.9	95.7	1
HCl	X		$151,106	383.1	282.1	0	120.8	140.0	2
CO_2		X	$88,280	261.5	152.5	0	109.9	90.7	1
CO_2	X	X	$150,037	293.0	248.1	0	120.8	135.0	2
CO_2			$88,548	272.5	167.7	0	109.9	90.7	1
CO_2	X		$150,513	308.0	282.1	0	120.8	135.0	2
Flow		X	$22,470	192.1	98.5	0	62.7	42.0	1
Flow	X	X	$25,095	205.5	128.8	0	69.1	43.2	2
Flow			$22,638	199.1	100.5	0	62.7	42.0	1
Flow	X		$25,371	214.6	131.6	0	69.1	43.2	2
Opacity		X	$22,033	192.1	98.5	0	62.7	6.0	1
Opacity	X	X	$24,657	205.5	128.8	0	69.1	7.2	2
Opacity			$22,201	199.1	100.5	0	62.7	6.0	1
Opacity	X		$24,933	214.6	131.6	0	69.1	7.2	2
CO		X	$88,366	332.5	152.5	0	109.9	90.7	1
CO	X	X	$150,130	368.5	248.1	0	120.8	135.0	2
CO			$88,634	342.7	167.7	0	109.9	90.7	1
CO	X		$150,606	383.1	282.1	0	120.8	135.0	2
SO_2		X	$88,366	332.5	152.5	0	109.9	90.7	1
SO_2	X	X	$150,130	368.5	248.1	0	120.8	135.0	2
SO_2			$88,634	342.7	167.7	0	109.9	90.7	1
SO_2	X		$150,606	383.1	282.1	0	120.8	135.0	2
O_2		X	$88,280	261.5	152.5	0	109.9	90.7	1
O_2	X	X	$150,037	293.0	248.1	0	120.8	135.0	2
O_2			$88,548	272.5	167.7	0	109.9	90.7	1
O_2	X		$150,513	308.0	282.1	0	120.8	135.0	2
PM		X	$28,855	211.2	153.9	0	64.7	27.1	1
PM	X	X	$36,482	224.9	200.4	0	71.1	28.6	2
PM			$29,223	218.2	155.9	0	64.7	27.1	1
PM	X		$37,158	234.0	203.2	0	71.1	28.6	2
THC		X	$85,086	332.9	152.7	0	109.9	93.2	1
THC	X	X	$143,350	369.3	248.5	0	120.8	137.5	2
THC			$85,354	343.1	167.9	0	109.9	93.2	1
THC	X		$143,826	383.9	282.5	0	120.8	137.5	2
Device Type	**In-Situ**								
CO/CO_2		X	$39,228	288.1	101.0	0	105.1	91.8	1
CO/CO_2	X	X	$45,992	328.8	151.9	0	122.1	137.2	2
CO/CO_2			$39,501	298.3	108.6	0	105.1	91.8	1
CO/CO_2	X		$46,479	343.0	167.5	0	122.1	137.2	2
CO		X	$38,028	283.8	97.4	0	105.1	91.8	1
CO	X	X	$43,592	320.3	144.7	0	122.1	137.2	2
CO			$38,301	294.0	105.0	0	105.1	91.8	1
CO	X		$44,079	334.5	160.3	0	122.1	137.2	2
SO_2		X	$38,028	283.8	97.4	0	105.1	91.8	1
SO_2	X	X	$43,592	320.3	144.7	0	122.1	137.2	2
SO_2			$38,301	294.0	105.0	0	105.1	91.8	1
SO_2	X		$44,079	334.5	160.3	0	122.1	137.2	2
O_2		X	$38,028	287.0	97.4	0	105.1	91.8	1
O_2	X	X	$43,592	323.5	144.7	0	122.1	137.2	2
O_2			$38,301	298.0	105.0	0	105.1	91.8	1
O_2	X		$44,079	338.5	160.3	0	122.1	137.2	2

Table 4.13: Coefficients for Calculating Total Capital Investment (TCI) for CEMS (Cont.)

Parameter Measured	Pre-Control Sample	Installation	*k1*	*k2* (hrs)	*k3* (hrs)	*k4* (hrs)	*k5* (hrs)	*k6* (hrs)	*k7* (hrs)
Device Type	**Extractive**								
Flow		X	$25,875	253.5	98.6	0	64.3	42.0	0.367
Flow	X	X	$32,737	290.6	158.8	0	71.2	86.4	0.733
Flow			$26,049	260.5	100.6	0	64.3	42.0	0.367
Flow	X		$33,223	302.9	167.2	0	71.2	86.4	0.733
SO_2/NO_x		X	$39,228	289.7	101.0	0	105.1	91.8	1
SO_2/NO_x	X	X	$45,992	330.4	151.9	0	122.1	137.2	2
SO_2/NO_x			$39,501	300.3	108.6	0	105.1	91.8	1
SO_2/NO_x	X		$46,479	345.0	167.5	0	122.1	137.2	2
$SO_2/NO_x/O_2$		X	$40,428	293.9	104.6	0	105.1	91.8	1
$SO_2/NO_x/O_2$	X	X	$48,392	338.9	159.1	0	122.1	137.2	2
$SO_2/NO_x/O_2$			$40,701	304.5	112.2	0	105.1	91.8	1
$SO_2/NO_x/O_2$	X		$48,879	353.5	174.7	0	122.1	137.2	2
Device Type	**FTIR**								
NO_x		X	$168,674	352.5	77.6	109.0	109.9	91.6	0
NO_x	X	X	$226,296	376.2	108.4	131.8	120.8	135.6	0
NO_x			$168,966	363.5	71.6	109.0	109.9	91.6	0
NO_x	X		$226,788	391.2	98.4	131.8	120.8	135.6	0
SO_2		X	$168,674	352.5	77.6	109.0	109.9	91.6	0
SO_2	X	X	$226,296	376.2	108.4	131.8	120.8	135.6	0
SO_2			$168,966	363.5	71.6	109.0	109.9	91.6	0
SO_2	X		$226,788	391.2	98.4	131.8	120.8	135.6	0
CO		X	$168,674	352.5	77.6	109.0	109.9	91.6	0
CO	X	X	$226,296	376.2	108.4	131.8	120.8	135.6	0
CO			$168,966	363.5	71.6	109.0	109.9	91.6	0
CO	X		$226,788	391.2	98.4	131.8	120.8	135.6	0
HCl		X	$168,674	352.5	77.6	109.0	109.9	91.6	0
HCl	X	X	$226,296	376.2	108.4	131.8	120.8	135.6	0
HCl			$168,966	363.5	71.6	109.0	109.9	91.6	0
HCl	X		$226,788	391.2	98.4	131.8	120.8	135.6	0
CO_2		X	$168,674	281.6	77.6	109.0	109.9	91.6	0
CO_2	X	X	$176,931	283.8	79.6	121.0	120.8	92.4	0
CO_2			$168,966	292.6	71.6	109.0	109.9	91.6	0
CO_2	X		$177,223	294.8	73.6	121.0	120.8	92.4	0
O_2		X	$168,674	281.6	77.6	109.0	109.9	91.6	0
O_2	X	X	$176,931	283.8	79.6	121.0	120.8	92.4	0
O_2			$168,966	292.6	71.6	109.0	109.9	91.6	0
O_2	X		$177,223	294.8	73.6	121.0	120.8	92.4	0
Flow		X	$184,793	301.3	115.1	89.0	62.7	72.0	1
Flow	X	X	$236,742	332.1	171.3	100.7	69.1	116.4	2
Flow			$184,993	312.3	109.1	89.0	62.7	72.0	1
Flow	X		$237,250	348.4	162.1	100.7	69.1	116.4	2

Table 4.13: Coefficients for Calculating Total Capital Investment (TCI) for CEMS

Parameter Measured	Pre-Control Sample	Installation	*k1*	*k2* (hrs)	*k3* (hrs)	*k4* (hrs)	*k5* (hrs)	*k6* (hrs)	*k7* (hrs)
Device Type	**Extractive**								
NO_x		X	$88,366	332.5	152.5	0	109.9	90.7	1
NO_x	X	X	$150,130	368.5	248.1	0	120.8	135.0	2
NO_x			$88,634	342.7	167.7	0	109.9	90.7	1
NO_x	X		$150,606	383.1	282.1	0	120.8	135.0	2
HCl	X		$88,866	332.5	152.5	0	109.9	95.7	1
HCl	X	X	$150,630	368.5	248.1	0	120.8	140.0	2
HCl			$89,134	342.7	167.7	0	109.9	95.7	1
HCl	X		$151,106	383.1	282.1	0	120.8	140.0	2
CO_2		X	$88,280	261.5	152.5	0	109.9	90.7	1
CO_2	X	X	$150,037	293.0	248.1	0	120.8	135.0	2
CO_2			$88,548	272.5	167.7	0	109.9	90.7	1
CO_2	X		$150,513	308.0	282.1	0	120.8	135.0	2
Flow		X	$22,470	192.1	98.5	0	62.7	42.0	1
Flow	X	X	$25,095	205.5	128.8	0	69.1	43.2	2
Flow			$22,638	199.1	100.5	0	62.7	42.0	1
Flow	X		$25,371	214.6	131.6	0	69.1	43.2	2
Opacity		X	$22,033	192.1	98.5	0	62.7	6.0	1
Opacity	X	X	$24,657	205.5	128.8	0	69.1	7.2	2
Opacity			$22,201	199.1	100.5	0	62.7	6.0	1
Opacity	X		$24,933	214.6	131.6	0	69.1	7.2	2
CO		X	$88,366	332.5	152.5	0	109.9	90.7	1
CO	X	X	$150,130	368.5	248.1	0	120.8	135.0	2
CO			$88,634	342.7	167.7	0	109.9	90.7	1
CO	X		$150,606	383.1	282.1	0	120.8	135.0	2
SO_2		X	$88,366	332.5	152.5	0	109.9	90.7	1
SO_2	X	X	$150,130	368.5	248.1	0	120.8	135.0	2
SO_2			$88,634	342.7	167.7	0	109.9	90.7	1
SO_2	X		$150,606	383.1	282.1	0	120.8	135.0	2
O_2		X	$88,280	261.5	152.5	0	109.9	90.7	1
O_2	X	X	$150,037	293.0	248.1	0	120.8	135.0	2
O_2			$88,548	272.5	167.7	0	109.9	90.7	1
O_2	X		$150,513	308.0	282.1	0	120.8	135.0	2
PM		X	$28,855	211.2	153.9	0	64.7	27.1	1
PM	X	X	$36,482	224.9	200.4	0	71.1	28.6	2
PM			$29,223	218.2	155.9	0	64.7	27.1	1
PM	X		$37,158	234.0	203.2	0	71.1	28.6	2
THC		X	$85,086	332.9	152.7	0	109.9	93.2	1
THC	X	X	$143,350	369.3	248.5	0	120.8	137.5	2
THC			$85,354	343.1	167.9	0	109.9	93.2	1
THC	X		$143,826	383.9	282.5	0	120.8	137.5	2
Device Type	**In-Situ**								
CO/CO_2		X	$39,228	288.1	101.0	0	105.1	91.8	1
CO/CO_2	X	X	$45,992	328.8	151.9	0	122.1	137.2	2
CO/CO_2			$39,501	298.3	108.6	0	105.1	91.8	1
CO/CO_2	X		$46,479	343.0	167.5	0	122.1	137.2	2
CO		X	$38,028	283.8	97.4	0	105.1	91.8	1
CO	X	X	$43,592	320.3	144.7	0	122.1	137.2	2
CO			$38,301	294.0	105.0	0	105.1	91.8	1
CO	X		$44,079	334.5	160.3	0	122.1	137.2	2
SO_2		X	$38,028	283.8	97.4	0	105.1	91.8	1
SO_2	X	X	$43,592	320.3	144.7	0	122.1	137.2	2
SO_2			$38,301	294.0	105.0	0	105.1	91.8	1
SO_2	X		$44,079	334.5	160.3	0	122.1	137.2	2
O_2		X	$38,028	287.0	97.4	0	105.1	91.8	1
O_2	X	X	$43,592	323.5	144.7	0	122.1	137.2	2
O_2			$38,301	298.0	105.0	0	105.1	91.8	1
O_2	X		$44,079	338.5	160.3	0	122.1	137.2	2

Table 4.14: Coefficients for Calculating Total Annual Costs (TAC) for CEMS (Cont.)

Parameter Measured	Pre-Control Sample	Installation	k8 (hrs)	k9 (hrs)	k10 (hrs)	k11 (hrs)	k12 (hrs)	k13 (hrs)	k14 (hrs)
Device Type	***Extractive***								
Flow		X	$1,875	26.4	485.1	0	1.7	42.5	0.05
Flow	X	X	$2,054	36.3	854.5	0	1.8	79.0	0.1
Flow			$1,875	26.4	485.1	0	1.7	42.5	0.05
Flow	X		$2,054	36.3	854.5	0	1.8	79.0	0.1
SO_2/NO_x		X	$4,948	48.0	502.5	0	1.8	77.9	0.05
SO_2/NO_x	X	X	$6,257	61.5	795.4	0	2.0	115.3	0.1
SO_2/NO_x			$4,948	48.0	502.5	0	1.8	77.9	0.05
SO_2/NO_x	X		$6,257	61.5	795.4	0	2.0	115.3	0.1
$SO_2/NO_x/O_2$		X	$4,948	52.3	598.5	0	1.9	78.4	0.05
$SO_2/NO_x/O_2$	X	X	$6,257	70.1	987.4	0	2.2	116.3	0.1
$SO_2/NO_x/O_2$			$4,948	52.3	598.5	0	1.9	78.4	0.05
$SO_2/NO_x/O_2$	X		$6,257	70.1	987.4	0	2.2	116.3	0.1
Device Type	**FTIR**								
NO_x		X	$22,375	35.5	30.2	301.2	1.7	76.9	0
NO_x	X	X	$24,861	41.7	36.2	439.9	1.8	113.9	0
NO_x			$22,375	35.5	30.2	301.2	1.7	76.9	0
NO_x	X		$24,861	41.7	36.2	439.9	1.8	113.9	0
SO_2		X	$22,375	35.5	30.2	301.2	1.7	76.9	0
SO_2	X	X	$24,861	41.7	36.2	439.9	1.8	113.9	0
SO_2			$22,375	35.5	30.2	301.2	1.7	76.9	0
SO_2	X		$24,861	41.7	36.2	439.9	1.8	113.9	0
CO		X	$22,375	35.5	30.2	301.2	1.7	76.9	0
CO	X	X	$24,861	41.7	36.2	439.9	1.8	113.9	0
CO			$22,375	35.5	30.2	301.2	1.7	76.9	0
CO	X		$24,861	41.7	36.2	439.9	1.8	113.9	0
HCl		X	$22,375	35.5	30.2	301.2	1.7	76.9	0
HCl	X	X	$24,861	41.7	36.2	439.9	1.8	113.9	0
HCl			$22,375	35.5	30.2	301.2	1.7	76.9	0
HCl	X		$24,861	41.7	36.2	439.9	1.8	113.9	0
CO_2		X	$22,375	33.5	30.2	300.1	1.7	74.7	0
CO_2	X	X	$24,674	39.7	30.8	435.0	1.8	75.4	0
CO_2			$22,375	33.5	30.2	300.1	1.7	74.7	0
O_2	X		$24,674	39.7	30.8	435.0	1.8	75.4	0
O_2		X	$22,375	33.5	30.2	300.1	1.7	74.7	0
O_2	X	X	$24,674	39.7	30.8	435.0	1.8	75.4	0
O_2			$22,375	33.5	30.2	300.1	1.7	74.7	0
O_2	X		$24,674	39.7	30.8	435.0	1.8	75.4	0
Flow		X	$2,616	27.1	397.6	21.2	0.0	34.0	0.05
Flow	X	X	$2,913	32.3	666.8	25.6	0.0	70.0	0.1
Flow			$2,616	27.1	397.6	21.2	0.0	34.0	0.05
Flow	X		$2,913	32.3	666.8	25.6	0.0	70.0	0.1

4.2 Total Capital Investment

Total Capital Investment (TCI) includes direct and indirect costs associated with purchasing and installing equipment. Costs include the equipment cost, which can be composed of the following components: CEM sampling system cost, monitor cost, DAS cost, auxiliary equipment cost, and both direct and indirect installation costs. The estimate includes costs associated with planning for the CEMS, equipment selection, purchase, installation, support facilities, performance

Table 4.14: Coefficients for Calculating Total Annual Costs (TAC) for CEMS

Parameter Measured	Pre-Control Sample	Installation	k8 (hrs)	k9 (hrs)	k10 (hrs)	k11 (hrs)	k12 (hrs)	k13 (hrs)	k14 (hrs)
Device Type	***Extractive***								
NO_x		X	$3,860	44.2	390.3	0	1.7	76.9	0.1
NO_x	X	X	$5,110	50.8	548.9	0	1.8	113.9	0.2
NO_x			$3,860	44.2	390.3	0	1.7	76.9	0.1
NO_x	X		$5,110	50.8	548.9	0	1.8	113.9	0.2
HCl		X	$4,360	44.2	390.3	0	1.7	80.9	0.1
HCl	X	X	$5,610	50.8	548.9	0	1.8	117.9	0.2
HCl			$4,360	44.2	390.3	0	1.7	80.9	0.1
HCl	X		$5,610	50.8	548.9	0	1.8	117.9	0.2
CO_2		X	$3,860	42.2	389.2	0	1.7	74.7	0.1
CO_2	X	X	$5,110	48.8	547.6	0	1.8	111.4	0.2
CO_2			$3,860	42.2	389.2	0	1.7	74.7	0.1
CO_2	X		$5,110	48.8	547.6	0	1.8	111.4	0.2
Flow		X	$1,655	22.1	386.6	0	0.0	34.0	0.05
Flow	X	X	$1,885	27.3	652.1	0	0.0	34.0	0.1
Flow			$1,655	22.1	386.6	0	0.0	34.0	0.05
Flow	X		$1,885	27.3	652.1	0	0.0	34.0	0.1
Opacity		X	$1,218	22.1	386.6	0	0.0	0.0	0.05
Opacity	X	X	$1,448	27.3	652.1	0	0.0	0.1	0.1
Opacity			$1,218	22.1	386.6	0	0.0	0.0	0.05
Opacity	X		$1,448	27.3	652.1	0	0.0	0.1	0.1
CO		X	$3,860	44.2	390.3	0	1.7	76.9	0.1
CO	X	X	$5,110	50.8	548.9	0	1.8	113.9	0.2
CO			$3,860	44.2	390.3	0	1.7	76.9	0.1
CO	X		$5,110	50.8	548.9	0	1.8	113.9	0.2
SO_2		X	$3,860	44.2	390.3	0	1.7	76.9	0.1
SO_2	X	X	$5,110	50.8	548.9	0	1.8	113.9	0.2
SO_2			$3,860	44.2	390.3	0	1.7	76.9	0.1
SO_2	X		$5,110	50.8	548.9	0	1.8	113.9	0.2
O_2		X	$3,860	42.2	389.2	0	1.7	74.7	0.1
O_2	X	X	$5,110	48.8	547.6	0	1.8	111.4	0.2
O_2			$3,860	42.2	389.2	0	1.7	74.7	0.1
O_2	X		$5,110	48.8	547.6	0	1.8	111.4	0.2
PM		X	$2,723	32.7	521.4	0	1.5	89.3	0.1
PM	X	X	$2,953	37.9	861.5	0	1.5	89.3	0.2
PM			$2,723	32.7	521.4	0	1.5	89.3	0.1
PM	X		$2,953	37.9	861.5	0	1.5	89.3	0.2
THC		X	$4,060	44.2	390.8	0	1.7	78.9	0.1
THC	X	X	$5,310	50.8	549.8	0	1.8	115.9	0.2
THC			$4,060	44.2	390.8	0	1.7	78.9	0.1
THC	X		$5,310	50.8	549.8	0	1.8	115.9	0.2
Device Type	**In-Situ**								
CO/CO_2		X	$4,948	48.0	502.3	0	1.8	77.6	0.05
CO/CO_2	X	X	$6,257	61.5	795.2	0	2.0	115.0	0.1
CO/CO_2			$4,948	48.0	502.3	0	1.8	77.6	0.05
CO/CO_2	X		$6,257	61.5	795.2	0	2.0	115.0	0.1
CO		X	$4,948	43.7	406.3	0	1.7	77.1	0.05
CO	X	X	$6,257	52.9	603.2	0	1.8	114.0	0.1
CO			$4,948	43.7	406.3	0	1.7	77.1	0.05
CO	X		$6,257	52.9	603.2	0	1.8	114.0	0.1
SO_2		X	$4,948	43.7	406.3	0	1.7	77.1	0.05
SO_2	X	X	$6,257	52.9	603.2	0	1.8	114.0	0.1
SO_2			$4,948	43.7	406.3	0	1.7	77.1	0.05
SO_2	X		$6,257	52.9	603.2	0	1.8	114.0	0.1
O_2		X	$4,948	41.7	405.2	0	1.7	74.9	0.05
O_2	X	X	$6,257	50.9	602.1	0	1.8	111.8	0.1
O_2			$4,948	41.7	405.2	0	1.7	74.9	0.05
O_2	X		$6,257	50.9	602.1	0	1.8	111.8	0.1

The Capital Recovery Factor, *CRF*, in Equation 4.2 can be calculated from the following equation:

$$CRF = \frac{i\,(1+i)^n}{\left[(1+i)^n - 1\right]} \tag{4.3}$$

where

i = interest rate (e.g., i = 0.07 for a 7% interest rate)
n = equipment life (in number of years)

For CEMS systems, the agency typically assumes an equipment life of 10 years.

4.5 Sample Calculation

What is the cost for a Extractive SO_2 gas analyzer on a new facility with sampling locations before and after the control device? Assume an interest rate of 7% and that the monitor has a 10-year life.

Step 1: Calculate Total Capital Investment, *TCI*, from Equation 4.1:

$$TCI = k1 + (k2 \times A) + (k3 \times B) + (k4 \times K) + (k5 \times D) + (k6 \times E) + (k7 \times F)$$

Loaded labor rates from Table 4.11:

A = CEE Rate = \$42.0/hr
B = Plant Technician I Rate = \$25.2/hr
C = Plant Technician II Rate = \$37.8/hr
D = Consultant Rate = \$89.1/hr
E = Test Crew Rate = \$52.8/hr

Equipment Cost from Table 4.12:

F = Equipment cost for an Extractive SO_2 CEMS = \$12,500

Coefficients k_1, k_2, k_3, k_4, k_5, k_6 and k_7 from Table 4.13:

k_1 = \$150,130
k_2 = 368.5 hrs

testing (Functional Acceptance Test), and quality assurance evaluations. Finally, the TCI includes the installation of any required platforms & ladders for routine access and service. TCI is calculated from the following equation:

$$TCI = k_1 + (k_2 \times A) + (k_3 \times B) + (k_4 \times C) + (k_5 \times D) + (k_6 \times E) + (k_7 \times F) \quad (4.1)$$

where k_1 through k_7 are the regression constants for capital costs given in Table 4.13. The cost factor variables A through F are the personnel and equipment cost factors as defined below:

A = CEE hourly cost (includes Rate, Overhead, and Fee)
B = Plant Technician hourly cost (includes Rate, Overhead, and Fee)
C = Plant Technician II hourly cost (includes Rate, Overhead, and Fee)
D = CEMS Consultant hourly cost (includes Rate, Overhead, and Fee)
E = Test Crew hourly cost (includes Rate, Overhead, and Fee)
F = Cost of Equipment

Default values for personnel cost factors A through E are given in Table 4.11. Default values for the equipment cost factor, F, are given in Table 4.12.

4.4.3 Total Annual Costs

Total annual cost (TAC) is the sum of the annual direct and indirect costs. Direct annual costs include variable, semi-variable, and fixed costs. Variable direct annual costs account for purchase of calibration gas, water, and electrical power or other consumables required by the CEMS. Fixed and semi-variable direct annual costs include operating and supervisory labor cost, maintenance cost, and equipment replacement cost. In general, indirect annual costs include the capital recovery cost, property taxes, insurance, administrative charges, and overhead. Capital recovery cost is based on the anticipated equipment lifetime and the annual interest rate employed. Equipment lifetime of 10 years is typical for CEMS. *TAC* is calculated from the following equation:

$$TCI = k1 + (k2 \times A) + (k3 \times B) + (k4 \times K) + (k5 \times D) + (k6 \times E) + (k7 \times F) \quad (4.2)$$

where *k8* through *k14* are the regression constants for annual cost given in Table 4.14 and A through F are the default cost factors given in Tables 4.11 and 4.12 as defined in the capital cost section. *TCI* is the total capital cost as calculated in the previous section and *CRF* is the Capital Recovery Factor.

k_{10} = 548.9 hrs
k_{11} = 0 hrs
k_{12} = 1.8 hrs
k_{13} = 113.9 hrs
k_{14} = 0.2

From Step 1, TCI = $214,750. Substituting these values into equation 4.2 gives:

$$TAC = \$5{,}110 + \$2{,}134 + \$13{,}832 + \$0 + \$160 + \$6{,}014 + \$2{,}500 + (CRF \times TCI)$$
$$= \$29{,}750 + (0.1424 \times \$214{,}750)$$
$$= \$60{,}330$$

The total capital investment is $214,750 and the total annual cost is $60,330 for a SO_2 extractive CEMS with sampling locations before and after the control device.

4.6 Acknowledgements

We gratefully acknowledge the following companies for contributing data to this section:

- CiSCO Systems
- Monitor Labs
- Analect Instruments

References

[1] Continuous Emissions Monitoring Systems Market Analysis, October 1994 p. 1-1

[2] Technical Support Document for the Regulatory Impact Analysis of the Enhanced Monitoring Rule. Mathtech, Inc. September 30,1993, p 2-23.

[3] Draft Periodic Monitoring Guidance, May 11, 1998

[3a] Continuous Emissions Monitoring Systems Market Analysis, October 1994, p2-1.

[4] User Manual: United States Environmental Protection Agency's Continuous Emissions Monitoring System Cost Model Version 3.0, p 3-2.

[5] Continuous Emissions Monitoring Systems Market Analysis, October 1994, p2-6.

k_3 = 248.1 hrs
k_4 = 0 hrs
k_5 = 120.8 hrs
k_6 = 135.0 hrs
k_7 = 2

Substituting these values into equation 4.1 gives:

$$TCI = \$150{,}130 + \$15{,}477 + \$6{,}252 + \$0 + \$10{,}763 + \$7{,}128 + \$25{,}000$$
$$= \$214{,}750$$

Step 2: Calculate Capital Recovery Factor, *CRF*, from Equation 4.3:

$$CRF = \frac{0.07 \times (1 + 0.07)^{10}}{\left[(1 + 0.07)^{10} - 1\right]}$$

$$CRF = 0.1424$$

Step 3: Calculate Total Annual Cost, *TAC* from Equation 4.2:

$$TAC = k_8 + (k_9 \times A) + (k_{10} \times B) + (k_{11} \times C) + (k_{12} \times D) + (k_{13} \times E) + (k_{14} \times F) + (CRF \times TCI)$$

Loaded labor rates from Table 4.11:

A = CEE Rate = $42.0/hr
B = Plant Technician I Rate = $25.2/hr
C = Plant Technician II Rate = $37.8/hr
D = Consultant Rate = $89.1/hr
E = Test Crew Rate = $52.8/hr

Equipment Cost from Table 4.12:

F = Equipment cost for an Extractive SO_2 monitor = $12,500

Coefficients k_8, k_9, k_{10}, k_{11}, k_{12}, k_{13} and k_{14} from Table 4.14:

k_8 = $5,110
k_9 = 50.8 hrs

k_{10} = 548.9 hrs
k_{11} = 0 hrs
k_{12} = 1.8 hrs
k_{13} = 113.9 hrs
k_{14} = 0.2

From Step 1, *TCI* = $214,750. Substituting these values into equation 4.2 gives:

$$TAC = \$5{,}110 + \$2{,}134 + \$13{,}832 + \$0 + \$160 + \$6{,}014 + \$2{,}500 + (CRF \times TCI)$$
$$= \$29{,}750 + (0.1424 \times \$214{,}750)$$
$$= \$60{,}330$$

The total capital investment is $214,750 and the total annual cost is $60,330 for a SO_2 extractive CEMS with sampling locations before and after the control device.

4.6 Acknowledgements

We gratefully acknowledge the following companies for contributing data to this section:

- CiSCO Systems
- Monitor Labs
- Analect Instruments

References

[1] Continuous Emissions Monitoring Systems Market Analysis, October 1994 p. 1-1

[2] Technical Support Document for the Regulatory Impact Analysis of the Enhanced Monitoring Rule. Mathtech, Inc. September 30,1993, p 2-23.

[3] Draft Periodic Monitoring Guidance, May 11, 1998

[3a] Continuous Emissions Monitoring Systems Market Analysis, October 1994, p2-1.

[4] User Manual: United States Environmental Protection Agency's Continuous Emissions Monitoring System Cost Model Version 3.0, p 3-2.

[5] Continuous Emissions Monitoring Systems Market Analysis, October 1994, p2-6.

[6] User Manual: United States Environmental Protection Agency's Continuous Emissions Monitoring System Cost Model Version 3.0, p 3-3.

[7] User Manual: United States Environmental Protection Agency's Continuous Emissions Monitoring System Cost Model Version 3.0, p 3-4.

[8] User Manual: United States Environmental Protection Agency's Continuous Emissions Monitoring System Cost Model Version 3.0, p 3-4, 3-5.

[9] Ohio EPA's Operation and Maintenance (O&M) Guidelines for Air Pollution Control Equipment, 1993, p 4-22,4-23.

[10] Ohio EPA's Operation and Maintenance (O&M) Guidelines for Air Pollution Control Equipment, 1993, p 9-13.

[11] Periodic Monitoring Technical Reference Document, April 1999, p 5-26.

[12] Periodic Monitoring Technical Reference Document, April 1999, p 5-23.

[13] Periodic Monitoring Technical Reference Document, April 1999, p 5-11.

[14] Ohio EPA's Operation and Maintenance (O&M) Guidelines for Air Pollution Control Equipment, 1993, p 6-22, 6-23.

[15] Ohio EPA's Operation and Maintenance (O&M) Guidelines for Air Pollution Control Equipment, 1993, p 2-16.

[16] Technical Support Document for the Regulatory Impact Analysis of the Enhanced Monitoring Rule. Mathtech, Inc. September 30,1993, p 2-21, 2-22.

[17] User Manual: United States Environmental Protection Agency's Continuous Emissions Monitoring System Cost Model Version 3.0, p 3-7

[18] Jahnke, James A. Continuous Emission Monitoring, Van Nostrand Reinhold, New York, c1993.

Appendix A

ASSUMPTIONS FOR PERSONNEL AND EQUIPMENT COST FACTORS

Appendix A consists of tables of assumed values for personnel and equipment cost factors. The total capital investment (TCI) and total annual cost (TAC) equations, Equations 4.1 and 4.2, were derived with these values built into them. These assumptions must be considered in determining the applicability of the equations to a specific source. See the CEMS Cost Model for additional information regarding these tables and their development.

Table 4.15: Default Personnel Travel and Per Diem Cost Factors

Cost Item	CEE Technicial	Plant Technician II	Plant Consultant	CEMS Personal	Test
Wage rate, $/hr w/o OH	30.00	18.00	27.00	27.00	16.00
Overhead (OH), % of wage rate	40	40	40	200	200
Fee, % profit	N/A	N/A	N/A	10	10
Hourly Rate1	42.00	25.20	37.	89.1	52.8

Table 4.16: Default Auxiliary Equipment Costs for CEMS ($)

Equipment	Extractive	In-situ	FTIR
Sampling system			
After control	40,000	1,000	38,000
Before control	50,000	2,000	48,000
Data acquisition system	20,000	20,000	16,000[a]
CEMS shelter	12,000	N/A	10,000
Fabrication of system in shelter	12,800	N/A	7,700
Monitor control unit	N/A	10,000	N/A

[a] Only needed if system includes opacity or PM monitor.

EPA/452/B-02-001

Section 3

VOC Controls

EPA/452/B-02-001

Section 3.1

VOC Recapture Controls

EPA/452/B-02-001

Chapter 1

Carbon Adsorbers

William M. Vatavuk
Innovative Strategies and Economics Group, OAQPS
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

William L. Klotz
Chas. T. Main, Inc.
Charlotte, NC 28224

Robert L. Stallings
Ozone Policy and Strategies Group, OZQPS
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

September 1999

Contents

1.0 Introduction 1-3

1.1 Types of Adsorbers 1-3
1.1.1 Fixed-bed Units 1-3
1.1.2 Cannister Units 1-5
1.1.3 Adsorption Theory 1-6

1.2 Design Procedure 1-12
1.2.1 Sizing Parameters 1-12
1.2.2 Determining Adsorption and Desorption Times 1-13
1.2.3 Estimating Carbon Requirement 1-15
1.2.3.1 Overview of Carbon Estimation Procedures 1-15
1.2.3.2 Carbon Estimation Procedure Used in Manual 1-15

1.3 Estimating Total Capital Investment 1-17
1.3.1 Fixed-Bed Systems 1-17
1.3.1.1 Carbon Cost 1-17
1.3.1.2 Vessel Cost 1-18
1.3.1.3 Total Purchased Cost 1-21
1.3.1.4 Total Capital Investment 1-23
1.3.2 Cannister Systems 1-23

1.4 Estimating Total Annual Cost 1-24
1.4.1 Direct Annual Costs 1-24
1.4.1.1 Steam1-281-25
1.4.1.2 Cooling Water 1-25
1.4.1.3 Electricity 1-26
1.4.1.4 Carbon Replacement 1-29
1.4.1.5 Solid Waste Disposal 1-29
1.4.1.6 Operating and Supervisory Labor 1-30
1.4.1.7 Maintenance Labor and Materials 1-30
1.4.2 Indirect Annual Costs 1-30
1.4.3 Recovery Credits 1-31
1.4.4 Total Annual Cost 1-32
1.4.5 Example Problem 1-32

1.0 Introduction

In air pollution control, adsorption is employed to remove volatile organic compounds (VOCs) from low to medium concentration gas streams, when a stringent outlet concentration must be met and/or recovery of the VOC is desired. Adsorption itself is a phenomenon where gas molecules passing through a bed of solid particles are selectively held there by attractive forces which are weaker and less specific than those of chemical bonds. During adsorption, a gas molecule migrates from the gas stream to the surface of the solid where it is held by physical attraction releasing energy—the "heat of adsorption", which typically equals or exceeds the heat of condensation. Adsorptive capacity of the solid for the gas tends to increase with the gas phase concentration, molecular weight, diffusivity, polarity, and boiling point. Gases form actual chemical bonds with the adsorbent surface groups. This phenomenon is termed "chemisorption".

Most gases ("adsorbates") can be removed ("desorbed") from the adsorbent by heating to a sufficiently high temperature, usually via steam or (increasingly) hot combustion gases, or by reducing the pressure to a sufficiently low value (vacuum desorption). The physically adsorbed species in the smallest pores of the solid and the chemisorbed species may require rather high temperatures to be removed, and for all practical purposes cannot be desorbed during regeneration. For example, approximately 3 to 5 percent of organics adsorbed on virgin activated carbon is either chemisorbed or very strongly physically adsorbed and is difficult to desorb during regeneration.[1]

Adsorbents in large scale use include activated carbon, silica gel, activated alumina, synthetic zeolites, fuller's earth, and other clays. This section is oriented toward the use of activated carbon, a commonly used adsorbent for VOCs.

1.1 Types of Adsorbers

Five types of adsorption equipment are used in collecting gases: (1) fixed regenerable beds; (2) disposable/rechargeable cannisters; (3) traveling bed adsorbers; (4) fluid bed adsorbers; and (5) chromatographic baghouses.[2] Of these, the most commonly used in air pollution control are fixed-bed systems and cannister types. This section addresses only fixed-bed and cannister units.

1.1.1 Fixed-bed Units

Fixed-bed units can be sized for controlling continuous, VOC-containing streams over a wide range of flow rates, ranging from several hundred to several hundred thousand cubic feet per minute (cfm). The VOC concentration of streams that can be treated by fixed-bed adsorbers can be as low as several parts per billion by volume (ppbv) in the case of some toxic chemicals or as high as 25% of the VOCs' lower explosive limit (LEL). (For most VOCs, the LEL ranges from 2500 to 10,000 ppmv.[3])

Fixed-bed adsorbers may be operated in either intermittent or continuous modes. In intermittent operation, the adsorber removes VOC for a specified time (the "adsorption time"), which corresponds to the time during which the controlled source is emitting VOC. After the adsorber and the source are shut down (e.g., overnight), the unit begins the desorption cycle during which the captured VOC is removed from the carbon. This cycle, in turn, consists of three steps: (1) regeneration of the carbon by heating, generally by blowing steam through the bed in the direction opposite to the gas flow;[1] (2) drying of the bed, with compressed air or a fan; and (3) cooling the bed to its operating temperature via a fan. (In most designs, the same fan can be used both for bed drying and cooling.) At the end of the desorption cycle (which usually lasts 1 to 1½ hours), the unit sits idle until the source starts up again.

In continuous operation a regenerated carbon bed is always available for adsorption, so that the controlled source can operate continuously without shut down. For example, two carbon beds can be provided: while one is adsorbing, the second is desorbing/idled. As each bed must be large enough to handle the entire gas flow while adsorbing, twice as much carbon must be provided than an intermittent system handling the same flow. If the desorption cycle is significantly shorter than the adsorption cycle, it may be more economical to have three, four, or even more beds operating in the system. This can reduce the amount of extra carbon capacity needed or provide some additional benefits, relative to maintaining a low VOC content in the effluent. (See Section 1.2 for a more thorough discussion of this.)

A typical two-bed, continuously operated adsorber system is shown in Figure 3.1. One of the two beds is adsorbing at all times, while the other is desorbing/idled. As shown here, the VOC-laden gas enters vessel #1 through valve A, passes through the carbon bed (shown by the shading) and exits through valve B, from whence it passes to the stack. Meanwhile, vessel #2 is in the desorption cycle. Steam enters through valve C, flows through the bed and exits through D. The steam-VOC vapor mixture passes to a condenser, where cooling water condenses the entire mixture. If part of the VOC is immiscible in water, the condensate next passes to a decanter, where the VOC and water layers are separated. The VOC layer is conveyed to storage. If impure, it may receive additional purification by distillation. Depending on its quality (i.e., quantity of dissolved organics), the water layer is usually discharged to a wastewater treatment facility.

[1] Although steam is the most commonly used regenerant, there are situations where it should not be used. An example would be a degreasing operation that emits halogenated VOCs. Steaming might cause the VOCs to decompose.

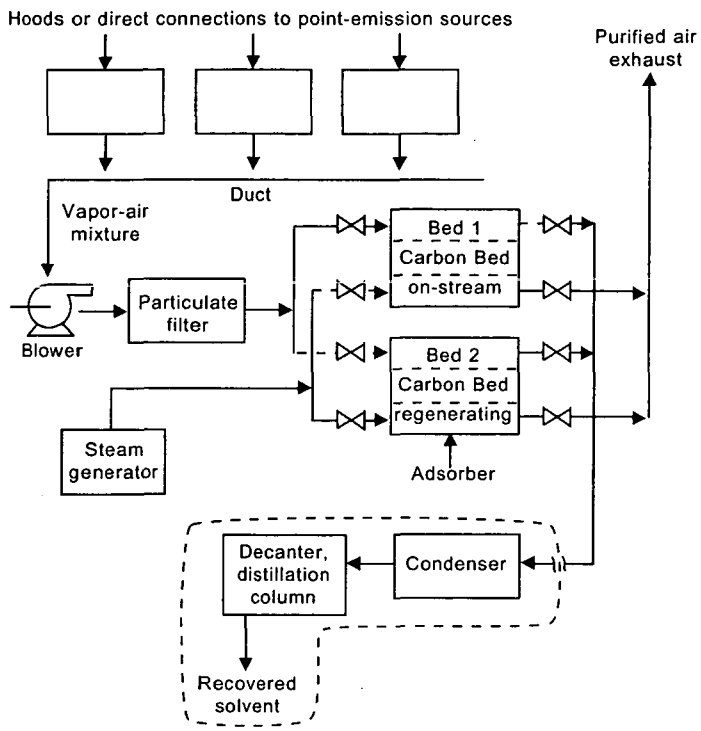


Figure 1.1: Typical-Two-Bed, Continuously Operated Fixed Bed Carbon Adsorber System

1.1.2 Cannister Units

Cannister-type adsorbers originally refered to relatively small returnable containers, such as 55-gallon drums. Cannisters differ from fixed-bed units, in that they are normally limited to controlling lower-volume and intermittent gas streams, such as those emitted by storage tank vents, where process economics dictate that off-site regeneration is appropriate. The carbon cannisters are not intended for desorption on-site. However, the carbon may be regenerated at a central facility. The term cannister is becoming something of a misnomer as much of the growth in the industry is in larger vessels without regeneration capabilities. Calgon provided information on standard systems as large as 8,000 cfm and carbon capacitites of 2,000 pounds; TIGG Corporation reported systems as large as 30,000 cfm. [4][5]

Once the carbon reaches a certain VOC content, the unit is shut down and either the carbon or the cannister is replaced. The carbon or cannister is then returned to a reclaimation facility or regenerated at the central facility. Each cannister unit consists of a vessel, activated carbon, inlet connection and distributer leading to the carbon bed, and an outlet connection for the purified gas stream. In one design (Calgon's Ventsorb®), 180 lbs of carbon are installed on an 8-

inch gravel bed, in a 55-gallon drum with an internal collector. The type of carbon used depends on the nature of the VOC to be treated [6]. Non-regenrable vessles can be placed in a series, this protects agianst breakthrough because in the event that the first cannister or vessel becomes saturated with VOC, the second then becomes the primary carbon adsorber. One option would be to periodically remove the most saturated cannister or carbon bed and add a fresh cannister or carbon bed to the clean end. Periodic sampling for breakthrough between the carbon beds would assure that replacement occured frequently enough to avoid breakthrough to the atmosphere. This approach also improves cost effectiveness of carbon replacement because the carbon is at or near its saturation point when it is replaced.

In theory, a cannister unit would remain in service longer than a regenerable unit would stay in its adsorption cycle due to a higher theoretical capacity for fresh carbon compared to carbon regenrated on-site. The service life is based on a service factor determined by the ratio of the theoretical capacity to the working capacity. Determining service factors help to insure the allowable outlet concentration from being exceeded. In reality, however, poor operating practice may result in the cannister remaining connected until the carbon is near or at saturation. This is because: (1) the carbon (and often the vessel) will probably be disposed of, so there is the temptation to operate it until the carbon is saturated; and (2) unlike fixed-bed units, whose outlet VOC concentrations are usually not monitored continuously (via flame ionization detectors, typically), canisters are usually not monitored. Adequate recordeeping and periodic monitoring for breakthrough can be supported by bed life modeling provided by vendors to ensure that cannister replacement occurs with sufficient frequency and that breakthrough does not occur.

1.1.3 Adsorption Theory

At equilibrium, the quantity of gas that is adsorbed on activated carbon is a function of the adsorption temperature and pressure, the chemical species being adsorbed, and the carbon characteristics, such as carbon particle size and pore structure. For a given adsorbent-VOC combination at a given temperature, an adsorption isotherm can be constructed which relates the mass of adsorbate per unit weight of adsorbent ("equilibrium adsorptivity") to the partial pressure of the VOC in the gas stream. The adsorptivity increases with increasing VOC partial pressure and decreases with increasing temperature.

A family of adsorption isotherms having the shape typical of adsorption on activated carbon is plotted in Figure 3.2. This and other isotherms whose shapes are convex upward throughout, are designated "Type I" isotherms. The Freundlich isotherm, which can be fit to a portion of a Type I curve, is commonly used in industrial design.[2]

$$W_e = k P_m \qquad (1.1)$$

where

w_e	=	equilibrium adsorptivity (lb adsorbate/lb adsorbent)
P	=	partial pressure of VOC in gas stream (psia)
k, m	=	empirical parameters

The treatment of adsorption from gas mixtures is complex and beyond the scope of this chapter. Except where the VOC in these mixtures have nearly identical adsorption isotherms, one VOC in a mixture will tend to displace another on the carbon surface. Generally, VOCs with lower vapor pressures will displace those with higher vapor pressure, resulting in the former displacing the latter previously adsorbed. Thus, during the course of the adsorption cycle the carbon's capacity for a higher vapor pressure constituent decreases. This phenomenon should be considered when sizing the adsorber. To be conservative, one would normally base the adsorption cycle requirements on the least adsorbable component in a mixture and the desorption cycle on the most adsorbable component.[1]

The equilibrium adsorptivity is the maximum amount of adsorbate the carbon can hold at a given temperature and VOC partial pressure. In actual control systems where there are not two beds operating in series, however, the entire carbon bed is never allowed to reach equilibrium. Instead, once the outlet concentration reaches a preset limit (the "breakthrough concentration"), the adsorber is shut down for desorption or (in the case of cannister units) replacement and disposal. At the point where the vessel is shut down, the average bed VOC concentration may only be 50% or less of the equilibrium concentration. That is, the carbon bed may be at equilibrium ("saturated") at the gas inlet, but contain only a small quantity of VOC near the outlet.

As Equation 3.1 indicates, the Freundlich isotherm is a power function that plots as a straight line on log-log paper. Conveniently, for the concentrations/partial pressures normally encountered in carbon adsorber operation, most VOC-activated carbon adsorption conforms to Equation 3.1. At very low concentrations, typical of breakthrough concentrations, a linear approximation (on arithmetic coordinates) to the Freundlich isotherm is adequate. However, the Freundlich isotherm does not accurately represent the isotherm at high gas concentrations and thus should be used with care as such concentrations are approached.

Adsorptivity data for selected VOCs were obtained from Calgon Corporation, a vendor of activated carbon.[6] The vendor presents adsorptivity data in two forms: a set of graphs displaying equilibrium isotherms [6] and as a modification of the Dubinin-Radushkevich (D-R) equation, a semi-empirical equation that predicts the adsorptivity of a compound based on its adsorption potential and polarizability.[8] In this Manual, the modified D-R equation is referred to as the Calgon fifth-order polynomial. The data displayed in the Calgon graphs [6] has been fit to the Freundlich equation. The resulting Freundlich parameters are shown in Table 1.1 for a limited number of chemicals. The adsorbates listed include aromatics (e.g., benzene, toluene), chlorinated aliphatics (dichloroethane), and one ketone (acetone). However, the list is far from all-inclusive.

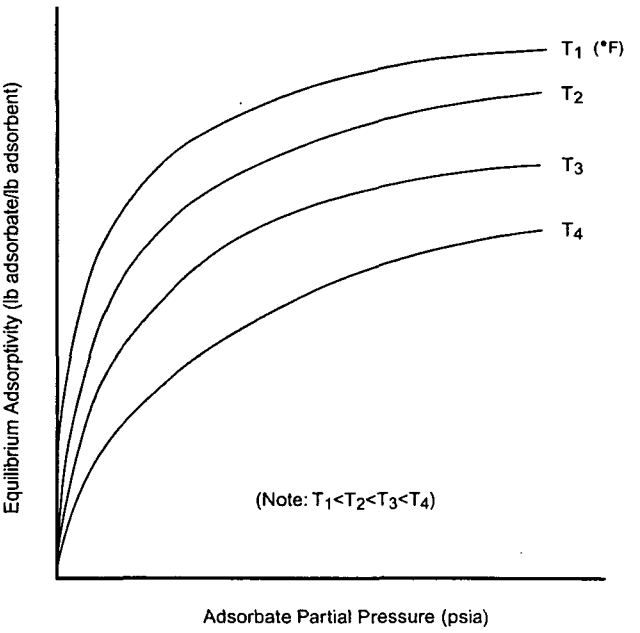


Figure 1.2: Type I Adsorption Isotherms for Hypothetical Adsorbate

Notice that a range of partial pressures is listed with each set of parameters, k and m. (Note: In one case (m-xylene) the isotherm was so curvilinear that it had to be split into two parts, each with a different set of parameters.) This is the range to which the parameters apply. Extrapolation beyond this range—especially at the high end—can introduce inaccuracy to the calculated adsorptivity.

But high-end extrapolation may not be necessary, as the following will show. In most air pollution control applications, the system pressure is approximately one atmosphere (14.696 psia). The upper end of the partial pressure ranges in Table 1.1 goes from 0.04 to 0.05 psia. According to Dalton's Law, at a total system pressure of one atmosphere this corresponds to an adsorbate concentration in the waste gas of 2,720 to 3,400 ppmv. Now, as discussed in Section 1.1.2, the adsorbate concentration is usually kept at 25% of the lower explosive limit (LEL).[2] For many VOCs, the LEL ranges from 1 to 1.5 volume %, so that 25% of the LEL would be 0.25 to 0.375% or 2,500 to 3,750 ppmv, which approximates the high end of the partial pressure ranges in Table 1.1.

[2] Although, Factory Mutual Insurance will reportedly permit operation at up to 50% of the LEL, if proper VOC monitoring is used.

Finally, each set of parameters applies to a fixed adsorption temperature, ranging from 77° to 104° F. These temperatures reflect typical operting conditions, although adsorption can take place as low as 32°F and even higher than 104°F. As the adsorption temperature increases to much higher levels, however, the equilibrium adsorptivity decreases to such an extent that VOC recovery by carbon adsorption may become economically impractical.

Table 1.1: Parameters for Selected Adsorption Isotherms [6][a]

Adsorbate	Adsorption Temp (°F)	Isotherm Parameters		Range of Isotherm[b] (psia)
		k	*m*	
Benzene	77	0.597	0.176	0.0001-0.05
Chlorobenzene	77	1.05	0.188	0.0001-0.01
Cyclohexane	100	0.505	0.210	0.0001-0.05
Dichloroethane	77	0.976	0.281	0.0001-0.04
Phenol	104	0.855	0.153	0.0001-0.03
Trichloroethane	77	1.06	0.161	0.0001-0.04
Vinyl Chloride	100	0.200	0.477	0.0001-0.05
m-Xylene	77	0.708	0.113	0.0001-0.001
	77	0.527	0.0703	0.001-0.05
Acrylonitrile	100	0.935	0.424	0.0001-0.015
Acetone	100	0.412	0.389	0.0001-0.05
Toluene	77	0.551	0.110	0.001-0.05

[a] Each isotherm is of the for $W_e=kP^p$. (See text for definition of terms.) Data are for adsorption of Calgon type "BPL" carbon.
[b] Equation should not be extrapolated outside these ranges.

The Calgon fifth-order polynomial is somewhat more accurate than the Freundilich parameters from Table 1.1. The polynomial contains a temperature parameter, and it allows one to estimate adsorption isotherms for compounds not shown in Table 1.1 if pure component data are available. The pure component data required are the saturation pressure, liquid molar volume, and the refractive index. It is, however, somewhat more complex to use than the Freundlich equation. The Calgon fifth-oder polynomial is as follows:

The mass loading, w_e, is calculated from

$$w_e = \frac{0.01\,G}{V_m}\,MW_{ads} \qquad (1.2)$$

where

w_e	=	mass loading, i.e., equilibrium adsorptivity (g adsorbate per g carbon)[3]
G	=	carbon loading at equilibrium (cm^3 liquid adsorbate per 100 g carbon)
V_m	=	liquid molar volume of adsorbate (cm^3 per g-mole).
MW_{ads}	=	molecular weight of Adsorbute

Note that the terms in Equation 1.2 are given in metric units, not English. This has been done because the carbon loading, G, is calculated from a regression equation in which all the terms are expressed in metric units. This equation for G is the Calgon fifth-order polynomial:

$$log_{10}(G) = A_0 + A_1 Y + A_2 Y^2 + A_3 Y^3 + A_4 Y^4 + A_5 Y^5 \qquad (1.3)$$

where

A_0	=	1.71
A_1	=	-1.46×10^{-2}
A_2	=	-1.65×10^{-3}
A_3	=	-4.11×10^{-4}
A_4	=	$+3.14 \times 10^{-5}$
A_5	=	-6.75×10^{-7}

and Y is calculated from several equations which follow.

The first step in calculating Y is to calculate χ. This can be done by calculating the adsorption potential, ε :

$$\varepsilon = RT \ln \left(\frac{P_s}{P_i} \right) \qquad (1.4)$$

where

R	=	1.987 (calories per g-mole-K
T	=	absolute temperature (K)
P_s	=	vapor pressure of adsorbate at the temperature T (kPa)
P_i	=	partial pressure of adsorbate (kPa).

The χ is calculated from:

$$\chi = \frac{\varepsilon}{(2.303 \; R \; V_m)} \qquad (1.5)$$

[3] This, of course, is equal to *lb* absorbate per *lb* carbon.

By substituting for ϵ in the above equation, χ can alternatively be calculated from[4]:

$$\chi = \left(\frac{T}{V_m}\right) log_{10}\left(\frac{P_s}{P_i}\right) \tag{1.6}$$

The next step in calculating Y is to calculate the relative polarizability, Γ.

$$\Gamma = \frac{\Theta_i}{\Theta_o} \tag{1.7}$$

where

Θ_i = polarizability of component i per unit volume, where component i is the adsorbate

Θ_o = polarizability of component o per unit volume, where component o is the reference component, n-heptane.

For the adsorbate or the reference compound, using the appropriate refractive index of adsorbate, n, the polarizability is calculated from:

$$\Theta = \frac{n^2 - 1}{n^2 + 2} \tag{1.8}$$

Once χ and Γ are known, Y can be calculated from:

$$Y = \frac{\chi}{\Gamma} \tag{1.9}$$

Calgon also has a proprietary, seventh-order form in which two additional coefficients are added to the Calgon fifth-order polynomial, but the degree of fit reportedly is improved only modestly.[8] Additional sources of isotherm data include the activated carbon vendors, handbooks (such as Perry's Chemical Engineer's Handbook), and the literature.

[4] Alternatively, if the available values for T, P_i, P_s, and Vm are in English units, they may be substitued into this equation without conversion. However, to make the result dimensionally consistent with Equation 1.3, it would have to be multiplied by a conversion factor, *34.7*.

1.2 Design Procedure

1.2.1 Sizing Parameters

Data received from adsorber vendors indicate that the size and purchase cost of a fixed-bed or cannister carbon adsorber system primarily depend on five parameters:

1. The volumetric flow of the VOC laden gas passing through the carbon bed(s);
2. The inlet and outlet *VOC* mass loadings of the gas stream;
3. The adsorption time (i.e,. the time a carbon bed remains on-line to adsorb VOC before being taken off-line for desorption of the bed);
4. The working capacity of the activated carbon in regenerative systems or the *equilibrium cpacity* in the case of non-regenerative systems,
5. The humidity of the gas stream, especially in the effect of humidy on capacity in relation to halogens.

In addition, the cost could also be affected by other stream conditions, such as the presence/ absence of excessive amounts of particulate, moisture, or other substances which would require the use of extensive pretreatment and/or corrosive-resistant construction materials.

The purchased cost depends to a large extent on the volumetric flow (usually measured in actual ft^3/min). The flow, in turn, determines the size of the vessels housing the carbon, the capacities of the fan and motor needed to convey the waste gas through the system, and the diameter of the ducting.

Also important are the VOC inlet and outlet gas stream loadings, the adsorption time, and the working or equilibium capacity of the carbon. These variables determine the amount and cost of carbon charged to the system initially and, in turn, the cost of replacing that carbon after it is exhausted (typically, five years after startup). Moreover, the amount of the carbon charge affects the size and cost of the auxiliary equipment (condenser, decanter, bed drying/cooling fan), because the sizes of these items are tied to the amount of VOC removed by the bed. The amount of carbon also has a bearing on the size and cost of the vessels.

A carbon adsorber vendor [9] supplied data that illustrate the dependency of the equipment cost on the amount of the carbon charge. Costs were obtained for fixed-bed adsorbers sized to handle three gas flow rates ranging from 4,000 to 100,000 scfm and to treat inlet VOC (toluene) concentrations of 500 and 5,000 ppm. Each adsorber was assumed to have an eight-hour

adsorption time. As one might expect, the equipment costs for units handling higher gas flow rates were higher than those handling lower gas flow rates. Likewise, at each of the gas flow rates, the units sized to treat the 5,000 ppm VOC streams had higher equipment costs than those sized to treat the 500 ppm concentration. These cost differences ranged from 23 to 29% and averaged 27%. These higher costs were partly needed to pay for the additional carbon required to treat the higher concentration streams. But some of these higher costs were also needed for enlarging the adsorber vessels to accommodate the additional carbon and for the added structural steel to support the larger vessels. Also, larger condensers, decanters, cooling water pumps, etc., were necessary to treat the more concentrated streams. (See Section 1.3.)

The VOC inlet loading is set by the source parameters, while the outlet loading is set by the VOC emission limit. (For example, in many states, the average VOC outlet concentration from adsorbers may not exceed 25 ppm.)

1.2.2 Determining Adsorption and Desorption Times

The relative times for adsorption and desorption and the adsorber bed configuration (i.e., whether single or multiple and series or parallel adsorption beds are used) establish the adsorption/ desorption cycle profile. The circle profile is important in determining carbon and vessel requirements and in establishing desorption auxiliary equipment and utility requirements. An example will illustrate. In the simplest case, an adsorber would be controlling a process which emits a relatively small amount of VOC intermittently—say, during one 8-hour shift per day. During the remaining 16 hours the system would either be desorbing or on standby. Properly sized, such a system would only require a single bed, which would contain enough carbon to treat eight hours worth of gas flow at the specified inlet concentration, temperature, and pressure. Multiple beds, operating in parallel, would be needed to treat large gas flows (>100,000 actual ft^3/min, generally)[9], as there are practical limits to the sizes to which adsorber vessels can be built. But, regardless of whether a single bed or multiple beds were used, the system would only be on-line for part of the day.

However, if the process were operating continuously (24 hours), an extra carbon bed would have to be installed to provide adsorptive capacity during the time the first bed is being regenerated. The amount of this extra capacity must depend on the number of carbon beds that would be adsorbing at any one time, the length of the adsorption period relative to the desorption period, and whether the beds were operating in parallel or in series. If one bed were adsorbing, a second would be needed to come on-line when the first was shut down for desorption. In this case, 100% extra capacity would be needed. Similarly, if five beds in parallel were operating in a staggered adsorption cycle, only one extra bed would be needed and the extra capacity would be 20% (i.e., 1/5)—provided, of course, that the adsorption time were at least five times as long as the desorption time. The relationship between adsorption time, desorption time, and the required extra capacity can be generalized.

$$M_c = M_{c_I} f \tag{1.10}$$

where

M_c, M_{ccI} = amounts of carbon required for continuous or intermittent control of a given source, respectively (lbs)

f = extra capacity factor (dimensionless)

This equation shows the relationship between M_c and M_{CI}. Section 1.2.3 shows how to calculate these quantities.

The factor, f, is related to the number of beds adsorbing (N_A) and desorbing (N_D) in a continuous system as follows:

$$f = 1 + \frac{N_D}{N_A} \tag{1.11}$$

(Note: N_A is also the number of beds in an intermittent system that would be adsorbing at any given time. The total number of beds in the system would be $N_A + N_D$.)

It can be shown that the number of desorbing beds required in a continuous system (N_D) is related to the desorption time (θ_D), adsorption time (θ_A), and the number of adsorbing beds, as follows:

$$\theta_D \leq \theta_A \left(\frac{N_D}{N_A} \right) \tag{1.12}$$

(Note: θ_D is the total time needed for bed regeneration, drying, and cooling.)

For instance, for an eight-hour adsorption time, in a continuously operated system of seven beds (six adsorbing, one desorbing) θ_D would have to be 1-1/3 hours or less (8 hours/6 beds). Otherwise, additional beds would have to be added to provide sufficient extra capacity during desorption.

1.2.3 Estimating Carbon Requirement

1.2.3.1 Overview of Carbon Estimation Procedures

Obtaining the carbon requirement (M_c or M_{cI}) is not as straightforward as determining the other adsorber design parameters. When estimating the carbon charge, the sophistication of the approach used depends on the data and calculational tools available.

One approach for obtaining the carbon requirement is a rigorous one which considers the unsteady-state energy and mass transfer phenomena occurring in the adsorbent bed. Such a procedure necessarily involves a number of assumptions in formulating and solving the problem. Such a procedure is beyond the scope of this Manual at the present time, although ongoing work in the Agency is addressing this approach.

In preparing this section of the Manual, we have adopted a rule-of-thumb procedure for estimating the carbon requirement. This procedure, while approximate in nature, appears to have the acceptance of vendors and field personnel. It is sometimes employed by adsorber vendors to make rough estimates of carbon requirement and is relatively simple and easy to use. It normally yields results incorporating a safety margin, the size of which depends on the bed depth (short beds would have less of a safety margin than deep beds), the effectiveness of regeneration, the particular adsorbate and the presence or absence of impurities in the stream being treated.

1.2.3.2 Carbon Estimation Procedure Used in Manual

The rule-of-thumb carbon estimation procedure is based on the "working capacity" (W_e, lb VOC/lb carbon). This is the difference per unit mass of carbon between the amount of VOC on the carbon at the end of the adsorption cycle and the amount remaining on the carbon at the end of the desorption cycle. It should not be confused with the "equilibrium capacity" (W_e,) defined above in Section 1.1.3. Recall that the equilibrium capacity measures the capacity of virgin activated carbon when the VOC has been in contact with it (at a constant temperature and partial pressure) long enough to reach equilibrium. In adsorber design, it would not be feasible to allow the bed to reach equilibrium. If it were, the outlet concentration would rapidly increase beyond the allowable outlet (or "breakthrough") concentration until the outlet concentration reached the inlet concentration. During this period the adsorber would be violating the emission limit. With non-regenerable (cannister) type systems, placing multiple vessles in a series can substantially decrease concerns of breakthrough.

The working capacity is some fraction of the equilibrium capacity. Like the equilibrium adsorptivity, the working capacity depends upon the temperature, the VOC partial pressure, and the VOC composition. The working capacity also depends on the flow rate and the carbon bed parameters.

The working capacity, along with the adsorption time and VOC inlet loading, is used to compute the carbon requirement for a cannister adsorber or for an intermittently operated fixed-bed adsorber as follows:

$$M_{c_I} = \frac{m_{voc}}{w_e}\,\theta_A \tag{1.13}$$

where

m_{voc} = VOC inlet loading (lb/h)

Combining this with Equations 1.10 and 1.11 yields the general equation for estimating the system total carbon charge for a continuously operated system:

$$M_c = \frac{m_{voc}}{w_c}\,\theta_A\left(1 + \frac{N_D}{N_A}\right) \tag{1.14}$$

Values for w_c may be obtained from knowledge of operating units. If no value for w_c is available for the VOC (or VOC mixture) in question, the working capacity may be estimated at 50% of the equilibrium capacity, as follows:

$$w_c \approx 0.5\,w_{e(max)} \tag{1.15}$$

where

$w_{e(\max)}$ = the equilibrium capacity (lb VOC/lb carbon) taken at the adsorber inlet (i.e., the point of maximum VOC concentration).

(Note: To be conservative, this 50% figure should be lowered if short desorption cycles, very high vapor pressure constituents, high moisture contents significant amounts of impurities, or difficult-to-desorb VOCs are involved. Furthermore, the presence of strongly adsorbed impurities in the inlet VOC stream may significantly shorten carbon life.)

As Equation 1.14 shows, the carbon requirement is directly proportional to the adsorption time. This would tend to indicate that a system could be designed with a shorter adsorption time to minimize the carbon requirement (and equipment cost). There is a trade-off here not readily apparent from Equation 1.14, however. Certainly, a shorter adsorption time would require less carbon. But, it would also mean that a carbon bed would have to be desorbed more frequently. This would mean that the regeneration steam would have to be supplied to the bed(s) more frequently to remove (in the long run) the same amount of VOC. Further, each time the bed is regenerated

the steam supplied must heat the vessel and carbon, as well as drive off the adsorbed VOC. And the bed must be dried and cooled after each desorption, regardless of the amount of VOC removed. Thus, if the bed is regenerated too frequently, the bed drying/cooling fan must operate more often, increasing its power consumption. Also, more frequent regeneration tends to shorten the carbon life. As a rule-of-thumb, the optimum regeneration frequency for fixed-bed adsorbers treating streams with moderate to high VOC inlet loadings is once every 8 to 12 hours.[1]

1.3 Estimating Total Capital Investment

Entirely different procedures should be used to estimate the purchased costs of fixed-bed and cannister-type adsorbers. Therefore, they will be discussed separately.

1.3.1 Fixed-Bed Systems

As indicated in the previous section, the purchased cost is a function of the volumetric flow rate, VOC inlet and outlet loadings, the adsorption time, and the working capacity of the activated carbon. As Figure 3.1 shows, the adsorber system is made up of several different items. Of these, the adsorber vessels and the carbon comprise from one-half to nearly 90% of the total equipment cost. (See Section 1.3.1.3.) There is also auxiliary equipment, such as fans, pumps, condensers, decanters, and internal piping. But because these usually comprise a small part of the total purchased cost, they may be "factored" from the costs of the carbon and vessels without introducing significant error. The costs of these major items will be considered separately.

1.3.1.1 Carbon Cost

Cabon Cost, C_c, in dollars ($) is simply the product of the initial carbon requirement (M_c) and the current price of carbon. As adsorber vendors buy carbon in very large quantities (million-pound lots or larger), their cost is somewhat lower than the list price. For larger systems (other than cannister) Calgon reports typical carbons costs to be $0.75 to $1.25 for virgin carbon and $0.50 to $0.75 for reactiviated carbon (mid-1999 dollars). A typical of $1.00/lb is used in the calculation below.[5][10] Thus:

$$C_c = 1.00 M_c \tag{1.16}$$

1.3.1.2 Vessel Cost

The cost of an adsorber vessel is primarily determined by its dimensions which, in turn, depend upon the amount of carbon it must hold and the superficial gas velocity through the bed that must be maintained for optimum adsorption. The desired superficial velocity is used to calculate the cross-sectional area of the bed perpendicular to the gas flow. An acceptable superficial velocity is established empirically, considering desired removal efficiency, the carbon particle size and bed

porosity, and other factors. For example, one adsorber vendor recommends a superficial bed velocity of 85 ft/min[9], while an activated carbon manufacturer cautions against exceeding 60 ft/min in systems operating at one atmosphere.[7] Another vendor uses a 65 ft/min superficial face velocity in sizing its adsorber vessels.[10] Lastly, there are practical limits to vessel dimensions which also influence their sizing. That is, due to shipping restrictions, vessel diameters rarely exceed 12 feet, while their length is generally limited to 50 feet.[10]

The cost of a vessel is usually correlated with its weight. However, as the weight is often difficult to obtain or calculate, the cost may be estimated from the external surface area. This is true because the vessel material cost—and the cost of fabricating that material—is directly proportional to its surface area. The surface area (S, ft^2) of a vessel is a function of its length (L, ft) and diameter (D, ft), which in turn, depend upon the superficial bed face velocity, the L/D ratio, and other factors.

Most commonly, adsorber vessels are cylindrical in shape and erected horizontally (as in Figure 1.1). Vessels configured in this manner are generally subjected to the constraint that the carbon volume occupies no more than 1/3 of the vessel volume [9][10]. It can be shown that this constraint limits the bed depth to no more than:

$$Maximum\ bed\ depth \approx \frac{\pi D}{12} \tag{1.17}$$

The vessel length, L, and diameter, D, can be estimated by solving two relationships, namely, (1) the equation relating carbon volume, and thus vessel volume, to L and D, and (2) the equation relating volumetric flow rate, superficial velocity, and cross-section normal to flow. If one assumes that the carbon bulk density is 30 lb/ft^3, then one can show that:

$$D = \frac{0.127\, M_C\, v_b}{Q} \tag{1.18}$$

$$L = \frac{7.87}{M_c'} \left(\frac{Q'}{v_b} \right)^2 \tag{1.19}$$

where

D	=	vessel diameter (ft)
L	=	vessel length (ft)
v_b	=	bed superficial velocity (ft/min)
M_c	=	carbon requirement *per vessel* (lbs)
Q	=	volumetric flow rate *per adsorbing vessel* (acfm)

Because the constants in Equations 1.18 and 1.19 are not dimensionless, one must be careful to use the units specified in these equations.

Although other design considerations can result in different values of L and D, these equations result in L and D which are acceptable from the standpoint of "study" cost estimation for horizontal, cylindrical vessels which are larger than 2-3 feet in diameter.

The carbon requirement and flow rate for each adsorber vessel can be calculated as follows:

$$M_c = \frac{M_e}{(N_A + N_D)}$$

$$Q = \frac{Q}{N_A} \qquad (1.20)$$

At gas flow rates (Q) of less than 9,000 scfm, it is usually more feasible to erect the adsorber vessels vertically instead of horizontally.[10] If so, the vessel diameter can be calculated from the volumetric flow rate per adsorbing vessel and the bed superficial velocity as follows:

$$D = \left(\frac{4Q'}{\pi\, v_b}\right)^{1/2} \qquad (1.21)$$

The vertical vessel length will depend principally on the carbon bed thickness. Additional space must be included below the carbon bed for bed support and above and below the bed for distribution and disengaging of the gas stream and for physical access to the carbon bed. In smaller diameter vessels, access to both sides of the bed is usually not required. However, 1 to 1½ feet must be provided on each side for gas distribution and disengagement, or 2 to 3 feet overall. For longer vessels, 2 to 3 feet at each end of the vessel is typically provided for access space.

Given the mass of carbon in the bed, the carbon bulk density, and the bed diameter (i.e., the cross-sectional area normal to flow), determining the carbon bed thickness is straight forward using the following equation:

$$t_b = \frac{volume\ of\ carbon}{cross-sectional\ area\ normal\ to\ flow} = \frac{\left(\frac{M_c'}{r_b}\right)}{\left(\frac{Q'}{v_b}\right)} \qquad (1.22)$$

where

ρ_b = carbon bulk density (lb/ft^3, assume 30 lb/ft^3)

The vessel length is, therefore,

$$L = t_b + t_{a,g} \tag{1.23}$$

where

$t_{a,g}$ = access / gas distribution allowance
= 2 to 6 feet (depending on vertical vessel diameter)

Finally, use the following equation to calculate the surface area of either a horizontal or vertical vessel:

$$S = \pi D \left(L + \frac{D}{2} \right) \tag{1.24}$$

$$C_v = 271 S^{0.778} \tag{1.25}$$

where

C_v = vessel cost (fall 1999 $), F.O.B. vender[5]

and

$$97 \leq S \leq 2110 ft^2 \tag{1.25a}$$

These units would be made of 304 stainless steel, which is the most common material used in fabricating adsorber vessels. [9][10] However, to obtain the cost of a vessel fabricated of another material, multiply C_v by and adjustment factor (F_m). A few of these factors are listed in Table 1.2.

1.3.1.3 Total Purchased Cost

As stated earlier, the costs of such items as the fans, pumps, condenser, decanter, instrumentation, and internal piping can be factored from the sum of the costs for the carbon and vessels. Based on four data points derived from costs supplied by an equipment vendor [10], we found that, depending on the total gas flow rate (Q), the ratio (R_c) of the total adsorber equipment cost to the cost of the vessels and carbon ranged from 1.14 to 2.24. These data points apanned a gas flow rate range of approximately 4,000 to 500,000 acfm. The following regression formula fit these four points:

$$R_c = 5.82 Q^{-0.133} \tag{1.26}$$

Table 1.2: Adjustment Factors to Obtain Costs for Fabricated Material

Material	F_m Factor	Reference(s)
Stainless steel, 316	1.3	[7,8,9]
Carpenter 20 CB-3	1.9	[9]
Montel-400	2.3	[7,9]
Nickel-200	3.2	[9]
Titanium	4.5	[9]

where

$$4,000 \quad Q\,(\text{acfm}) \quad 500,000$$
$$\text{Correlation coefficient (r)} = 0.872$$

Similar equations can be developed for other vessel shapes, configurations, etc.

Based on vendor data, we developed a correlation between adsorber vessel cost and surface area:[10]

The total adsorber equipment cost (C_A) would be the product of R_e and the sum of the carbon and vessel costs, or:

$$C_A = R_c \left[C_c + C_v \left(N_A + N_D\right)\right] \tag{1.27}$$

1.3.1.4 Total Capital Investment

As discussed in Section 1, in the methodology used in this Manual, the total capital investment (TCI) is estimated from the total purchased cost via an overall direct/indirect installation cost factor. A breakdown of that factor for carbon adsorbers is shown in Table 1.3. As Section 1.2 indicates, the TCI also includes costs for land, working capital, and off-site facilities, which are not included in the direct/indirect installation factor. However as these items are rarely required with adsorber systems, they will not be considered here. Further, no factors have been provided for site preparation (SP) and buildings (Bldg.), as these site-specific costs depend very little on the purchased equipment cost.

[5] Two vendors provided information for the 1999 updates, neither felt that modifications to the capital costs of adsorber system between 1989 and 1999 were appropriate. The major change for 1999 was a decreases in the price of carbon.[4][5]

Table 1.3: Capital Cost Factors for Carbon Adsorbers [11]

Direct Costs	
Purchased equipment costs	
Adsorber + auxiliary equipment[a]	As estimated, A
Instrumentation[b]	0.10 A
Sales taxes	0.03 A
Freight	0.05 A
Purchased equipment cost, PEC	B = 1.18 A
Direct installation costs	
Foundations & supports	0.08 B
Handling & erection	0.14 B
Electrical	0.04 B
Piping	0.02 B
Insulation	0.01 B
Painting	0.01 B
Direct installation costs	0.30 B
Site preparation	As required, SP
Buildings	As required, Bldg.
Total Direct Costs, DC	**1.30 B + SP + Bldg.**
Indirect Costs (installation)	
Engineering	0.10 B
Construction and field expenses	0.05 B
Contractor fees	0.10 B
Start-up	0.02 B
Performance test	0.01 B
Contingencies	0.03 B
Total Indirect Costs, IC	**0.31 B**
Total Capital Investment = DC + IC	**1.61 B + SP + Bldg.**

[a] Ductwork and any other equipment normally not included with unit furnished by adsorber vendor.
[b] Instrumentation and controls often furnished with the adsorber, and thus included in the EC.

Note that the installation factor is applied to the total purchased equipment cost, which includes the cost of such auxiliary equipment as the stack and external ductwork and such costs as freight and sales taxes (if applicable). ("External ductwork" is that ducting needed to convey the exhaust gas from the source to the adsorber system, and then from the adsorber to the stack. Costs for ductwork and stacks are shown elsewhere in this Manual) Normally, the adjustment would also cover the instrumentation cost, but this cost is usually included with the adsorber equipment cost. Finally, note that these factors reflect "average" installation conditions and could vary considerably, depending upon the installation circumstances.

1.3.2 Cannister Systems

Once the carbon requirement is estimated using the above procedure, the number of cannisters is determined. This is done simply by dividing the total carbon requirement (M_c) by the amount of carbon contained by each cannister (typically, 150 lbs.). This quotient, rounded to the next highest digit, yields the required number of cannisters to control the vent in question.

Costs for a typical cannister (Calgon's Ventsorb®) are listed in Table 1.4. These costs include the vessel, carbon, and connections, but do not include taxes, freight, or installation charges. Note that the cost per unit decreases as the quantity purchased increases. Each cannister contains Calgon's "BPL" carbon (4 x 10 mesh), which is commonly used in industrial adsorption. However, to treat certain VOCs, more expensive specialty carbons (e.g., "FCA 4 x 10") are needed. These carbons can increase the equipment cost by 60% or more.[6] As is indicated in the caption of Table 1.4, these prices are in mid-1999 dollars.

The price of 180 pound carbon cannisters is approximately $700 in small quantities and approximately $600 (mid - 1999 dollars) in larger quantities. [5] Current trends are toward the use of larger cannisters. Calgon supplies large cannisters of 1,000 to 2,000 pound capacity where the carbon is typically exchanged in the field. Calgon reports costs per pound of carbon ranging from $0.50 to $2.00 depending on mesh, activity and type with a conservative median price of $1.50 per pound. Calgon also reports the widespread use of larger non-regenerable fixed beds either 12 feet in diameter (113 square feet capable of of handling 6,000 cfm) and 8x20 foot beds (160 square feet capable of handling 8,000 cfm). These non-regenerable fixed bed absorbers are usually atmospheric designs made of thin steel with an internal coating to inhibit corrosion. For 1,000 pound cannisters, Calgon reports typical installation costs of $3,200 and equipment costs of $5,600 and for 2,000 pound cannisters these costs are $4,600 and $7,800 respectively. For

Table 1.4: Equipment Costs (Mid 1999$) for a Typical Canister Adsorber [5]

Quantity	Equipment Costs for Each Piece[a] in Dollars ($)
1-3	679
4-9	640
10-29	600
30	585

[a] These costs are F.O.B., Pittsburgh, PA. They do not include taxes and freight charges.

onsite carbon replacement services, Calgon estimates carbon costs to be $2.00 to $2.50 per pound for virgin carbon and $1.50 to $1.80 for reactivated carbon. Annual maintenance costs are reported to range from 3% to 10% of the installed capital costs.

As fewer installation materials and labor are required to install a cannister unit than a fixed-bed system, the composite installation factor is consequently lower. The only costs required are those needed to place the cannisters at, and connect them to, the source. This involves a small amount of piping only; little or no electrical work, painting, foundations, or the like would be needed. Twenty percent of the sum of the cannister(s) cost, freight charges, and applicable sales taxes would cover this installation cost.

1.4 Estimating Total Annual Cost

As Section of this Manual explains, the total annual cost is comprised of three components: direct costs, indirect costs, and recovery credits. These will be considered separately.

1.4.1 Direct Annual Costs

These include the following expenditures: steam, cooling water, electricity, carbon replacement, operating and supervisor labor, and maintenance labor and materials. Of these, only electricity and solid waste disposal or carbon replacement/regeneration would apply to the cannister-type adsorbers.

1.4.1.1 Steam

As explained in Section 1.1, steam is used during the desorption cycle. The quantity of steam required will depend on the amount of carbon in the vessel, the vessel dimensions, the type and amount of VOC adsorbed, and other variables. Experience has shown that the steam requirement ranges from approximately 3 to 4 lbs of steam/lb of adsorbed VOC.[9][10] Using the midpoint of this range, we can develop the following expression for the annual steam cost:

$$C_s = 3.50 \times 10^{-3} \, m_{voc} \, \theta_s \, p_s \qquad (1.28)$$

where

Cs	=	steam cost ($/yr)
$_s$	=	system operating hours (h/yr)
m_{voc}	=	VOC inlet loading (lbs/h)
p_s	=	steam price ($/thous. lbs)

If steam price data are unavailable, one can estimate its cost at 120% of the fuel cost. For example, if the local price of natural gas were $5.00/million BTU, the estimated steam price would be $6.00/million BTU which is approximately $6.00/thousand lbs. (The 20% factor covers the capital and annual costs of producing the steam.)

1.4.1.2 Cooling Water

Cooling water is consumed by the condenser in which the steam-VOC mixture leaving the desorbed carbon bed is totally condensed. Most of the condenser duty is comprised of the latent heat of vaporization (ΔH_v) of the steam and VOC. As the VOC ΔH_v are usually small compared to the steam ΔH_v, (about 1000 BTU/lb), the VOC ΔH_v may be ignored. So may the sensible heat of cooling the water-VOC condensate from the condenser inlet temperature (about 212°F) to the outlet temperature. Therefore, the cooling water requirement is essentially a function of the steam usage and the allowable temperature rise in the coolant, which is typically 30° to 40°F.[9] Using the average temperature rise (35°F), we can write:

$$C_{cw} = 3.43 \frac{C_s}{p_s} p_{cw} \tag{1.29}$$

where

C_{cw} = cooling water cost ($/yr)
p_{cw} = cooling water price ($/thous. gal.)

If the cooling water price is unavailable, use $0.15 to $0.30/thousand gallons.

1.4.1.3 Electricity

In fixed-bed adsorbers, electricity is consumed by the system fan, bed drying/cooling fan, cooling water pump, and solvent pump(s). Both the system and bed fans must be sized to overcome the pressure drop through the carbon beds. But, while the system fan must continuously convey the total gas flow through the system, the bed cooling fan is only used during a part of the desorption cycle (one-half hour or less).

For both fans, the horsepower needed depends both on the gas flow and the pressure drop through the carbon bed. The pressure drop through the bed (ΔP_b) depends on several variables, such as the adsorption temperature, bed velocity, bed characteristics (e.g., void fraction), and thickness. But, for a given temperature and carbon, the pressure drop per unit thickness depends solely on the gas velocity. For instance, for Calgon's "PCB" carbon (4 x 10 mesh), the following relationship holds:[8]

$$\frac{\Delta P_b}{t_b} = 0.03679\ v_b + 1.107 \times 10^{-4}\ v_b^2 \tag{1.30}$$

where

$\Delta P_b/t_b$ = pressure drop through bed (inches of water/foot of carbon)
v_b = superficial bed velocity (ft/min)

As Equation 1.22 shows, the bed thickness (t_b, ft) is the quotient of the bed volume (V_b) and the bed cross-sectional area (A_b). For a 30 lb/ft^3 carbon bed density, this becomes:

$$t_b = \frac{V_b}{A_b} = \frac{0.0333\ M_c'}{A_b} \tag{1.31}$$

(For vertically erected vessels, $A_b = Q/v_b$, while for horizontally erected cylindrical vessels, A= LD.) Once ΔP_b is known, the system fan horsepower requirement (hp_{sf}) can be calculated:

$$hp_{sf} = 2.50 \times 10^{-4}\ Q\ \Delta P_s \tag{1.32}$$

where

Q = gas volumetric flow through system (acfm)
ΔP_s = total system pressure drop = $\Delta P_b + 1$

(The extra inch accounts for miscellaneous pressure losses through the external ductwork and other parts of the system.[9][6] However, if extra long duct runs and/or preconditioning equipment are needed, the miscellaneous losses could be much higher.)

This equation incorporates a fan efficiency of 70% and a motor efficiency of 90%, or 63% overall.

The horsepower requirement for the bed drying/cooling fan (hp_{cf}) is computed similarly. While the bed fan pressure drop would still be ΔP_b, the gas flow and operating times would be different. For typical adsorber operating conditions, the drying/cooling air requirement would be 50 to 150 ft^3/lb carbon, depending on the bed moisture content, required temperature drop, and other factors. The operating time (θ_{cf}) would be the product of the drying/cooling time per desorption cycle and the number of cycles per year. It can be shown that:

$$\theta_{cf} = 0.4\ \theta_D \left(\frac{N_A \theta_s}{\theta_A} \right) \tag{1.33}$$

(The "0.4" allows for the fact that as a rule-of-thumb, approximately 40% of the desorption cycle is used for bed drying/cooling.)

The cooling water pump horsepower requirement (hp_{cwp}) would be computed as follows:

$$hp_{cwp} = \frac{2.52 \times 10^{-4} q_{cw} H s}{\eta} \qquad (1.34)$$

where

q_{cw}	=	cooling water flow (gal/min)
H	=	required head (nominally 100 feet of water)
s	=	specific gravity of fluid relative to water at 60°F
η	=	combined pump-motor efficiency.

The annual operating hours for the cooling water pump (1_{cwp}) would be computed using Equation 1.33, after substituting "0.6" for 0.4. The 0.6 factor accounts for the fact that the cooling water pump is only used during the steaming portion of the regeneration, while the condenser is in operation.

Equation 1.34 may also be used to compute the solvent pump horsepower requirement. In the latter case, the flow (q_s) would be different of course, although the same head—100 ft. of water—could be used. The specific gravity would depend on the composition and temperature of the condensed solvent. For example, the specific gravity of toluene at 100°F would be approximately 0.86 at 70°F. (However, the solvent pump horsepower is usually very small—usually < 0.1 hp.—so its electricity consumption can usually be neglected.)

Once the various horsepowers are calculated, the electricity usage (in kWh) is calculated, by multiplying each horsepower value by 0.746 (the factor for converting hp to kilowatts) and the number of hours each fan or pump operates annually. For the system fan, the hours would be the annual operating hours for the system (θ_s). But, as discussed above, the operating times for the bed drying/cooling fan and cooling water pump would be different.

To obtain the annual electricity cost, simply multiply kWh by the electricity price (in $/kWh) that applies to the facility being controlled.

For cannister units, use Equation 1.32 to calculate the fan horsepower requirement. However, instead of P_b use the following to compute the total cannister pressure drop P_e inches of water:[7]

$$\Delta P_c = 0.0471 Q_c + 9.29 \times 10^{-4} Q_c^2$$

[6] To obtain a more precise estimate of ductwork pressure drop, refer to Section 2 of this Manual.

where

Q_c = flow through the cannister (acfm).

1.4.1.4 Carbon Replacement

As discussed above, the carbon has a different economic life than the rest of the adsorber system. Therefore, its replacement cost must be calculated separately. Employing the procedure detailed in Section 1, we have:

$$\Delta P_c = 0.0471\ Q_c + 9.29 \times 10^{-4}\ Q_c^2 \qquad (1.35)$$

where

CRF_c = capital recovery factor for the carbon
1.08 = taxes and freight factor
C_c, C_{cl} = initial cost of carbon (F.O.B. vendor) and carbon replacement labor cost, respectively ($)

The replacement labor cost covers the labor cost for removing spent carbon from vessels and replacing it with virgin or regenerated carbon. The cost would vary with the amount of carbon being replaced, the labor rates, and other factors. For example, to remove and replace a 50,000 pound carbon charge would require about 16 person-days, which, at typical wage rates, is equivalent to approximately $0.05/lb replaced.[13]

A typical life for the carbon is five years. However, if the inlet contains VOCs that are very difficult to desorb, tend to polymerize, or react with other constituents, a shorter carbon lifetime—perhaps as low as *two* years—would be likely.[1] For a five-year life and 7% interest rate, CRF_c = 0.2439.

1.4.1.5 Solid Waste Disposal

Disposal costs are rarely incurred with fixed-bed adsorbers, because the carbon is almost always regenerated in place, not discarded. The carbon in cannister units should also be regenerated in most cases. For larger vessels, common practice is for a carbon vendor to pick up the spent carbon and replace it with fresh carbon. The spent carbon is then returned to a central facility for regeneration. EPA encourages both solvent recovery and reuse of spent carbon as pollution prevention and waste minimzation techniques.

In some cases, the nature of the solvents, including their extremely hazardous nature or the difficulty in desorbong them from the carbon may make disposal the preferred option. In these cases, an entire cannister—carbon, drum, connections, etc.— may be shipped to a secure landfill. The cost of landfill disposal could vary considerably, depending on the number of cannisters disposed of, the location of the landfill, etc. Based on data obtained from two large landfills, for instance, the disposal cost would range from approximately $35 to $65 per cannister excluding transportation costs.[14][15]

1.4.1.6 Operating and Supervisory Labor

The operating labor for adsorbers is relatively low, as most systems are automated and require little attention. One-half operator hour per shift is typical.[12] The annual labor cost would then be the product of this labor requirement and the operating labor wage rate ($/h) which, naturally, would vary according to the facility location, type of industry, etc. Add to this 15% to cover supervisory labor, as Section 1.2 suggests.

1.4.1.7 Maintenance Labor and Materials

Use 0.5 hours/shift for maintenance labor [12] and the applicable maintenance wage rate. If the latter data are unavailable, estimate the maintenance wage rate at 110% of the operating

labor rate, as Section 1 suggests. Finally, for maintenance materials, add an amount equal to the maintenance labor, also per Section 1.

1.4.2 Indirect Annual Costs

These include such costs as capital recovery, property, taxes, insurance, overhead, and administrative costs ("G&A"). The capital recovery cost is based on the equipment lifetime and the annual interest rate employed. (See Section 1.2 for a thorough discussion of the capital recovery cost and the variables that determine it.) For adsorbers, the system lifetime is typically ten years, except for the carbon, which, as stated above, typically needs to be replaced after five years. Therefore, when figuring the system capital recovery cost, one should base it on the installed capital cost less the cost of replacing the carbon (i.e., the carbon cost plus the cost of labor necessary to replace it). Substituting the initial carbon and replacement labor costs from Equation 1.36, we obtain:

$$CRC_s = \left[TCI - \left(1.08C_c + C_{cl}\right)\right] CRF_s \qquad (1.37)$$

where

CRC_s	=	capital recovery cost for adsorber system ($/yr)
TCI	=	total capital investment ($)
1.08	=	taxes and freight factor
C_c, C_{cl}	=	initial carbon cost (F.O.B. vendor) and carbon replacement cost, respectively ($)
CRF_s	=	capital recovery factor for adsorber system (defined in Section 1.2).

For a ten-year life and a 7% annual interest rate, the CRF_s would be 0.1424.

As Section 1.2 indicates, the suggested factor to use for property taxes, insurance, and administrative charges is 4% of the *TCI*. Finally, the overhead is calculated as 60% of the sum of operating, supervisory, and maintenance labor, and maintenance materials.

The above procedure applies to cannister units as well, except in those case where the entire unit and not just the carbon is replaced. The piping and ducting cost can usually be considered a capital investment with a useful life of ten years. However, whether the cannister itself would be treated as a capital or an operating expense would depend on the particular application and would need to be evaluated on a case-by-case basis.

1.4.3 Recovery Credits

These apply to the VOC which is adsorbed, then desorbed, condensed, and separated from the steam condensate. If the recovered VOC is sufficiently pure, it can be sold. However, if the VOC layer contains impurities or is a mixture of compounds, it would require further treatment, such as distillation. Purification and separation costs are beyond the scope of this chapter. Needless to say, the costs of these operations would offset the revenues generated by the sale of the VOC. Finally, as an alternative to reselling it, the VOC could be burned as fuel and valued accordingly. In any case, the following equation can be used to calculate these credits:

$$RC = m_{voc}\ \theta_s\ P_{voc}\ E \tag{1.38}$$

where

RC	=	recovery credit ($/yr)
m_{voc}	=	VOC inlet loading (lbs/h)
θ_s	=	system operating hours (h/yr)
p_{voc}	=	resale value of the recovered VOC ($/lb)
E	=	adsorber VOC control efficiency

By definition, the efficiency (E) is the difference between the inlet and outlet VOC mass loading, divided by the inlet loading. However, during an adsorption cycle the outlet VOC loading will increase from essentially zero at the start of the cycle to the breakthrough concentration at the end of the cycle. Because the efficiency is a function of time, it should be calculated via integration over the length of the adsorption cycle. To do this would require knowledge of the temporal variation of the outlet loading during the adsorption cycle. If this knowledge is not available to the Manual user, a conservative approximation of the efficiency may be made by setting the outlet loading equal to the breakthrough concentration.

1.4.4 Total Annual Cost

Finally, as explained in Section 1, the total annual cost (TAC) is the sum of the direct and indirect annual costs, less any recovery credits, or:

$$TAC = DC + IC - RC \tag{1.39}$$

1.4.5 Example Problem

A source at a printing plant emitting 100 lb/h of toluene is to be controlled by a carbon adsorber. The plant proposes to operate the adsorber in a continuous mode for 8,640 h/yr (360 days). While operating, two carbon beds will be adsorbing, while a third will be desorbing/on stand by. For its convenience, the plant has selected adsorption and desorption times of 12 and 5 hours, respectively. The total waste gas flow is 10,000 acfm at the adsorber inlet conditions (one atmosphere and 77°F). The waste gas contains negligible quantities of particulate matter and moisture. Further, the applicable VOC regulation requires the adsorber to achieve a mean removal efficiency of 98% during the entire adsorption cycle. Finally, assume that the recovered toluene is recycled at the source. Estimate the total capital investment and total annual cost for the adsorber system.

Carbon Working Capacity: At the stated flow and pollutant loading, the toluene inlet concentration is 710 ppm. This corresponds to a partial pressure of 0.0104 psia. Substituting this partial pressure and the toluene isotherm parameters (from Table 1) into Equation 1, we obtain an equilibrium capacity of 0.333 lb/lb. By applying the rule-of-thumb discussed above (page 4-19), we obtain a working capacity of 0.167 lb/lb (*i.e.*, 0.333/2).

Carbon Requirement: As stated above, this adsorber would have two beds on-line and a third off-line. Equation 3.12 can answer this question. Substitution of the adsorption time and numbers of adsorbing and desorbing beds yields:

$$Desorption\ time = \theta_D \leq \theta_A \left(\frac{N_D}{N_A} \right) = 12h \times 0.5 = 6h$$

Because the stated desorption time (5 hours) is less than 6 hours, the proposed bed configuration is feasible. Next, calculate the carbon requirement (M_c) from Equation 3.14:

$$M_c = \frac{m_{voc}}{w_c}\theta_A\left(1+\frac{N_D}{N_A}\right) = \frac{100\,lb/h \times (12h) \times (1+0.5)}{0.167\,lb/lb} = 10{,}800lbs$$

From Equation 3.16, the carbon cost is:

$$C_c = 1.00\ M_c = \$10{,}800$$

Adsorber Vessel Dimensions: Assume that the vessels will be erected horizontally and select a superficial bed velocity (v_b) of 75 ft/min. Next, calculate the vessel diameter (D), length (L), and surface area (S) from Equations 3.18, 3.19, and 3.24, respectively. [Note: In these equations, $M_c = M_c\,(N_A + N_D) = 3{,}600$ lb and $Q = Q/N_A = 5{,}000$ acfm.]

$$D = \frac{0.127\ M_c\ v_b}{Q'} = \frac{0.127\ (3{,}600)\ (75)}{5{,}000} = 6.86\ ft$$

$$L = \frac{7.87}{M_c'}\left(\frac{Q'}{v_b}\right)^2 = \frac{7.87}{3{,}600}\left(\frac{5{,}000}{75}\right)^2 = 9.72\ ft$$

$$S = \pi\ D\left(L + \frac{D}{2}\right) = 283\ ft$$

Because S falls between 97 and 2,110 ft^2, Equation 1.25 can be used to calculate the cost per vessel, C_v (assuming 304 stainless steel construction). Thus:

$$C_v = 271\ S^{0.778} = \$21{,}900$$

Adsorber Equipment Cost: Recall that the adsorber equipment cost is comprised of the adsorber vessels, carbon, and the condenser, decanter, fan, pumps and other equipment usually included in the adsorber price. The cost of the latter items are "factored" from the combined cost of the vessels and carbon. Combining Equations 1.26 and 1.27, we have:

$$C_A = 5.82\ Q^{-.0133}\left[C_c + \left(N_A + N_D\right) C_v\right]$$

Substitution of the above values yields:

$$C_A = \$130{,}800$$

Cost of Auxiliary Equipment: Assume that costs for the following auxiliary equipment have been estimated from data in other parts of the Manual:

Ductwork	$16,500
Dampers	7,200
Stack	8,500
Total	$32,000

Total Capital Investment: The total capital investment is factored from the sum of the adsorber unit and auxiliary equipment cost, as displayed in Table 1.5. Note that no line item cost has been shown for instrumentation, for this cost is typically included in the adsorber price.

Therefore:

Purchased Equipment Cost = "B" = 1.08 x "A"= 1.08 x ($130,800 + $32,200) = $176,040

And:

Total Capital Investment (rounded) = 1.61 x "B" = $283,400

Annual Costs: Table 1.5 gives the direct and indirect annual costs for the carbon adsorber system, as calculated from the factors in Section 1.1.4. Except for electricity, the calculations in the table show how these costs were derived. The following discussion will deal with the electricity cost.

First, recall that the electricity includes the power for the system fan, bed drying/cooling fan, and the cooling water pump. (The solvent pump motor is normally so small that its power consumption may be neglected.) These consumptions are calculated as follows:

System fan is calculated from Equation 1.32:

$$kWh_{sf} = 0.746\,kW/hp \times 250 \times 10^{-4}\ Q\ \Delta P_s\ \theta_s$$

Table 1.5: Capital Cost Factors for Carbon Adsorber System Example Problem

Cost Item	Factor
Direct Costs	
Purchased equipment costs	
Adsorber vessels and carbon	$130,800
Auxiliary equipment	32,200
Sum = A	$163,000
Instrumentation, 0.1 A	
Sales taxes, 0.03 A	4,890
Freight, 0.05 A	8,150
Purchased equipment cost, PEC	$176,040
Direct installation costs	
Foundations & supports, 0.08 B	14,083
Handling & erection, 0.14 B	24,646
Electrical, 0.04 B	7,042
Piping, 0.02 B	3,521
Insulation for ductwork, 0.01 B	1,760
Painting, 0.02 B	1,760
Direct installation costs	$52,812
Site preparation	
Buildings	
Total Direct Costs, DC	$228,852
Indirect Costs (installation)	
Engineering, 0.10 B	17,604
Construction and field expenses, 0.05 B	8,802
Contractor fees, 0.10 B	17,604
Start-up, 0.02 B	3,521
Performance test, 0.01 B	1,760
Contingencies, 0.03 B	5,281
Total Indirect Costs, IC	$54,572
Total Capital Investment (rounded)	$283,400

The cost for this is included in the adsorber equipment cost.

But:

$$\Delta P_s\ (inches\ water) = \Delta P_s + 1 = t_b \left(0.03679\ v_b + 1.107 \times 10^{-4}\ v_b^2\right) + 1$$

(The latter expression was derived from Equation 1.30, assuming that the carbon used in this example system is Calgon's "PCB", 4 x 10 mesh size.)

By assuming a carbon bed density, of 30 lb/ft^3, Equation 1.31 can be used to calculate the bed thickness (t_b):

$$Bed\ thickness = t_b = \frac{0.0333\ M_c'}{A_b} = \frac{0.0333\ M_c'}{LD} = 1.80\ ft$$

Thus:

$$\Delta P_s = 1 + 1.80 \left(0.03679 \times 75 + 1.107 \times 10^{-4} \times 75^2\right) = 7.09\ inches$$

and finally:

$$kWh_{sf} = 0.746 \times 2.5 \times 10^{-4} \times 7.09 in. \times 10{,}000\ acfm \times 8{,}640 h/yr$$
$$= 114{,}200\ kWh/yr$$

Bed drying/cooling fan: During the drying/cooling cycle, the pressure drop through the bed also equals P_b. However, as Section 1.4.1.3 indicates, the flow and operating time are different. For the air flow, take the midpoint of the range (100 ft^3 air/lb carbon) and divide by 2 hours (the bed drying/cooling time), yielding: 100 ft^3/lb x 3,600 lbs x 1/120 min = 3,000 acfm. Substituting this into Equation 1.32 results in:

$$2.50 \times 10^{-4} \times 7.09\ inches \times 3{,}000\ acfm = 5.32\ hp$$

From Equation 1.33, we get:

$$\Theta_{cf} = 0.4 \times 5h \times 2 \times \frac{8{,}640h}{12h} = 12h$$

Thus:

$$kWh_{cf} = 0.746\,kW/hp \times 5.32hp \times 2{,}880h = 11{,}400\,kWh/yr$$

Cooling water pump: The cooling water pump horsepower is calculated from Equation 1.34. Here, let = 63% and H = 100 ft. The cooling water flow (q_{cw}) is the quotient of the annual cooling water requirement and the annual pump operating time. From the data in Table 1.6, we obtain the cooling water requirement: 10,400,000 gal/yr. The pump annual operating time is obtained from Equation 1.33 (substituting 0.6 for 0.4), or cwp = (0.6)(5 h)(2)(8,640)/12 = 4,320 h/yr.

Thus:

$$hp_{cwp} = \frac{(2.52 \times 10^{-4})\,(100\,ft)}{0.63} \times \frac{10{,}400{,}000\,gal/yr}{4{,}320\,h/yr \times 60\,min/yr} = 1.60hp$$

Thus:

$$hp_{cwp} = \frac{(2.52 \times 10^{-4})\,(100\,ft)}{0.63} \times \frac{10{,}400{,}000\,gal/yr}{4{,}320\,h/yr \times 60\,min/yr} = 1.60hp$$

And:

$$kWh_{cwp} = 0.746\,kW/h \times 1.60hp \times 4{,}320\,h/yr = 5{,}160\,kWh/yr$$

Summing the individual power consumptions, we get the value shown in Table 1.5: 131,000kWh/yr

Recovery Credit: As Table 3.6 indicates, a credit for the recovered toluene has been taken. However, to account for miscellaneous losses and contamination, the toluene is arbitrarily valued at one-half the November 1989 market price of $0.0533/lb(= $111/ton).[16]

Total Annual Cost: The sum of the direct and indirect annual costs, less the toluene recovery credit, yields a net total annual cost of $76,100. Clearly, this "bottom line" is very sensitive to the recovery credit and, in turn, the value given the recovered toluene. For instance, if it had been valued at the full market price ($221/ton), the credit would have doubled and the total annual cost would have been $29,200. Thus when incorporating recovery credits, it is imperative to select the value of the recovered product carefully.

Table 1.6: Annual Costs for Carbon Adsorber System Example Problem

Cost Item		Calculations	Cost
Direct Annual Costs, DC			
	Operating Labor		
	Operator	0.5h/shift x 3 shifts/day x 360 days/yr x $12/hr	$6,480
	Supervisor	15% of operator = 0.15 x 6,480	970
	Operating materials		
	Maintenance		
	Labor	0.5h/shift x 3 shifts/day x 360 days/yr x $13.20/hr	7,130
	Material	100% of maintenance labor	7,130
	Replacement Parts, Carbon	(5-year life)	
	Replacement Labor	0.2439 ($0.05/lb x 10,800 lb)	130
	Carbon Cost[a]	0.2439 ($21,600 x 1.08)	5,690
	Utilities		
	Electricity	$0.06 kWh x 131,000 kWh/yr	7,860
	Steam	3.5 lb/lb VOC x $6/103 x 100 lb VOC/hr x 8,640 hr/yr	18,140
	Cooling Water	3.43 gal/lb steam x [(3.5 x 100 x 8,640) lb steam x $0.20/103 gal.] / yr	2,070
Total DC			$55,600
Indirect Annual Costs, IC			
	Overhead	60% of sum of operating labor, maintenance labor, & maintenance materials = 0.6 (6,480 + 970 + 7,130 + 7,130)	13,030
	Administrative charges	2% of Total Capital Investment = 0.02 ($316,000)	6,320
	Property tax	1% of Total Capital Investment = 0.01 ($316,000)	3,160
	Insurance	1% of Total Capital Investment = 0.01 ($316,000)	3,160
	Capital recovery[b]	0.1424 [316,000 - 0.05 (10,800) - 1.08 (21,600)]	41,600
Total IC			$67,270
Recovery Credit (toluene)			(46,820)
Total Annual Cost (rounded)			$76,100

[a] The 1.08 factor is for freight and sales taxes.

[b] The capital recovery cost factor, CRF, is a function of the adsorber or equipment life and the opportunity cost of the capital (i.e., interest rate). For example, for a 10 year equipment life and a 7% interest rate, CRF = 0.1424.

References

[1 Correspondence: Robert L. Stallings and William Klotz (Research Triangle Institute, Research Triangle Park, NC) to William M. Vatavuk (U.S. EPA, OAQPS, Research Triangle Park, NC), June 24, 1986.

[2] Calvert, Seymour and Englund, Harold M. (eds.), *Handbook of Air Pollution Control Technology*, John Wiley & Sons, New York, 1984, pp. 135-192.

[3] *Handbook of Chemistry and Physics*, 54th Edition, The Chemical Rubber Company, Cleveland, 1973-74, pp. D85-D92.

[4] Correspondence: Wayne G. Schulinger and Georgiana Riley (TIGG Corporation, Pittsburgh, PA) to Janet Groeber (SAIC, Cincinnati, OH), October11,1 999.

[5] Correspondence: Mark H. Stenzel (Calgon Carbon Corporation, Pittsburgh, PA) to Janet Groeber (SAIC, Cincinnati, OH), August 24,1 999.

[6] "Calgon Ventsorb® for Industrial Air Purification" (Bulletin 23-56a), Calgon Corporation, Pittsburgh, 1986.

[7] *Adsorption Handbook*, Calgon Corporation, Pittsburgh, 1980.

[8] Rogers, Tony, "Comparison of BED SIZE and Calgon Adsorption Isotherms", Research Triangle Institute (Research Triangle Park, NC), January 20, 1988.

[9] Correspondence: Richard Selznick (Baron Blakeslee, Inc., Westfield. NJ) to William M. Vatavuk (U.S. EPA, OAQPS, Research Triangle Park, NC), April 23, 1986.

[10] Correspondence: Denny Clodfelter(M&W Industries, Inc., Rural Hall, NC) to William M. Vatavuk (U.S. EPA, OAQPS, Research Triangle Park, NC), September 25, 1989.

[11] Matley, Jay (ed.), *Modern Cost Engineering*, McGraw-Hill Publications Co., New York, 1984, p. 142.

[12] Vatavuk, William M. and Neveril, Robert, "Estimating Costs of Air Pollution Control Systems, Part II: Factors for Estimating Capital and Operating Costs," *Chemical Engineering*, November 3, 1980, pp. 157-162.

[13] Telephone conversation: Robert L. Stallings (Research Triangle Institute, Research Triangle Park, NC) with William M. Vatavuk (U.S. EPA, OAQPS, Research Triangle Park, NC), September 11, 1986.

[14] Correspondence: William Kitto (Chemwaste, Sulphur, LA) to William M. Vatavuk (U.S. EPA, OAQPS, Research Triangle Park, NC), July 25, 1986.

[15] Correspondence: Jerry Locklear (GSX, Pinewood, SC) to William M. Vatavuk (U.S. EPA, OAQPS, Research Triangle Park, NC), July 25, 1986.

[16] *Chemical Marketing Reporter*, December 2, 1989.

EPA/452/B-02-001

Section 3

VOC Controls

EPA/452/B-02-001

Section 3.1

VOC Recapture Controls

EPA/452/B-02-001

Chapter 2

Refrigerated Condensers

Gunseli Sagun Shareef
Wiley J. Barbour
Susan K. Lynch
W. Richard Pelt
Radian Corporation
Research Triangle Park, NC 27709

William M. Vatavuk
Innovative Strategies and Economics Group, OAQPS
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

December 1995

Contents

2.1 Introduction 2-3
2.1.1 System Efficiencies and Performance 2-3

2.2 Process Description 2-3
2.2.1 VOC Condensers 2-5
2.2.2 Refrigeration Unit 2-6
2.2.3 Auxiliary Equipment 2-7

2.3 Design Procedures 2-7
2.3.1 Estimating Condensation Temperature 2-9
2.3.2 VOC Condenser Heat Load 2-10
2.3.3 Condenser Size 2-13
2.3.4 Coolant Flow Rate 2-14
2.3.5 Refrigeration Capacity 2-15
2.3.7 Auxiliary Equipment 2-15
2.3.8 Alternate Design Procedure 2-16

2.4 Estimating Total Capital Investment 2-16
2.4.1 Equipment Costs for Packaged Solvent Vapor Recovery Systems 2-17
2.4.2 Equipment Costs for Nonpackaged (Custom) Solvent Vapor Recovery Systems 2-20
2.4.3 Equipment Costs for Gasoline Vapor Recovery Systems 2-21
2.4.4 Installation Costs 2-22

2.5 Estimating Total Annual Cost 2-26
2.5.1 Direct Annual Costs 2-26
2.5.2 Indirect Annual Costs 2-27
2.5.3 Recovery Credit 2-28

2.6 Example Problem 1 2-29
2.6.3 Equipment Costs 2-33
2.6.4 Total Annual Cost 2-34

2.7 Example Problem 2 2-36
2.7.1 Required Information for Design 2-36

2.8 Acknowledgments 2-36

References 2-37

Appendix A 2-39

Appendix B 2.42

2.1 Introduction

Condensers in use today may fall in either of two categories: refrigerated or non-refrigerated. Non-refrigerated condensers are widely used as raw material and/or product recovery devices in chemical process industries. They are frequently used prior to control devices (e.g., incinerators or absorbers). Refrigerated condensers are used as air pollution control devices for treating emission streams with high VOC, concentrations (usually > 5,000 ppmv) in applications involving gasoline bulk terminals, storage, etc.

Condensation is a separation technique in which one or more volatile components of a vapor mixture are separated from the remaining vapors through saturation followed by a phase change. The phase change from gas to liquid can be achieved in two ways: (a) the system pressure can be increased at a given temperature, or (b) the temperature may be lowered at a constant pressure. In a two-component system where one of the components is noncondensible (e.g., air), condensation occurs at dew point (saturation) when the partial pressure of the volatile compound is equal to its vapor pressure. The more volatile a compound (e.g., the lower the normal boiling point), the larger the amount that can remain as vapor at a given temperature; hence the lower the temperature required for saturation (condensation). Refrigeration is often employed to obtain the low temperatures required for acceptable removal efficiencies. This chapter is limited to the evaluation of refrigerated condensation at constant (atmospheric) pressure.

2.1.1 System Efficiencies and Performance

The removal efficiency of a condenser is dependent on the emission stream characteristics including the nature of the VOC in question (vapor pressure/temperature relationship), VOC concentration, and the type of coolant used. Any component of any vapor mixture can be condensed if brought to a low enough temperature and allowed to conic to equilibrium. Figure 2.1 shows the vapor pressure dependence on temperature for selected compounds.[1] A condenser cannot lower the inlet concentration to levels, below the saturation concentration at the coolant temperature. Removal efficiencies above 90 percent can be achieved with coolants such as chilled water, brine solutions, ammonia, or chlorofluorocarbons, depending on the VOC composition and concentration level of the emission stream.

2.2 Process Description

Figure 2.2 depicts a typical configuration for a refrigerated surface condenser system as an emission control device. The basic equipment required for a refrigerated condenser system includes a VOC condenser, a, refrigeration unit(s). and auxiliary, equipment (e.g., precooler, recovery/ storage tank, pump/blower, and piping).

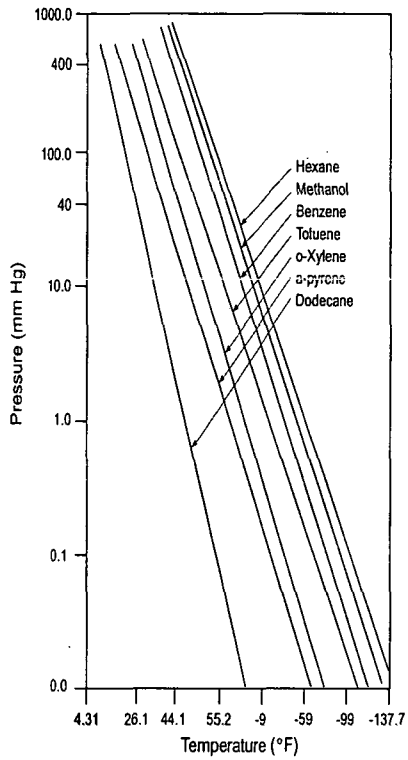


Figure 2.1: Vapor Pressures of Selected Compounds vs. Temperature [1]

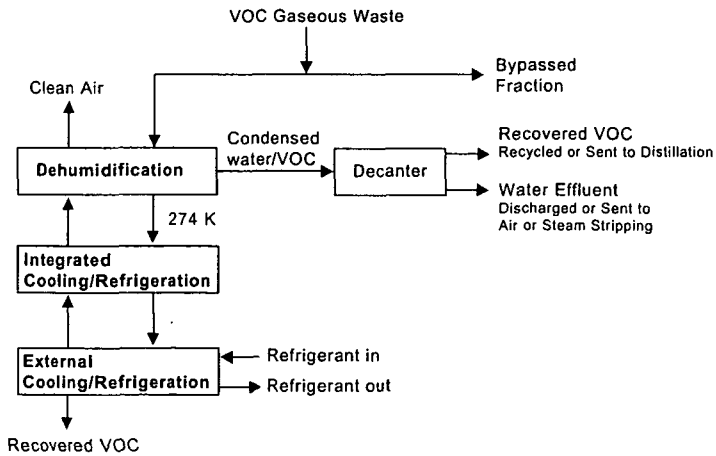


Figure 2.2: Schematic Diagram for a Refrigerated Condenser System

2.2.1 VOC Condensers

The two most common types of condensers used are surface and contact condensers.[21] In surface condensers, the coolant does not contact tile gas stream. Most surface condensers in refrigerated systems are the shell and tube type (see Figure 2.3).[3] Shell and tube condensers circulate the coolant through tubes. The VOCs condense on the outside of the tubes (e.g., within the shell). Plate and frame type heat exchangers are also used as condensers in refrigerated systems. In these condensers, the coolant and the vapor flow separately over thin plates. In either design, the condensed vapor forms a film on the cooled surface and drains away to a collection tank for storage, reuse, or disposal.

In contrast, to surface condensers where the coolant does not contact either the vapors or the condensate, contact condensers cool tile vapor stream by spraying either a liquid at ambient temperature or a chilled liquid directly into the gas stream.

Spent coolant containing the VOCs from contact condensers usually can not be reused directly and can be a waste disposal problem. Additionally, VOCs in the spent coolant can not be directly recovered without further processing. Since the coolant from surface condensers does not contact the vapor stream, it is not contaminated and can be recycled in a closed loop. Surface condensers also allow for direct recovery of VOCs from tile gas Stream. This chapter addresses the design and costing of refrigerated surface condenser systems only.

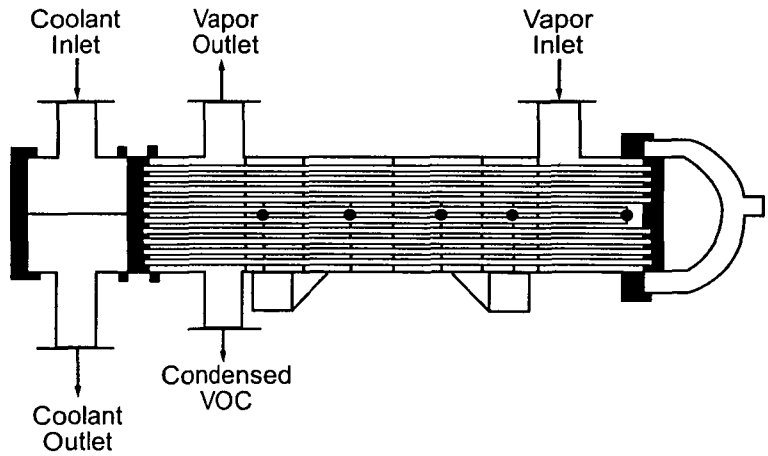


Figure 2.3: Schematic Diagram for a Refrigerated Condenser System

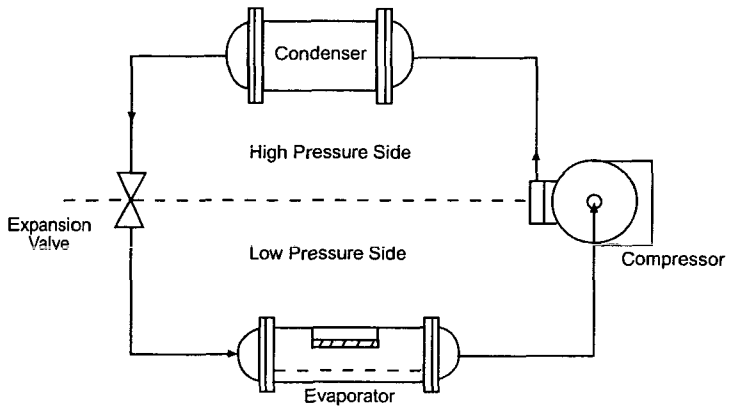


Figure 2.4: Basic Refrigeration Cycle [4]

2.2.2 Refrigeration Unit

The commonly used mechanical vapor compression cycle to produce refrigeration consists of four stages: evaporation, compression, condensation, and expansion (see Figure 2.4).[4] Tile cycle which is used for single-stage vapor compression involves two pressures, high and low, to enable a continuous process to produce a cooling effect. Heat absorbed from tile gas stream evaporates the liquid coolant (refrigerant). Next, the, refrigerant (now in vapor phase) is compressed to a higher temperature and pressure by tile system compressor. Then, the superheated refrigerant vapor is condensed, rejecting its sensible and latent heat in the condenser. Subsequently, the liquid refrigerant flows from the condenser through the expansion valve, where pressure and temperature are reduced to those in the evaporator, thus completing the cycle.

The capacity of a refrigeration unit is the rate at which heat is removed, expressed in tons of refrigeration. One ton of refrigeration is the refrigeration produced by melting one ton of ice at 32°F in 24 hours. It is the rate of removing heat equivalent to 12,000 Btu/h or 200 Btu/min. For more details on refrigeration principles, see References [5] and [6].

Applications requiring low temperatures (below about -30°F), multistage refrigeration systems are frequently employed.[4] Multistage systems are designed and marketed in two different types-compound and cascade. In compound systems, only one refrigerant is used. In a cascade system, two or more separate refrigeration systems are interconnected in such a manner that one provides a means of heat rejection for the other. Cascade systems are desirable for applications requiring temperatures between -50 and -150°F and allow the use of different refrigerants in each cycle,[4] Theoretically, any number of cascaded stages are possible, each stage requiring an additional condenser and an additional stage of compression.

In refrigerated condenser systems, two kinds of refrigerants are used, primary and secondary. Primary refrigerants are those that undergo a phase change from liquid to gas after absorbing heat. Examples are ammonia (R-717), and chlorofluorocarbons such as chlorodifluoromethane (R-22) or dichlorodifluorormethane (R-12). Recent concerns about the latter causing depletion of the ozone layer is prompting development of substitute refrigerants. Secondary refrigerants such as brine solutions act only as heat carriers and remain in liquid phase.

Conventional systems use a closed primary refrigerant loop that cools the secondary loop through the heat transfer medium in the evaporator. The secondary heat transfer fluid is then pumped to a VOC vapor condenser where it is used to cool the, air/VOC vapor stream. In some applications. however, the primary refrigeration fluid is directly used to cool the vapor stream.

2.2.3 Auxiliary Equipment

As shown in Figure 2.2, some applications may require auxiliary equipment such as precoolers, recovery/storage tanks, pumps/blowers, and piping.

If water vapor is present in the treated gas stream or if the VOC has a high freezing point (e.g., benzene), ice or frozen hydrocarbons may form on the condenser tubes or plates. This will reduce the heat transfer efficiency of the condenser and thereby reduce the removal efficiency. Formation of ice will also increase the pressure drop across the condenser. In such cases, a precooler may be needed to condense the moisture prior to the VOC condenser. This precooler would bring the temperature of the stream down to approximately 35 to 40°F, effectively removing the moisture from the gas. Alternatively, an intermittent heating cycle can be used to melt away ice build-up. This may be accomplished by circulating ambient temperature brine through the condenser or by the use of radiant heating coils. If a system is not operated continuously, the ice can also be removed by circulating ambient air.

A VOC recovery tank for temporary storage of condensed VOC prior to reuse, reprocessing, or transfer to a larger storage tank may be necessary in some cases. Pumps and blowers are typically used to transfer liquid (e.g., coolant or recovered VOC) and gas streams, respectively, within the system.

2.3 Design Procedures

In this section are presented two procedures for designing (sizing) refrigerated surface condenser systems to remove VOC from air/VOC mixtures. With the first procedure presented, one calculates the condenser exit temperature needed to obtain a given VOC recovery efficiency. In the second procedure, which is the inverse of the first, the exit temperature is given and the recover efficiency corresponding to it is calculated.

The first procedure depends on knowledge of the following parameters:

1. Volumetric flow rate of the VOC-containing gas stream;
2. Inlet temperature of the gas stream;
3. Concentration and composition of the VOC in the gas stream;
4. Required removal efficiency of the VOC;
5. Moisture content of the emission stream; and
6. Properties of the VOC (assuming the VOC is a pure compound):
   - Heat of condensation,
   - Heat capacity, and
   - Vapor pressure.

The design of a refrigerated condenser system requires determination of the VOC condenser size and the capacity of the refrigeration unit. For a given VOC removal efficiency, the condensation temperature and the heat load need to be calculated to determine these parameters. The data necessary to perform the sizing procedures below as well as the variable names and their respective units are listed in Table 8.1.

Table 8.1: Required Input Data

Data	Variable Name	Units
Inlet Stream Flow Rate	Q_{in}	SCFM (77°F:1 atm)
Inlet Stream Temperature	T_{in}	°F
VOC Inlet Volume Fraction	Y_{voc},in	volume fraction
Required VOC Removal Efficiency	n	-
Antoine Equation Constants[a]	A,B,C	Btu/lb-mole
Heat of Condensation of the VOC[a]	deltaH	Btu/lb-mole-°F
Heat Capacity of the VOC[a]	$C_{p,voc}$	Btu/lb-°F
Specific Heat of the Coolant	$C_{p,cool}$	Btu/lb-mole-°F
Heat Capacity of Air	$C_{p,air}$	

[a]See Appendix A for these properties of selected organic compounds.

The steps outlined below for estimating condensation temperature and the heat load apply to a two-component mixture (VOC/air) in which one of the two components is considered to be noncondensible (air). The VOC component is assumed to consist of a single compound. Also, the emission stream is assumed to be free of moisture. The calculations are based on the assumptions of ideal gas and ideal solution to simplify the sizing procedures. For a more rigorous analysis, See Reference [5].

2.3.1 Estimating Condensation Temperature

The temperature necessary to condense the required amount of VOC must, be estimated to determine the heat load. The first step is to determine the VOC concentration at the outlet of the condenser for a given removal efficiency. This is calculated by first determining the partial pressure of the VOC at the outlet, P_{voc}. Assuming that the ideal gas law applies, P_{voc} is given by:

$$P_{VOC}\ (outlet) = 760\ \frac{Moles\ VOC\ in\ outlet\ stream}{Moles\ inlet\ stream\ -\ Moles\ VOC\ removed} \qquad (2.1)$$

where

P_{VOC} = Partial pressure of the VOC in the exit stream (mm Hg).

and the condenser is assumed to operate at a constant pressure of one atmosphere (760 mm Hg).

However:

Moles VOC in outlet stream = (Moles VOC in inlet stream)(1-η) (2.2)
Moles VOC in inlet stream = (Moles in inlet stream) $y_{voc,in}$ (2.3)
Moles VOC removed = (Moles VOC in inlet stream) (2.4)
where

η = removal efficiency of the condenser system (fractional)
= Moles VOC removed/Moles VOC in inlet
$y_{voc,in}$ = Volume fraction of VOC in inlet stream

After substituting these variables in Equation 2.1, we obtain:

$$P_{VOC} = 760\left[\frac{y_{VOC,\,in}\ (1\ -\ \eta)}{\left[1\ -\ \left(\eta\ y_{VOC,\,in}\right)\right]}\right] \qquad (2.5)$$

At the condenser outlet, the VOC in the gas stream is assumed to be at equilibrium with the V0C condensate. At equilibrium, the partial pressure of the VOC in the gas stream is equal to its vapor pressure at that temperature assuming the condensate is pure VOC (e.g., vapor pressure P_{voc}). Therefore, by determining the temperature at which this condition occurs, the condensation temperature can be specified. This calculation is based on the Antoine equation that defines the relationship between vapor pressure and temperature for a particular compound:

$$\log P_{VOC} = A - \frac{B}{T_{con} + C} \qquad (2.6)$$

where T_{con} is the condensation temperature (°C). Note that T_{con} is in degrees Centigrade in this equation. In Equation 2.6, *A, B,* and *C* are VOC-specific constants pertaining to temperature expressed in °C and pressure in mm Hg (see Appendix 8A). Solving for T_{con} and converting to degrees Fahrenheit:

$$T_{con} = \left(\frac{B}{A - \log_{10}(P_{VOC})} - C \right) 1.8 + 32 \qquad (2.7)$$

The calculation methods for a gas stream containing multiple VOCs are complex, particularly when there are significant departures from the ideal behavior of gases and liquids. However, the temperature necessary for condensation of a mixture of VOCs can be estimated by the weighted average of the temperatures necessary to condense each VOC in the gas stream at a concentration equal to the total VOC concentration.[1]

2.3.2 VOC Condenser Heat Load

Condenser heat load is the amount of heat that must be removed from the inlet stream to attain the specified removal efficiency. It is determined from an energy balance, taking into account the enthalpy change due to the temperature change of the VOC, the enthalpy change due to the condensation of the VOC, and the enthalpy change due to the temperature change of the air. Enthalpy change due to the presence of moisture in the inlet gas stream is neglected in the following analysis.

For the purpose of this estimation, it is assumed that, the total heat load on the system is equal to the VOC Condenser heat load. Realistically, when calculating refrigeration capacity requirements

for low temperature cooling units, careful consideration should be given to the, process line, losses and heat input of the process pumps. Refrigeration unit capacities are typically rated in terms of net output and do not reflect any losses through process pumps or process lines.

First, the number of lb-moles of VOC per hour in the inlet stream must be calculated by the following expression:

$$M_{VOC,\,in} = \frac{Q_{in}}{392\ ft^3}\left(y_{VOC,\,iu}\right) 60\frac{min}{hr} \tag{2.8}$$

where $M_{voc,in}$ is the molar flow rate of VOC in the inlet stream and Q is the volume flow rate in standard ft^3/min(scfm). The factor 392 is the volume (ft^3) occupied by one lb-mole of inlet gas stream at standard conditions (77°F and 1 atm). The number of lb-moles of VOC per hour in the outlet gas stream is calculated as follows:

$$M_{voc,\,out} = M_{voc,\,in}\,(1 - \eta) \tag{2.9}$$

where $M_{voc,out}$ is the molar flow rate of VOC in the exit stream. Finally, the number of lb-moles of VOC per hour that are condensed is calculated as follows:

$$M_{voc,\,con} = M_{voc,\,in} - M_{voc,\,out} \tag{2.10}$$

where $M_{voc,con}$ is the flow rate of the VOC that is condensed.

The condenser heat load is then calculated by the following equation:

$$H_{load} = \Delta H_{con} + \Delta H_{uncon} + \Delta H_{noncon} \tag{2.11}$$

where

H_{load} = condenser heat load (Btu/hr)
H_{con} = enthalpy change associated with the condensed VOC (Btu/hr)
H_{uncon} = enthalpy change associated with the uncondensed VOC (Btu/lir)
H_{noncon} = enthalpy change associated with the noncondensible air (Btu/hr).

The change in enthalpy of the condensed VOC is calculated as follows:

$$\Delta H_{con} = M_{VOC,\,con}\left[\Delta H_{VOC} + C_{p,\,VOC}\left(T_{in} - T_{con}\right)\right] \tag{2.12}$$

where H_{voc} is the molar heat of condensation and $C_{p,voc}$ is the molar heat capacity of the VOC. Each parameter varies as function of temperature. In Equation 8.12, H_{voc} and $C_{p,voc}$ are evaluated at the mean temperature:

$$T_{mean} = \frac{Tin + T_{con}}{2} \tag{2.12a}$$

The heat of condensation at a specific temperature, T_2, (°R), can be calculated from the heat of condensation at a reference temperature, T_1 (°R), using the Watson Equation:[7]

$$\left(\Delta H_{VOC}\ at\ T_2\right) = \left(\Delta H_{VOC}\ at\ T_1\right)\left(\frac{1 - \frac{T_2}{T_c}}{1 - \frac{T_1}{T_c}}\right)^{0.38} \tag{2.13}$$

where T_c (°R) is the VOC critical temperature.

The heat capacity can also be calculated for a specific temperature, T_2, if heat capacity constants (*a, b, c,* and *d*) are known for the particular compound. The heat capacity equation is:

$$C_{p,voc} = a + bT_2 + cT_2^2 + dT_2^3 \tag{2.14}$$

However, to simplify the heat load analysis, $C_{p,voc}$ can be assumed to remain constant over the temperature range of operation (i.e., T_{in} - T_{con}) without much loss of accuracy in the heat load calculations, as the sensible heat change in Equation 8.12 is relatively small compared to the enthalpy change due to condensation.

Heat of condensation and heat capacity data are provided in Appendix A. The heat of condensation for each compound is reported at its boiling point, while its heat capacity is given at 77°F. To estimate the heat of condensation at another temperature, use Equation 2.13. However, the Appendix A heat capacity data may be used to approximate $C_{p,voc}$ at other temperatures, since sensible heat changes are usually small, compared to condensation enthalpy changes.

The enthalpy change associated with the uncondensed VOC is calculated by the following expression:

$$H_{uncon} = M_{VOC,out} C_{p,VOC} (T_{in} - T_{con}) \quad (2.15)$$

Finally, the enthalpy change of the noncondensible air is calculated as follows:

$$\Delta H_{noncon} = \left[\left(\frac{Q_{in}}{392\ ft^3} 60 \frac{min}{hr} \right) - M_{VOC,in} \right] C_{p,air} (T_{in} - T_{con}) \quad (2.16)$$

where $C_{p,air}$ is the specific heat of air. In both Equations 2.15 and 2.16, the C_p's are evaluated at the mean temperature, as given by Equation 2.12a.

2.3.3 Condenser Size

Condensers are sized based on the heat load, the logarithmic mean temperature difference between the emission and coolant streams, and the overall heat transfer coefficient. The overall heat transfer coefficient, U, can be estimated from individual heat transfer coefficients of the gas stream and the coolant. The overall heat transfer coefficients for tubular heat exchangers where organic solvent vapors in noncondensible gas are condensed on the shell side and water/brine is circulated on the tube side typically range from 20 to 60 Btu/hr-ft^2-°F according to Perry's Chemical Engineers' Handbook[4]. To simplify the calculations, a single *"U"* value may be used to size these condensers. This approximation is acceptable for purposes of making study cost estimates.

Accordingly, an estimate of 20 Btu/hr-ft^2-°F can be used to obtain a conservative estimate of condenser size. The following equation is used to determine the required heat transfer area:

$$A_{con} = \frac{H_{load}}{U\ \Delta T_{lm}} \quad (2.17)$$

where

A_{con} = condenser surface area (ft^2)
U = overall heat transfer coefficient (Btu/hr-ft^2-°F)
T_{lm} = logarithmic mean temperature difference (°F).

The logarithmic mean temperature difference is calculated by the following equation, which is based on the use of a countercurrent flow condenser:

$$\Delta T_{lm} = \frac{\left(T_{in} - T_{cool,\,out}\right) - \left(T_{con} - T_{cool,\,in}\right)}{\ln\left(\frac{T_{in} - T_{cool,\,out}}{T_{con} - T_{cool,\,in}}\right)} \qquad (2.18)$$

where

$T_{cool,in}$ = coolant inlet temperature (°F)
$T_{cool,out}$ = coolant outlet temperature (°F).

The temperature difference ("approach") at the condenser exit can be assumed to be 15°F. In other words, the coolant inlet temperature, $T_{cool,in}$, will be 15°F less than the calculated condensation temperature, T_{con}. Also, the temperature rise of the coolant is specified as 25°F. (These two temperatures- the condenser approach and the coolant temperature rise-reflect good design practice that, if used, will result in an acceptable condenser size.) Therefore, the following equations can be applied to determine the coolant inlet and outlet temperature:

$$T_{cool,\,in} = T_{con} - 15\ ^{\circ}\mathrm{F} \qquad (2.19)$$

$$T_{cool,\,out} = T_{cool,\,in} + 25\ ^{\circ}\mathrm{F} \qquad (2.20)$$

2.3.4 Coolant Flow Rate

The heat removed from the emission stream is transferred to the coolant. By a simple energy balance, the flow rate of the coolant can be calculated as follows:

$$W_{cool} = \frac{H_{load}}{C_{p,\,cool}\left(T_{cool,\,out} - T_{cool,\,in}\right)} \qquad (2.21)$$

where W_{cool} is the coolant flow rate (lb/hr), and $C_{p,cool}$ is the coolant specific heat (Btu/lb-°F). $C_{p,cool}$ will vary according to the type of coolant used. For a 50-50 (volume %) mixture of ethylene glycol and water, $C_{p,cool}$ is approximately 0.65 Btu/lb-°F. The specific heat of brine (salt water), another commonly used coolant, is approximately 1.0 Btu/lb-°F.

2.3.5 Refrigeration Capacity

The refrigeration unit is assumed to supply the coolant at the required temperature to the condenser. The required refrigeration capacity is expressed in terms of refrigeration tons as follows:

$$R = \frac{H_{load}}{12{,}000} \tag{2.22}$$

Again, as explained in section 2.3.2, H_{load} does not include any heat losses.

2.3.6 Recovered VOC

The mass of VOC recovered in the condenser can be calculated using the following expression:

$$W_{voc,\,con} = M_{voc,\,con} \times MW_{voc} \tag{2.23}$$

where

$W_{voc,con}$ = mass of VOC recovered (or condensed) (lb/hr)
MW_{voc} = molecular weight of the VOC (lb/lb-mole).

2.3.7 Auxiliary Equipment

The auxiliary equipment for a refrigerated surface condenser system may include:

- precooler,
- recovered VOC storage tank,
- pumps/blowers, and
- piping/ductwork.

If water vapor is present in the treated gas stream, a precooler may be needed to remove moisture to prevent ice from forming in the VOC condenser. Sizing of a precooler is influenced by the moisture concentration and the temperature of the emission stream. As discussed in Section 2.2.3, a precooler may not be necessary for intermittently operated refrigerated surface condenser systems where the ice will have time to melt between successive operating periods.

If a precooler is required, a typical operating temperature is 35 to 40°F. At this temperature, almost all of the water vapor present will be condensed without danger of freezing. These condensation temperatures roughly correspond to a removal efficiency range of 70 to 80 percent

if the inlet stream is saturated with water vapor at 77°F. The design procedure outlined in the previous sections for a VOC condenser can be used to size a precooler, based on the psychometric chart for the air-water vapor system (see Reference [4]).

Storage/recovery tanks are used to store the condensed VOC when direct recycling is not a suitable option. The size of these tanks is determined from the amount of VOC condensate to be collected and the amount of time necessary before unloading. Sizing of pumps and blowers is based on the liquid and gas flow rates, respectively, as well as the system pressure drop between the inlet and outlet. Sizing of the piping and ductwork (length and diameter) primarily depends upon the stream flow rate, duct/pipe velocity, available space, and system layout.

2.3.8 Alternate Design Procedure

In some applications, it may be desirable to size and cost a refrigerated condenser system that will use a specific coolant and provide a particular condensation temperature. The design procedure to be implemented in such a case would essentially be the same as the one presented in this section except that instead of calculating the condenser exit temperature needed to obtain a specified VOC recovery efficiency, the exit temperature is given and the corresponding recovery efficiency is calculated.

The initial calculation would be to estimate the partial (=vapor) pressure of the VOC at the given condenser exit temperature, T_{con}, using Equation 2.6. Next, calculate η using Equation 2.24, by rearranging Equation 2.5:

$$\eta = \frac{\left(760\, y_{VOC,\,in}\right) - P_{VOC}}{y_{VOC,\,in}\left(760 - P_{VOC}\right)} \qquad (2.24)$$

Finally, substitute the calculated P_{voc} into this equation to obtain η. In the remainder of the calculations to estimate condenser heat load, refrigeration capacity, coolant flow rate, etc., follow the procedure presented in Sections 2.3.2 through 2.3.7.

2.4 Estimating Total Capital Investment

This section presents the procedures and data necessary for estimating capital costs for refrigerated surface condenser systems in solvent vapor recovery and gasoline vapor recovery applications. Costs for packaged and nonpackaged solvent vapor recovery systems are presented in Sections 2.4.1 and 2.4.2, respectively. Costs for packaged gasoline vapor recovery systems are

described in Section 2.4.3. Costs are calculated based on the design/sizing procedures discussed in Section 2.3.

Total capital investment, TCI, includes equipment cost, EC, for the entire refrigerated condenser unit, auxiliary equipment costs, taxes, freight charges, instrumentation, and direct, and indirect installation costs. All costs in this chapter are presented 3rd quarter 1990 dollars.[1]

For these control systems, the total capital investment is a battery limit cost estimate and does not include the provisions for bringing utilities, services, or roads to the site; the, backup facilities; the land; the working capital, the research and development required; or the process piping and instrumentation interconnections that may be required in the process generating tile waste gas. These costs are based on new plant installations; no retrofit cost considerations are included. The retrofit cost factors are so site specific that no attempt has been made to provide them.

The expected accuracy of the cost estimates presented in this chapter is ±30 percent (e.g.., "study" estimates). It must be kept in mind that even for a given application, design and manufacturing procedures vary from vendor to vendor, so costs may vary.

In the next two sections, equipment costs are presented for packaged and nonpackaged (custom) solvent vapor recovery systems. respectively. With the packaged systems the equipment cost is factored from the refrigeration the custom systems, the equipment cost is determined as the sum of the costs of the individual system components. Finally. equipment costs for packaged gasoline, vapor recovery systems are given in Section 2.4.3.

2.4.1 Equipment Costs for Packaged Solvent Vapor Recovery Systems

Vendors were asked to provide refrigerated unit cost estimates for a wide range of applications. The equations shown below for refrigeration unit equipment costs, EC_r, are multivariable regressions of data provided by two vendors and are only valid for the ranges listed in Table 2.2.[8,9] In this table, the capacity range of refrigeration units for which cost data were available are shown as a function of temperature.

Single Stage Refrigeration Units (less than 10 tons)

$$EC_r = \left(9.83 - 0.014T_{con} + 0.340 \ln R\right) \quad (2.25)$$

Table 1.2: Applicability Ranges for the Refrigeration Unit Cost Equations (Equations 8.25 to 8.27)

Temperature T_{con}(°F)[a]	Minimum Size Available *R*(tons)		Maximum Size Available *R*(tons)	
	Single Stage	Multistage	Single Stage	Multistage
40	0.85	NA[b]	174	NA
30	0.63	NA	170	NA
20	0.71	NA	880	NA
10	0.44	NA	200	NA
0 to -5	0.32	NA	133	NA
-10	0.21	3.50	6.6	81
-20 to -25	0.13	2.92	200	68
-30	NA	2.42	NA	85
-40	NA	1.92	NA	68
-45 to -50	NA	1.58	100[c]	55
-55 to -60	NA	1.25	100[c]	100
-70	NA	1.33	NA	42
-75 to -80	NA	1.08	NA	150
-90	NA	0.83	NA	28
-100	NA	0.67	NA	22

[a]For condensation temperatures that lie between the levels shown, round off to the nearest level (e.g., if T_{con} = 16°F, use 20°F) to determine minimum and maximum available size.
[b]NA = System not available based on vendor data collected in this study.
[c]Only one data point available.

Single Stage Refrigeration Units (greater than or equal to 10 tons)

$$EC_r = \left(9.26 - 0.007T_{con} + 0.627 \ln R\right) \quad (2.26)$$

Multistage Refrigeration Units

$$EC = \exp\left(9.73 - 0.012T_{con} + 0.584 \ln R\right) \quad (2.27)$$

Equations 2.25 and 2.26 provide costs for refrigeration units based on single stage designs, while Equation 2.27 gives costs for multistage units. Equation 2.27 covers both types of multistage units, "cascade" and "compound". Data provided by a vendor show that the costs of cascade and

compound units compare well, generally differing by less than 30%.[8] Thus, only one cost equation is provided. Equation 2.25 applies to single stage refrigeration units smaller than 10 tons and Equation 2.26 applies to single stage refrigeration units as large or larger than 10 tons. Single stage units typically achieve temperatures between 40 and -20°F, although there are units that are capable of achieving -60°F in a single stage.[8, 10] Multistage units are capable of lower temperature operation between -10 and -100°F.

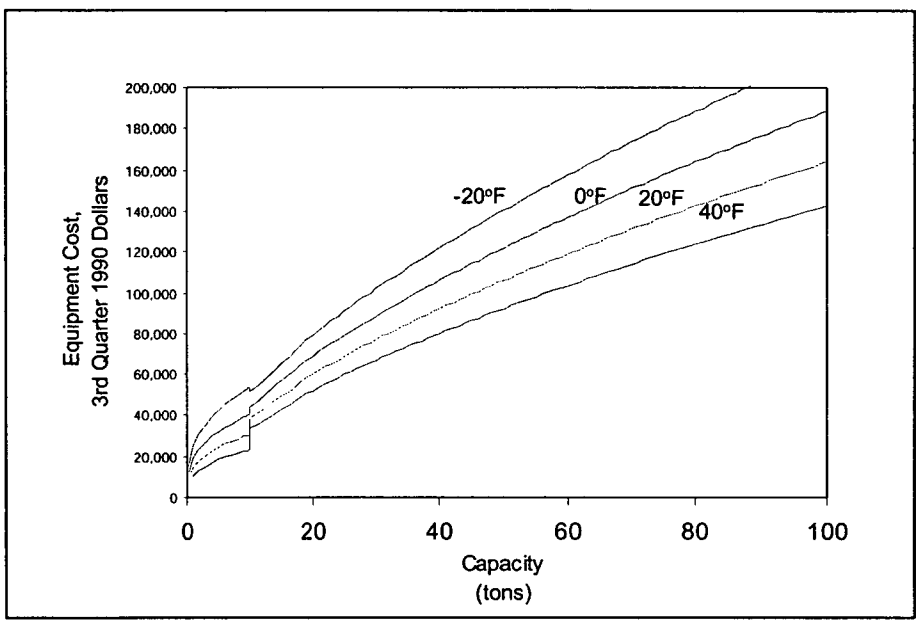


Figure 2.5: Refrigeration Unit Equipment Cost (Single Stage) [8.9]

Single stage refrigeration unit costs are depicted graphically for selected temperatures in Figure 2.5. Figure 2.6 shows the equipment cost curves for multistage refrigeration units.

(NOTE: In Figure 2.5, the discontinuities in the curves at the 10 ton capacity are a result of the two regression equations used. Equation 2.25 is used for capacities of less than 10 tons; Equation 2.26 is used for capacities greater than or equal to 10 tons.)

These costs are for outdoor models that are skid-mounted on steel beams and consist of the following components: walk-in weatherproof enclosure, air-cooled low temperature refrigeration machinery with dual pump design, storage reservoir, control panel and instrumentation, vapor condenser, and necessary piping. AR refrigeration units have two pumps: a system pump and a bypass pump to short-circuit the vapor condenser during no-load conditions. Costs for heat transfer fluids (brine) are not included.

The equipment cost of packaged solvent vapor recovery systems (EC_p) is estimated to be 25 percent greater than the cost of the refrigeration unit alone [9]. The additional cost includes VOC condenser, recovery tank, the necessary connections, piping, and additional instrumentation. Thus:

$$EC_p = 1.25\, EC_r \tag{2.28}$$

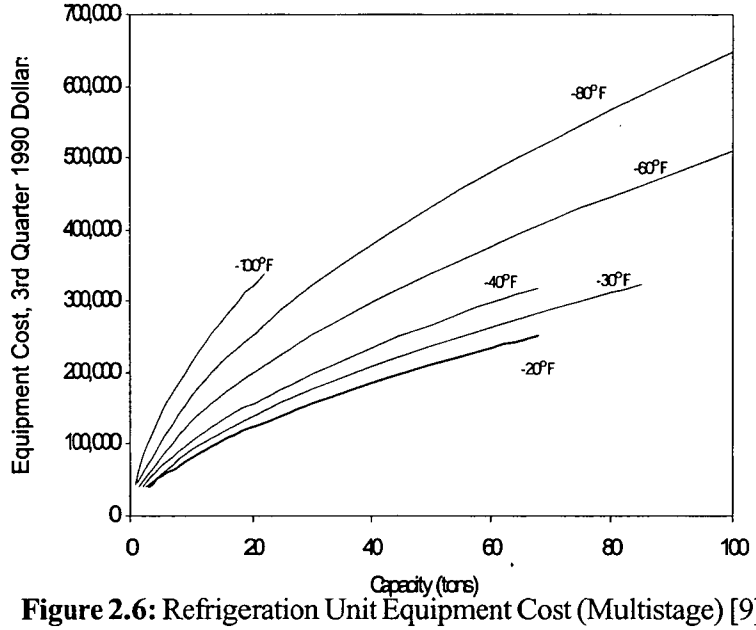


Figure 2.6: Refrigeration Unit Equipment Cost (Multistage) [9]

Purchased equipment cost, PEC, includes the packaged equipment cost, EC_p, and factors for sales taxes (0.03) and freight (0.05). Instrumentation and controls are included with the packaged units. Thus,

$$PEC_p = EC_p\,(1 + 0.03 + 0.05) = 1.08\, EC_p \tag{2.29}$$

2.4.2 Equipment Costs for Nonpackaged (Custom) Solvent Vapor Recovery Systems

To develop cost estimates for nonpackaged or custom refrigerated systems, information was solicited from vendors on costs of refrigeration units, VOC condensers, and VOC storage/ recovery tanks [9, 11, 12]. Quotes from the vendors were used to develop the estimated costs for each type of equipment. Only one set of vendor data was available for each type of equipment.

Equations 2.25, 2.26, and 2.27 shown above are applicable for estimating the costs for the

refrigeration units. Equation 8.30 shows the equation developed for the VOC condenser cost estimates[11]:

$$EC_{con} = 34\,A_{con} + 3{,}755 \qquad (2.30)$$

This equation is valid for the range of 38 to 800 ft^2 and represents costs for shell arid tube type heat exchangers with 304 Stainless Steel tubes.

The following equation represents the storage/recovery tank cost data obtained from one vendor[12]:

$$EC_{tank} = 2.72\,V_{tank} + 1{,}960 \qquad (2.31)$$

These costs are applicable for the range of 50 to 5,000 gallons and pertain to 316 stainless steel vertical tanks.

Costing procedures for a precooler (EC_{pre}) that includes a separate condenser/refrigeration unit and a recovery tank are similar to that for a custom refrigerated condenser system. Hence, Equations 2.25 through 2.31 would be applicable, with the exception of Equation 2.27, which represents multistage systems. Multistage systems operate at much lower temperatures than that required by a precooler.

Costs for auxiliary equipment such as ductwork, piping, fans, or pumps are designated as EC_{aux}, These items should be costed separately using methods described elsewhere in this Manual.

The total equipment cost for custom systems, EC_c is then expressed as:

$$EC_c = EC_r + EC_{con} + EC_{tank} + EC_{pre} + EC_{aux} \qquad (2.32)$$

The purchased equipment cost including EC_c and factors for sales taxes (0.03), freight (0.05), and instrumentation and controls (0.10) is given below:

$$PEC_c = EC_c\,(1 + 0.03 + 0.05 + 0.10) = 1.18\,EC_c \qquad (2.33)$$

2.4.3 Equipment Costs for Gasoline Vapor Recovery Systems

Separate quotes were obtained for packaged gasoline vapor recovery systems because these systems are specially designed for controlling gasoline vapor emissions from such sources as storage tanks, gasoline bulk terminals, and marine vessel loading and unloading operations. Systems that control marine vessel gasoline loading and unloading operations also must meet U.S. Coast

Guard safety requirements.

Quotes obtained from one vendor were used to develop equipment cost estimates for these packaged systems (see Figure 2.7). The cost equation shown below is a least squares regression of these cost data and is valid for tile range 20 to 140 tons.[91]

$$EC_p = 4{,}910\,R + 212{,}000 \qquad (2.34)$$

The vendor data in process flow capacity (gal/min) vs cost ($) were transformed into Equation 2.34 after applying the design procedures in Section 2.3. Details of the data transformation are given in Appendix B.

The cost estimates apply to skid-mounted refrigerated VOC condenser systems for hydrocarbon vapor recovery primarily at gasoline loading/storage facilities. The systems are intermittently operated at -80 to -120°F allowing 30 to 60 minutes per day for defrosting by circulation of warm brine. Multistage systems are employed to achieve these lower temperatures. Tile achievable VOC removal efficiencies for these systems are in the range of 90 to 95 percent.

The packaged systems include the refrigeration unit with the necessary pumps, compressors, condensers/evaporators, coolant reservoirs, the VOC condenser unit and VOC recovery tank, precooler, instrumentation and controls, and piping. Costs for heat transfer fluids (brines) are not included. The purchased equipment cost for these systems includes sales tax and freight and is calculated using Equation 2.29.

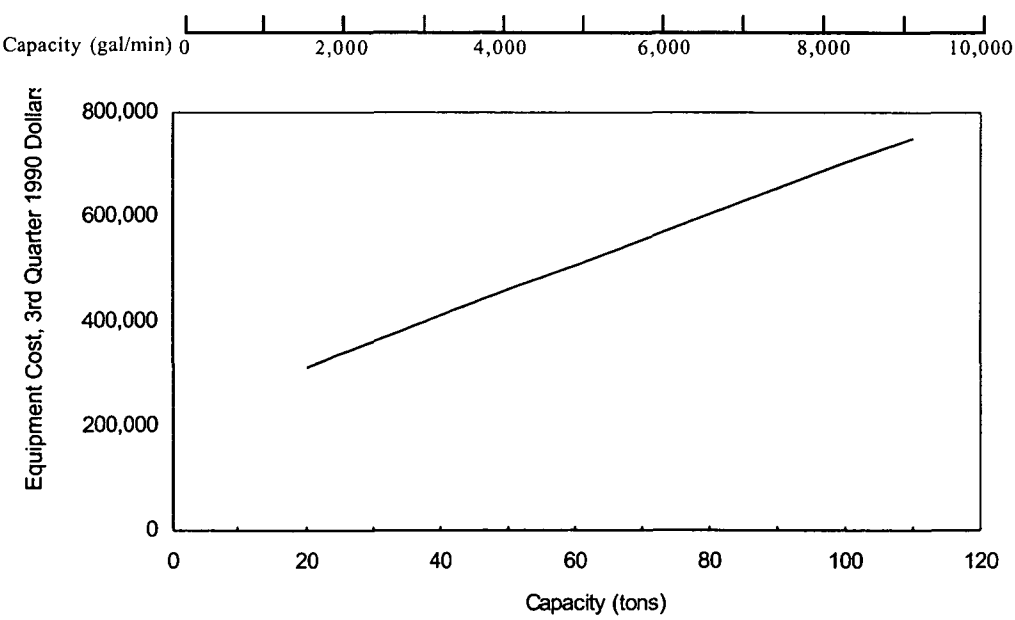


Figure 2.7: Gasoline Vapor Recovery System Equipment Cost [9]

Table 2.3: Capital Cost Factors for Nonpackaged (Custom) Refrigerated Condenser Systems

Cost Item	Factor
Purchased Equipment Costs	
Refrigerated condenser system, EC	As estimated, A
Instrumentation	0.10 A
Sales Taxes	0.03 A
Freight	0.05 A
Purchased equipment costs, PEC	B=1.18 A[a]
Direct Installation Costs	0.08 B
Foundations & Supports	0.14 B
Handling & Erection	0.08 B
Electrical	0.08 B
Piping	0.02 B
Insulation	0.10 B
Painting	0.01 B
Direct Installation Costs	0.43 B
Site Preparation	As Required, SP
Buildings	As Required, Bldg.
Total Direct Costs, DC	1.43 B + SP + Bldg.
Indirect Costs (Installation)	
Engineering	0.10 B
Construction and Field Expenses	0.05 B
Contractor Fees	0.10 B
Start-Up	0.02 B
Performance Test	0.01 B
Contingencies	0.03 B
Total Indirect Costs, IC	0.31 B
Total Capital Investment = DC + IC	**1.74 b + SP + Bldg.[b]**

[a] Purchased equipment cost factor for packaged systems is 1.08 with instrumentation included.
[b] For packaged systems, total capital investment = $1.15PEC_p$.

Table 2.4: Suggested Annual Cost Factors for Refrigerated Condenser Systems

Cost Item	Factor
Direct Annual Cost, DC	
Operating Labor	1/2 hour per shift
Operator	15% of operator
Supervisor	
Operating Materials	
Maintenance	
Labor	1/2 hour per shift
Material	100% of maintenance labor
Electricity	
at 40°F	1.3 kW/ton
at 20°F	2.2 kW/ton
at -20°F	4.7 kW/ton
at -50°F	5.0 kW/ton
at -100°F	11.7 kW/ton
Indirect Annual Costs, IC	
Overhead	60% of total labor and maintenance material costs
Administrative Charges	2% of Total Capital Investment
Property Tax	1% of Total Capital Investment
Insurance	1% of Total Capital Investment
Capital Recovery[a]	0.1098 x Total Capital Investment
Recovery Credits, RC	
Recoverd VOC	Quality recovered x operating hours
Total Annual Cost	**DC + IC - RC**

[a] Assuming a 15 year life at 7% [13]. See Chapter 2.

2.4.4 Installation Costs

The total capital investment, TCl, for packaged systems is obtained by multiplying the purchased equipment cost, PEC_p by the total installation factor:[13]

Table 2.5: Electricity Requirements

Electricity (E, kW/ton)	Temperature (°F)
1.3	40
2.2	20
4.7	-20
5.0	-50
11.7	-100

$$TCI = 1.15\ PEC_p \tag{2.35}$$

For nonpackaged (custom) systems, the total installation factor is 1.74:

$$TCI = 1.74\ PEC_c \tag{2.36}$$

An itemization of tile total installation factor for nonpackaged systems is shown in Table 2.3. Depending on the site conditions., the installation costs for a given system could deviate significantly, from costs generated by there average factors. Guidelines are available for adjusting these average installation factors.[14]

2.5 Estimating Total Annual Cost

The total annual cost (TAC) is the sum of the direct and indirect annual costs. The bases used in calculating annual cost factors are given in Table 2.4.

2.5.1 Direct Annual Costs

Direct annual costs, DC, include-labor (operating and supervisory), maintenance (labor and materials), and electricity. Operating labor is estimated at 1/2-hour per 8-hour shift. The supervisory labor cost is estimated at 15% of the operating labor cost. Maintenance labor is

estimated at 1/2-hour per 8-hour shift.

Maintenance materials costs are assumed to equal maintenance labor costs.

Utility costs for refrigerated condenser systems include electricity requirements for the refrigeration unit and any pumps/blowers. The power required by the pumps/blowers is negligible when compared with the refrigeration unit power requirements. Electricity requirements for refrigerated condenser systems are summarized in Table 2.5:
These estimates were developed from product literature obtained from one vendor.[9] The electricity cost, C_e, can then be calculated from the following expression:

$$C_e = \frac{R}{\eta_{compressor}} E \ \theta_s \ p_e \qquad (2.37)$$

where

θ_s = system operating hours (hr/yr)
p_e = electricity cost
$\eta_{compressor}$ = mechanical efficiency of the compressor

2.5.2 Indirect Annual Costs

Indirect annual costs, IC, are calculated as the sum of capital recovery costs plus general and administrative (G&A), overhead, property tax, and insurance costs. Overhead is assumed to be equal to 60 percent of the sum of operating, supervisory, and maintenance labor, and maintenance materials. Overhead cost is discussed in Section 1 of this Manual.

The system capital recovery cost, CRC, is based on an estimated 15-year equipment life.[13] (See Section 1 of the Manual for a discussion of the capital recovery cost.) For a 15-year life and an interest rate of 7 percent, the capital recovery factor is 0.1098. The system capital recovery cost is then estimated by:

$$C_e = \frac{R}{\eta_{compressor}} E \ \theta_s \ p_e \qquad (2.37)$$

G&A costs, property tax, and insurance are factored from total capital investment. typically at 2 percent, 1 percent, and 1 percent, respectively.

Table 2.6: Example Problem Data

Vent Stream Parameters	**Value**
Inlet Stream Flow Rate	100 scfm[a]
Inlet Stream Temperature	86°F
VOC to be Condensed	Acetone
VOC Inlet Volume Fraction	0.0375
Required VOC Removal Efficiency	.90
Antoine Equation Constants for Acetone:	
A	7.117
B	1210.595
C	229.664
Heat of Condensation of Acetone[b]	12,510 Btu/lb-mole
Heat Capacity of Acetone[c]	17.90 Btu/lb-mole-°F
Specific Heat of Coolant[c] (ethylene glycol)	0.65 Btu/lb-°F
Heat Capacity of Air[c]	6.95 Btu/lb-mole-°F
Annual Cost Parameter	**Value**
Operating Labor	$15.64/hr
Maintenance Labor	$17,2/hr
Electricity	$0.0461/kWh
Acetone Resale Value	$0.10/lb

[a]Standard conditions: 77°F and 1 atmosphere.
[b]Evaluated at the acetone boiling point (134°F).
[c]These properties were evaluated at 77°F.

2.5.3 Recovery Credit

If the condensed VOC can be directly reused or sold without further treatment, then the credit from this operation is incorporated in the total annual cost estimates. The following equation can be used to estimate the VOC, recovery credit, *RC*:

$$CRC = 0.1098\ TCI \tag{2.38}$$

where

p_{voc} = resale value of recovered VOC ($/lb)
$W_{voc,con}$ = quantity of VOC recovered (lb/hr).

2.5.4 Total Annual Cost

The total annual cost (TAC) is calculated as the sum of the direct and indirect annual costs, minus the recovery credit:

$$RC = W_{vox,\,con}\ p_{voc} \tag{2.39}$$

2.6 Example Problem 1

The example problem described in this section shows how to apply the refrigerated condenser system sizing and costing procedures to the control of a vent stream consisting of acetone, air, and a negligible amount of moisture. This example problem assumes a required removal efficiency and calculates the temperature needed to achieve this level of control.

2.6.1 Required Information for Design

The first step in the design procedure is to specific the gas stream to be processed. Gas stream parameters to be used in this example problem are listed in Table 2.6. The values for the Antoine equation constants, heat, of condensation, and heat capacity of acetone are obtained from Appendix 8A. Specific heat of the coolant is obtained from Perry's *Chemical Engineers' Handbook*[4].

2.6.2 Equipment Sizing

The first step in refrigerated condenser sizing is determining the partial pressure of the VOC at the outlet of the condenser for a given removal efficiency. Given the stream flow rate, inlet VOC concentration, and the required removal efficiency, the partial pressure of the VOC at the outlet can be calculated using Equation 2.5.

$$P_{VOC} = 760\ \frac{0.375\ (1 - 0.90)}{1 - 0.375\ (0.90)} = 43\ \text{mm Hg}$$

Next, the temperature necessary to condense the required amount of VOC must be determined using Equation 8.7:

$$T_{con} = \left(\frac{1210.595}{7.117 - \log_{10}(43)} - 229.664 \right) 1.8 + 32 = 16^\circ \text{ F}$$

The next step is to estimate the VOC condenser heat load. Calculate: (1) the VOC flow rate for the inlet/outlet emission streams, (2) the flow rate of the condensed VOC, and (3) the condenser heat balance. The flow rate of VOC in the inlet stream is calculated from Equation 2.8.

$$M_{voc,\,in} = \frac{100}{393} (0.375)\, 60 = 5.74 \ \frac{\text{lb-moles}}{\text{hr}}$$

The flow rate of VOC in the outlet stream is calculated using Equation 2.7 as follows:

$$M_{voc,\,out} = 5.74\ (1 - 0.90) = 0.574 \ \frac{\text{lb-moles}}{\text{hr}}$$

Finally, the flow rate of condensed VOC is calculated with Equation 2.10:

$$M_{voc,\,con} = 5.74 - 0.574 = 5.166 \ \frac{\text{lb-moles}}{\text{hr}}$$

Next, the condenser heat balance is conducted. As indicated in Table 2.6, the acetone heat of condensation is evaluated at its boiling point, 134°F. However, it is assumed (for simplicity) that all of the acetone condenses at the condensation temperature, T_{con} = 16°F. To estimate the heat of condensation at 16°F, use the Watson equation (Equation 8.13) with the following inputs:

T_c = 918°R(Appendix A)
T_1 = 134 + 460 = 594°R
T_2 = 16 + 460 = 476°R.

Upon substitution, we obtain:

$$\left(\Delta H_{VOC}\ at\ 16^{\circ}\ F\right) = 12{,}510\left(\frac{1-\frac{476}{918}}{1-\frac{594}{918}}\right)^{0.38}$$

$$= 14{,}080\ \frac{Btu}{lb-mole}$$

As Table2.6 shows, the heat capacities of acetone and air and the specific heat of the, coolant were all evaluated at, 77°F. This temperature is fairly close to the condenser mean operating temperature, i.e., (86 + 16)/2 = 51°F. Consequently, using tile 77°F values would not add significant additional error to the heat load calculation.

The change in enthalpy of the condensed VOC is calculated using Equation 2.12:

$$\Delta H_{con} = 5.166\left[14{,}080 + 17.90\ (86 - 16)\right] = 79{,}210\ \frac{Btu}{hr}$$

The enthalpy change associated with the uncondensed VOC is calculated from Equation 2.15:

$$\Delta H_{uncon} = (0.574)\ (17.90)\ (86 - 16) = 719\ \frac{Btu}{hr}$$

Finally, the enthalpy change of the noncondensible air is estimated from Equation 2.16:

$$\Delta H_{noncon} = \left[\left(\frac{100}{392}\ 60\right) - 5.74\right] 6.95\ (86 - 16) = 4{,}654\ \frac{Btu}{hr}$$

The condenser heat load is then calculated by substituting H_{con}, H_{uncon}, and H_{noncon} in Equation 2.11:

$$H_{load} = 79{,}210 + 719 + 4{,}654 = 84{,}583\ \frac{\text{Btu}}{\text{hr}}$$

The next step is estimation of the VOC condenser size. The logarithmic mean temperature difference is calculated using Equation 2.18. In this calculation:

$T_{cool,in}$ = 16 - 15 = 1°F
$T_{cool,out}$ = 1 + 25 = 26°F

from Equations 2.19 and 2.20, respectively:

$$\Delta T_{lm} = \frac{(86 - 26) - (16 - 1)}{\left(\ln \frac{86 - 26}{16 - 1}\right)} = 32.5° \text{ F}$$

The condenser surface area can then be calculated using Equation 2.17.

$$A_{con} = \frac{84{,}583}{20\ (32.5)} = 130\ ft^2$$

In this equation, a conservative value of 20 Btu//hr-ft^2-°F is used as the overall heat transfer coefficient.

The coolant flow rate can be calculated using Equation 2.21.

$$W_{cool} = \frac{84{,}583}{0.65\ (26 - 1)} = 5{,}205\ \frac{lb}{hr}$$

The refrigeration capacity can be estimated from Equation 2.22 as follows:

$$R = \frac{84{,}583}{12{,}000} = 7.05\ tons$$

Finally, the quantity of recovered VOC can be estimated using Equation 2.23:

$$W_{voc,con} = 5.166 \times 58.08 = 300 \text{ lb/hr}$$

where the molecular weight of acetone is obtained from Appendix A.

Note that in this example case, the partial pressure of acetone at the condenser exit is relatively high (43 mm Hg). In applications where much lower outlet concentrations are desired, a second control device (e.g., incinerator, adsorber) to operate in series with the condenser may need to be considered.

2.6.3 Equipment Costs

Once the system sizing parameters have been determined, the equipment costs can be calculated. For the purpose of this example, a custom refrigerated condenser system, including a refrigeration unit, a VOC condenser, and a recovery tank will be costed.

From Table 2.2, a single stage refrigeration unit appears to be suitable for the example problem with an estimated condensation temperature of 16°F and capacity of 7.05 tons. Hence Equation 2.25, which is applicable to units less than 10 tons. is selected for estimating costs. Application of this equation results in the following value for the refrigeration unit cost:

$$EC = \exp\left[9.83 - 0.014\ (16) + 0.340\ \ln(7.05)\right] = \$28{,}855$$

VOC condenser cost is computed using Equation 2.30 as follows:

$$EC_{con} = 34\ (130) + 3{,}775 = \$8{,}195$$

Recovery tank cost can be calculated from Equation 2.31. For this case, $W_{voc,con} = 300$ lb/hr, which is equivalent to 45.5 gal/hr (density of acetone is about 6.6 lb/gal). Assuming an 8-hour daily operation, the interim storage capacity requirement would be 364 gallons. Application of Equation 2.31 leads to the following:

$$EC_{tank} = 2.72\ (364) + 1{,}960 = \$2{,}950$$

Assuming there are no additional costs due to precooler or other auxiliary equipment, the total equipment cost is calculated from Equation 2.32:

$$EC_c = 28{,}855 + 8{,}195 + 2{,}950 + 0 + 0 = \$40{,}000$$

The purchased equipment cost including instrumentation, controls, taxes, and freight is estimated using Equation 2.33:

$$PEC_c = 1.18\ (40{,}000) = \$47{,}200$$

The total capital investment is calculated using Equation 2.36:

$$TCI = 1.74\ (47{,}000) = \$82{,}128$$

2.6.4 Total Annual Cost

Table 2.7 summarizes the estimated annual costs for the, example problem. The cost calculations are shown in the table. Direct annual costs for refrigerated systems include labor, materials, and utilities. Labor costs are based on 8-hr/day, 5-day/week operation. Supervisory labor is computed at 15 percent of operating labor, and operating and maintenance labor are each based on 1/2 hr per 8-hr shift. The electricity cost is based on a requirement of 2.2 kW/ton, because the condensation temperature (16°F) is close to the 20°F temperature given for this value. Indirect annual costs include overhead, capital recovery, administrative charges, property tax, and insurance.

Total annual cost is estimated using Equation 2.40. For this example case, application of refrigerated condensation as a control measure results in an annual savings of $37,500. As Table 2.7 shows, the acetone recovery credit is over twice the direct and Indirect costs combined. Clearly, this credit has more influence on the total annual cost than any other Component. Although the credit depends on three parameters-the acetone recovery rate, the annual operating hours, and the acetone salvage value ($0.10/lb)-the last parameter is often the most difficult to estimate. This is mainly because the salvage value varies according to the facility location as well as the current state of the chemical market.

Table 1.7: Annual Cost for Refrigerated Condenser System Example Problem

Cost Item	Calculations	Cost
Direct Annual Costs, DC		
Operating Labor		
Operator	$\frac{0.5\text{h}}{\text{shift}} \times \frac{\text{shift}}{8\text{h}} \times \frac{2{,}080\text{h}}{\text{yr}} \times \frac{\$15.64}{\text{h}}$	$2,030
Supervisor	15% of operator = 0.15 x 2,030	300
Operating materials		-
Maintenance		
Labor	$\frac{0.5\text{h}}{\text{shift}} \times \frac{\text{shift}}{8\text{h}} \times \frac{2{,}080\text{h}}{\text{yr}} \times \frac{\$17.21}{\text{h}}$	2,240
Material	100% maintenance labor	2,240
Utilities		
Electricity	$\frac{7.05\text{tons}}{0.85\ \text{Ton}} \times \frac{2.2\text{kw}}{\text{yr}} \times \frac{2{,}080\text{h}}{\text{kwh}} \times \$0.0461$	1,750
Total DC		$8,560
Indirect Annual Costs, JC		
Overhead	60% of total labor and maintenance material = 0.6 (2,030 + 305 + 2,240 + 2,240)	4,090
Administrative charges	2% of Total Capital Investment = 0.02($82,100)	1,640
Property tax	1% of Total Capital Investment = 0.01($82,100)	820
Insurance	1% of Total Capital Investment = 0.01($82,100)	820
Capital recovery[a]	0.1098 x $82,100	9,010
Total IC		$16,380
Recovery Credits, RC		
Recovered Acetone	$\frac{300\text{lb}}{\text{h}} \times \frac{2{,}080\text{h}}{\text{yr}} \times \frac{\$0.10}{\text{lb}}$	($62,400)
Total Annual Cost (rounded)		($37,500)
		(Savings)

[a] The capital recovery cost factor, CRF, is function of the refrigerated condenser equipment life and the opportunity cost of the capital (i.e., interest rate). For example, for a 15 year equipment life and a 7% interest rate, CRF = 0.1098.

2.7 Example Problem 2

In this example problem, the alternate design procedure described in Section 2.3.8 is illustrated. The temperature of condensation is given, and the resultant removal efficiency is calculated. The example stream inlet parameters are identical to Example Problem 1 with the exception that removal efficiency is not specified and the required temperature of condensation is assumed to be 16°F.

2.7.1 Required Information for Design

The first step is to calculate the partial pressure of the VOC at the specified temperature (16°F) using Equation 2.6 to solve for P_{VOC}:

$$H_{load} = 143\ Q_g$$

Remember to convert T_{con} to degrees Centigrade, i.e., 16°F = -8.9°C.

Substituting the values for the Antoine equation constants for acetone as listed in Table 2.6:

$$R = \frac{H_{load}}{12{,}000} = 0.0119\ Q_g$$

$$P_{VOC} = 43 \text{ mm Hg.}$$

Using Equation 2.24, the removal efficiency is:

$$Q_g = 83.9\ R$$

The remainder of the calculations in this problem are identical to those in Example Problem 1.

2.8 Acknowledgments

The authors gratefully acknowledge the following companies for contributing data to this chapter:

- Edwards Engineering (Pompton Plains, NJ)
- Piedmont Engineering (Charlotte, NC)

- Universal Industrial Refrigeration (Gonzales, LA)
- ITT Standard (Atlanta, GA)
- XChanger (Hopkins, MN)
- Buffalo Tank Co. (Jacksonville, FL)

References

[1] Erikson, D.G., *Organic Chemical Manufacturing Volume 5: Adsorption, Condensat ion, and Absorption Devices*, U.S. Environmental Protection Agency. Research Triangle Park, North Carolina, Publication No. EPA 450/3-80-027, December 1980.

[2] Vatavuk, W.M., and R.B. Neveril, "Estimating Costs of Air Pollution Control Sys tems: Part XV1. Costs of Refrigeration Systems", *Chemical Engineering*, May 16, 1983, pp. 95-98.

[3] McCabe, W.L., and J.C. Smith, *Unit Operations of Chemical Engineering* (Third Edition), McGraw-Hill Book Company, New York, 1976.

[4] Perry, R.H. and C.H. Chilton, Eds. *Chemical Engineers' Handbook* (Sixth Edition), McGraw-Hill Book Company, New York, 1989.

[5] Kern, D.Q., *Process Heat Transfer*, McGraw-Hill Book Company, New York, 1950

[6] Smith, J.M., and M.C. VanNess, *Introduction to Chemical Engineering Thermody namics* (Third Edition), McGraw-Hill Book Company, New York, 1975.

[7] Reid, Robert C., John M. Prausnitz, and Bruce E. Poling, *Properties of Gases & Liquids* (Fourth Edition), McGraw-Hill Book Company, New York, 1987.

[8] Letter and attachment from Robert V. Sisk. Jr. of Piedmont Engineering, Pineville, North Carolina, to Wiley Barbour of Radian Corporation, Research Triangle Park, North Carolina, January 28, 1991.

[9] Letter and attachment from Waldrop, R., and V. Sardo of Edwards Engineering

Corp., Pompton Plains, New Jersey, to Wiley Barbour of Radian Corporation, Research Triangle Park, North Carolina, October 1, 1990.

[10] Price, Brian C., "Know the Range and Limitations of Screw Compressors," *Chemical Engineering Progress*, 87(2):50-56.

[11] Letter and attachment from Bob Hansek of ITT Corporation, Atlanta, Georgia to Wiley Barbour of Radian Corporation, Research Triangle Park, North Carolina, October 10, 1990.

[12] Letter and attachment from Avery Cooke of Liquid Handling Equipment, Inc., Charlotte, North Carolina to Rich Pelt of Radian Corporation, Research Triangle Park, North Carolina, September 20, 1990.

[13] Letter from Richard Waldrop of Edwards Engineering Corp., Pompton Plains, New Jersey to William Vatavuk, P.E., Durham, North Carolina, August 29, 1988.

Appendix A

Properties of Selected Compounds

Table 2.8: Properties of Selected Compounds

Compound	Critical Temp.[a] (°R)	Boiling Point (°F)	Molecular Weight (lb/lb-mole)	Heat of Condensation[b] (Btu/lb-mole)	Heat Capacity[c] (Btu/lbmole °F)	State
Acetone	918	134	58.08	12,510	30.22	Liquid
					17.90	Gas
Acetylene	555	-119	26.02	7,290	10.50	Gas
Acrylonitrile	-	171	53.06	14,040	15.24	Gas
Aniline	1259	364	93.13	19,160	45.90	Liquid
					25.91	Gas
Benzene	1012	176	78.11	13,230	19.52	Liquid
Benzonitrile	1259	376	103.12	19,800	26.07	Gas
Butane	766	31	58.12	9,630	23.29	Gas
Chloroethane	829	54	64.52	10,610	14.97	Gas
Chloroform	966	143	119.39	12,740	15.63	Gas
Chloromethane	750	-12	50.49	9,260	9.74	Gas
Cyclobutane	-	55	56.10	10,410	17.26	Gas
Cyclohexane	997	177	84.16	12,890	37.4	Liquid
					25.40	Gas
Cyclopentane	921	121	70.13	11,740	30.80	Liquid
					19.84	Gas
Cyclopropane	716	-27	42.08	8,630	13.37	Gas
Diethyl ether	840	94	74.12	11,480	40.8	Liquid
					26.89	Gas
Dimethylamine	788	44	45.09	11,390	16.50	Gas
Ethylbenzene	1111	277	106.17	15,300	30.69	Gas
Ethylene oxide	845	51	44.05	10,980	11.54	Gas
Heptane	973	209	100.12	13,640	53.76	Liquid
					39.67	Gas
Hexane	914	156	86.18	12,410	45.2	Liquid
					34.20	Gas
Methanol	923	148	32.04	14,830	19.40	Liquid
					10.49	Gas
Octane	1024	258	114.23	14,810	45.14	Gas
Pentane	846	97	72.15	11,090	28.73	Gas
Toluene	1065	231	92.14	14,270	37.58	Liquid
					24.77	Gas
o - Xylene	1135	292	106.17	15,840	44.9	Liquid
					31.85	Gas
m - Xylene	1111	282	106.17	15,640	43.8	Liquid
					30.49	Gas
p - Xylene	1109	281	106.17	15.480	30.32	Gas

[a] Reprinted with permission from *Lange's Handbook of Chemistry* (12th edition), Table 9-7.[15]
[b] Reprinted with permission from *Lange's Handbook of Chemistry* (12th edition), Table 9-4.[15] (Measured at boiling point.)
[c] Reprinted with permission from *Lange's Handbook of Chemistry* (12th edition), Table 9-2.[15] (Measured at 77°F.)

Table 2.9: Antoine Equation Constants for Selected Compounds[a]

Compound	Antoine Equation Constants			Valid Temperature Range (°F)
	A	B	C	
Acetone	7.117	1210.59555	229.66	Liquid
Acetylene	7.100	711.0	253.4	-116 to -98
Acrylonitrile	7.039	1232.53	222.47	-4 to 248
Aniline	7.320	1731.515	206.049	216 to 365
Benzene	6.905	1211.033	220.790	46 to 217
Benzonitrile	6.746	1436.72	181.0	Liquid
Butane	6.809	935.86	238.73	-107 to 66
Chloroethane	6.986	1030.01	238.61	-69 to 54
Chloroethylene	6.891	905.01	239.48	-85 to 9
Chloroform	6.493	929.44	196.03	-31 to 142
Chloromethane	7.0933	948.58	249.34	-103 to 23
Cyanic acid	7.569	1251.86	243.79	-105 to 21
Cyclobutane	6.916	1054.54	241.37	-76 to 54
Cyclohexane	6.841	1201.53	222.65	68 to 178
Cyclopentane	6.887	1124.16	231.36	-40 to 162
Cyclopropane	6.888	856.01	246.50	-130 to -26
Diethyl ether	6.920	1064.07	228.80	-78 to 68
Diethylamine	5.801	583.30	144.1	88 to 142
Dimethylamine	7.082	960.242	221.67	-98 to 44
Dioxane - 1,4	7.432	1554.68	240.34	68 to 221
Ethyl benzene	6.975	1424.255	213.21	79 to 327
Ethylene oxide	7.128	1054.54	2371.76	-56 to 54
Heptane	6.897	1264.90	216.54	28 to 255
Hexane	6.876	1171.17	224.41	-13 to 198
Methanol	7.897	1474.08	229.13	7 to 149
Octane	6.919	1351.99	209.15	66 to 306
Pentane	6.853	1064.84	233.01	-58 to 136
Toluene	6.955	1344.8	219.48	43 to 279
Vinyl acetate	7.210	1296.13	226.66	72 to 162
o - Xylene	6.999	1474.679	213.69	90 to 342
m - Xylene	7.009	1462.266	215.11	82 to 331
p - Xylene	6.991	1453.430	215.31	81 to 331

[a] Reprinted with permission from *Lange's Handbook of Chemistry* (12th edition), Table 10-8.[15]

Appendix B

Documentation for Gasoline Vapor Recovery System Cost Data

As mentioned in Section 2.4.3, vendor cost data were obtained that related the equipment cost ($) of packaged gasoline vapor recovery systems to the process flow capacity (gal/min). These data needed to be transformed, in order to develop Equation 2.34, which relates equipment cost ($) to system refrigeration capacity (R, tons), as follows:

$$EC_p = 4{,}910R + 212{,}000$$

To make this transformation, we needed to develop an expression relating flow capacity to refrigeration capacity. The first step was to determine the inlet partial pressure ($P_{VOC,in}$) of the VOC-gasoline, in this case. As was done in Section 3.1, we assumed that the VOC vapor was saturated and, thus, in equilibrium with the VOC liquid. This, in turn, meant that we could equate the partial pressure to the vapor pressure. The "model" gasoline had a Reid vapor pressure of 10 and a molecular weight of 66 lb/lb-mole, as shown in Section 4.3 of *Compilation of Air Pollutant Emission Factors* (FPA publication AP-42, Fourth Edition, September 1985). For this gasoline, the following Antoine equation constants were used:

$$A = 12.5733$$
$$B = 6386.1$$
$$C = 613$$

These constants were obtained by extrapolating available vapor pressure vs. temperature data for gasoline found in Section 4.3 of AP-42. Upon substi-tuting these constants and an assumed inlet temperature of 77 °F (25 °C) into the Antoine equation and solving for the inlet partial pressure ($P_{voc,in}$) We obtain:

$$\log P_{VOC,in} = A - \frac{B}{T_{in} + C}$$

$$= 12.5733 - \frac{6386.1}{25 + 613}$$

$$P_{VOC,in} = 366\ mm\ Hg$$

If the system operates at atmospheric pressure (760 mm Hg), this partial pressure would correspond to a VOC volume fraction in the inlet stream of:

$$y_{VOC,in} = \frac{366\ mm}{760\ mm} = 0.482$$

The outlet partial pressure ($P_{voc,out}$) and volume fraction are calculated in a similar way. The condensation (outlet) temperature used in these calcu-lations is -80 °F (-62 °C), the typical operating Temperature for the gasoline vapor recovery units for which the vendor supplied costs.

$$\log P_{VOC,out} = 12.5733 - \frac{6386.1}{-62 + 613}$$

$$P_{VOC,out} = 9.62\ mm\ Hg$$

This corresponds to a volume fraction in the outlet stream ($y_{voc,out}$) of:

$$y_{VOC,out} = \frac{9.62\ mm}{760\ mm} = 0.0127$$

Substitution of $P_{VOC,out}$ and $y_{VOC,in}$ into Equation 2.24 yields the condenser removal efficiency (η):

$$\eta = \frac{(760 \times 0.482) - 9.62}{0.482\,(760 - 9.62)} = 0.986$$

The next step in determining the inlet and outlet VOC hourly molar flow rates ($M_{voc,in}$ and $M_{voc,out}$, respectively). As Equation 2.8 shows, $M_{voc,in}$ is a function of $y_{voc,in}$ and the total inlet volumetric flow rate, Q_{in}, (scfm).

Now, because the gasoline vapor flow rates are typically expressed in gallons/minute, we have to convert them to scfm. This is done as follows:

$$Q_{in} = Q_g\left(\frac{gal}{\min}\right) \times \frac{1\ ft^3}{7.48\ gal} = 0.134 Q_g\ scfm$$

Substituting these variables into Equation 2.8, we obtain:

$$M_{VOC,in} = \frac{0.134 Q_g}{392}(0.482)(60) = 0.00989 Q_g\left(\frac{lb - mole}{hr}\right)$$

We obtain $M_{VOC,out}$ from Equation 2.9:

$$M_{VOC,out} = 0.00989 Q_g\,(1 - 0.986) = 1.38 \times 10^{-4} Q_g\left(\frac{lb - moles}{hr}\right)$$

And according to Equation 2.10, the amount of gasoline vapor condensed ($M_{voc,con}$) is the difference between $M_{VOC,in}$ and $M_{voc,out}$:

$$M_{VOC,con} = 0.00975 Q_g \left(\frac{lb\text{-}moles}{hr} \right)$$

The final step is to calculate the condenser heat load. This load is a function of the inlet, outlet, and condensate molar flow rates, the inlet and conden-sation temperatures, the heat capacities of the VOC and air, and the VOC heat of condensation. The VOC heat capacity and heat of condensation data used are based on pentane and butane chemical properties, the largest components of gasoline, and were obtained from *CHRIS Hazardous Chemical Data* (U.S. Coast Guard, U.S. Department of Transportation, June 1985).

Heat capacities (Btu/lb-mole-°F):

$$C_{p,VOC} = 26.6$$
$$C_{p,air} = 6.95$$

Heat of condensation of VOC: 9,240 Btu/lb-mole

Substitution of these data, the molar flow rates, and the temperatures into Equations 2.12, 2.15 and 2.16 yields the following enthalpy changes in Btu/hr:

$$\Delta H_{con} = 130.8 Q_g$$
$$\Delta H_{uncon} = 0.572 Q_g$$
$$\Delta H_{noncon} = 11.6 Q_g$$

The condenser heat load (H_{load}) is the sum of these three enthalpy changes (Equation 2.11):

$$H_{load} = 143 Q_g$$

The refrigeration capacity (R, tons) is computed from Equation 2.22:

$$R = \frac{H_{load}}{12{,}000} = 0.0119 Q_g$$

This last equation relates the refrigeration capacity (tons) to the inlet gaso-line vapor flow rate (gal/min). Solving for Q_g, in terms of R, we obtain:

$$Q_g = 83.9 R$$

Finally, we substitute this relationship into the equipment cost ($) vs. vapor flow rate ($Q_g$) correlation, which-was developed from the vendor cost data:

$$\begin{aligned} EC_p &= 58.5Q_g + 212{,}000 \\ &= 58.5(83.9R) + 212{,}000 \\ &= 4{,}910R + 212{,}000 \end{aligned}$$

Note that this last expression is identical to Equation 2.34.

EPA/452/B-02-001

Section 3

VOC Controls

EPA/452/B-02-001

Section 3.2

VOC Destruction Controls

EPA/452/B-02-001

Chapter 1

Flares

Leslie B. Evans, Organic Chemicals Group
William M. Vatavuk, Innovative Strategies and Economics Group
Office of Air Quality Planning and Standards
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

Diana K. Stone
Susan K. Lynch
Richard F. Pandullo
Radian Corporation
Research Triangle Park, NC 27709

Walter Koucky
E.H. Pechan and Associates, Inc.
Durham, NC 27707

September 2000

Contents

1.1 Introduction 1-4
1.1.1 Flare Types 1-4
1.1.2 Applicability 1-6
1.1.3 Performance 1-7
1.1.3.1 Factors Affecting Efficiency 1-7
1.1.3.2 Flare Specifications 1-7

1.2 Process Description 1-8
1.2.1 Gas Transport Piping 1-9
1.2.2 Knock-out Drum 1-9
1.2.3 Liquid Seal 1-10
1.2.4 Flare Stack 1-10
1.2.5 Gas Seal 1-15
1.2.6 Burner Tip 1-15
1.2.7 Pilot Burners 1-15
1.2.8 Steam Jets 1-15
1.2.9 Controls 1-16

1.3 Design Procedures 1-17
1.3.1 Auxiliary Fuel Requirement 1-17
1.3.2 Flare Tip Diameter 1-18
1.3.3 Flare Height 1-19
1.3.4 Purge Gas Requirement 1-21
1.3.5 Pilot Gas Requirement 1-22
1.3.6 Steam Requirement 1-22
1.3.7 Knock-out Drum 1-23
1.3.8 Gas Mover System 1-25

1.4 Estimating Total Capital Investment 1-26
1.4.1 Equipment Costs 1-26
1.4.2 Installation Costs 1-28

1.5 Estimating Total Annual Costs 1-32
1.5.1 Direct Annual Costs 1-32
1.5.2 Indirect Annual Costs 1-33

1.6 Example Problem 1-35
1.6.1 Required Information for Design 1-35
1.6.2 Capital Equipment 1-35
1.6.2.1 Equipment Design 1-36
1.6.2.2 Equipment Costs 1-38
1.6.3 Operating Requirements 1-39
1.6.4 Total Annual Costs 1-41

1.7 Acknowledgments 1-43

References 1-43

1.1 Introduction

Flaring is a volatile combustion control process for organic compound (VOC) in which the VOCs are piped to a remote, usually elevated, location and burned in an open flame in the open air using a specially designed burner tip, auxiliary fuel, and steam or air to promote mixing for nearly complete (> 98%) VOC destruction. Completeness of combustion in a flare is governed by flame temperature, residence time in the combustion zone, turbulent mixing of the components to complete the oxidation reaction, and available oxygen for free radical formation. Combustion is complete if all VOCs are converted to carbon dioxide and water. Incomplete combustion results in some of the VOC being unaltered or converted to other organic compounds such as aldehydes or acids.

The flaring process can produce some undesirable by-products including noise, smoke, heat radiation, light, sulfur oxides (SO_x), nitrogen oxides (NO_x), carbon monoxide (CO), and an additional source of ignition where not desired. However, by proper design these can be minimized.

1.1.1 Flare Types

Flares are generally categorized in two ways: (1) by the height of the flare tip (i.e., ground or elevated), and (2) by the method of enhancing mixing at the flare tip (i.e., steam-assisted, air-assisted, pressure-assisted, or non-assisted). Elevating the flare can prevent potentially dangerous conditions at ground level where the open flame (i.e., an ignition source) is located near a process unit. Further, the products of combustion can be dispersed above working areas to reduce the effects of noise, heat, smoke, and objectionable odors.

In most flares, combustion occurs by means of a diffusion flame. A diffusion flame is one in which air diffuses across the boundary of the fuel/combustion product stream toward the center of the fuel flow, forming the envelope of a combustible gas mixture around a core of fuel gas. This mixture, on ignition, establishes a stable flame zone around the gas core above the burner tip. This inner gas core is heated by diffusion of hot combustion products from the flame zone.

Cracking can occur with the formation of small hot particles of carbon that give the flame its characteristic luminosity. If there is an oxygen deficiency and if the carbon particles are cooled to below their ignition temperature, smoking occurs. In large diffusion flames, combustion product vortices can form around burning portions of the gas and shut off the supply of oxygen. This localized instability causes flame flickering, which can be accompanied by soot formation.

As in all combustion processes, an adequate air supply and good mixing are required to complete combustion and minimize smoke. The various flare designs differ primarily in their accomplishment of mixing.

Steam-Assisted Flares

Steam-assisted flares are single burner tips, elevated above ground level for safety reasons, that burn the vented gas in essentially a diffusion flame. They reportedly account for the majority of the flares installed and are the predominant flare type found in refineries and chemical plants.[1,2]

To ensure an adequate air supply and good mixing, this type of flare system injects steam into the combustion zone to promote turbulence for mixing and to induce air into the flame. Steam-assisted flares are the focus of the chapter and will be discussed in greater detail in Sections 1.2 through 1.4.

Air-Assisted Flares

Some flares use forced air to provide the combustion air and the mixing required for smokeless operation. These flares are built with a spider-shaped burner (with many small gas orifices) located inside but near the top of a steel cylinder two feet or more in diameter. Combustion air is provided by a fan in the bottom of the cylinder. The amount of combustion air can be varied by varying the fan speed. The principal advantage of the air-assisted flares is that they can be used where steam is not available. Although air assist is not usually used on large flares (because it is generally not economical when the gas volume is large[3]) the number of large air-assisted flares being built is increasing.[4]

Non-Assisted Flares

The non-assisted flare is just a flare tip without any auxiliary provision for enhancing the mixing of air into its flame. Its use is limited essentially to gas streams that have a low heat content and a low carbon/hydrogen ratio that burn readily without producing smoke.[5] These streams require less air for complete combustion, have lower combustion temperatures that minimize cracking reactions, and are more resistant to cracking.

Pressure-Assisted Flares

Pressure-assisted flares use the vent stream pressure to promote mixing at the burner tip. Several vendors now market proprietary, high pressure drop burner tip designs. If sufficient vent stream pressure is available, these flares can be applied to streams previously requiring steam or air assist for smokeless operation. Pressure-assisted flares generally (but not necessarily) have the burner arrangement at ground level, and consequently, must be located in a remote area of the plant where there is plenty of space available. They have multiple burner heads that are staged to operate based on the quantity of gas being released. The size, design, number, and group arrangement of the burner heads depend on the vent gas characteristics.

Enclosed Ground Flares

An enclosed flare's burner heads are inside a shell that is internally insulated. This shell reduces noise, luminosity, and heat radiation and provides wind protection. A high nozzle pressure drop is usually adequate to provide the mixing necessary for smokeless operation and air or steam assist is not required. In this context, enclosed flares can be considered a special class of pressure-assisted or non-assisted flares. The height must be adequate for creating enough draft to supply sufficient air for smokeless combustion and for dispersion of the thermal plume. These flares are always at ground level.

Enclosed flares generally have less capacity than open flares and are used to combust continuous, constant flow vent streams, although reliable and efficient operation can be attained over a wide range of design capacity. Stable combustion can be obtained with lower Btu content vent gases than is possible with open flare designs (50 to 60 Btu/scf has been reported)[2], probably due to their isolation from wind effects. Enclosed flares are typically found at landfills.

1.1.2 Applicability

Flares can be used to control almost any VOC stream, and can handle fluctuations in VOC concentration, flow rate, heating value, and inerts content. Flaring is appropriate for continuous, batch, and variable flow vent stream applications. The majority of chemical plants and refineries have existing flare systems designed to relieve emergency process upsets that require release of large volumes of gas. These large diameter flares designed to handle emergency releases, can also be used to control vent streams from various process operations. Consideration of vent stream flow rate and available pressure must be given for retrofit applications. Normally, emergency relief flare systems are operated at a small percentage of capacity and at negligible pressure. To consider the effect of controlling an additional vent stream, the maximum gas velocity, system pressure, and ground level heat radiation during an emergency release must be evaluated. Further, if the vent stream pressure is not sufficient to overcome the flare system pressure, then the economics of a gas mover system must be evaluated, If adding the vent stream causes the maximum velocity limits or ground level heat radiation limits to be exceeded, then a retrofit application is not viable.

Many flare systems are currently operated in conjunction with baseload gas recovery systems. These systems recover and compress the waste VOC for use as a feedstock in other processes or as fuel. When baseload gas recovery systems are applied, the flare is used in a backup capacity and for emergency releases. Depending on the quantity of usable VOC that can be recovered, there can be a considerable economic advantage over operation of a flare alone.

Streams containing high concentrations of halogenated or sulfur containing compounds are not usually flared due to corrosion of the flare tip or formation of secondary pollutants (such as SO_2). If these vent types are to be controlled by combustion, thermal incineration, followed by scrubbing to remove the acid gases, is the preferred method.[3]

1.1.3 Performance

This section discusses the parameters that affect flare VOC destruction efficiency and presents the specifications that must be followed when flares are used to comply with EPA air emission standards.

1.1.3.1 Factors Affecting Efficiency

The major factors affecting flare combustion efficiency are vent gas flammability, auto-ignition temperature, heating value (Btu/scf), density, and flame zone mixing.

The flammability limits of the flared gases influence ignition stability and flame extinction. The flammability limits are defined as the stoichiometric composition limits (maximum and minimum) of an oxygen-fuel mixture that will burn indefinitely at given conditions of temperature and pressure without further ignition. In other words, gases must be within their flammability limits to burn. When flammability limits are narrow, the interior of the flame may have insufficient air for the mixture to burn. Fuels, such as hydrogen, with wide limits of flammability are therefore easier to combust.

For most vent streams, the heating value also affects flame stability, emissions, and flame structure. A lower heating value produces a cooler flame that does not favor combustion kinetics and is also more easily extinguished. The lower flame temperature also reduces buoyant forces, which reduces mixing.

The density of the vent stream also affects the structure and stability of the flame through the effect on buoyancy and mixing. By design, the velocity in many flares is very low; therefore, most of the flame structure is developed through buoyant forces as a result of combustion. Lighter gases therefore tend to burn better. In addition to burner tip design, the density also directly affects the minimum purge gas required to prevent flashback, with lighter gases requiring more purge.[5]

Poor mixing at the flare tip is the primary cause of flare smoking when burning a given material. Streams with high carbon-to-hydrogen mole ratio (greater than 0.35) have a greater tendency to smoke and require better mixing for smokeless flaring.[3] For this reason one generic steam-to-vent gas ratio is not necessarily appropriate for all vent streams. The required steam rate is dependent on the carbon to hydrogen ratio of the gas being flared. A high ratio requires more steam to prevent a smoking flare.

1.1.3.2 Flare Specifications

At too high an exit velocity, the flame can lift off the tip and flame out, while at too low a velocity, it can burn back into the tip or down the sides of the stack.

The EPA requirements for flares used to comply with EPA air emission standards are specified in 40 CFR Section 60.18. The requirements are for steam-assisted, air-assisted, and non-assisted flares. Requirements for steam-assisted, elevated flares state that the flare shall be designed for and operated with:

- an exit velocity at the flare tip of less than 60 ft/sec for 300 Btu/scf gas streams and less than 400 ft/sec for >1,000 Btu/scf gas streams. For gas streams between 300-1,000 Btu/scf the maximum permitted velocity (V_{max}, in ft/sec) is determined by the following equation:

$$\log_{10}(V_{max}) = \frac{B_v + 1{,}214}{852} \qquad (7.1)$$

- where B_v is the net heating value in Btu/scf.

- no visible emissions. A five-minute exception period is allowed during any two consecutive hours.

- a flame present at all times when emissions may be vented. The presence of a pilot flame shall be monitored using a thermocouple or equivalent device.

- the net heating value of the gas being combusted being 300 Btu/scf or greater.

In addition, owners or operators must monitor to ensure that flares are operated and maintained in conformance with their design.

1.2 Process Description

The elements of an elevated steam-assisted flare generally consist of gas vent collection piping, utilities (fuel, steam, and air), piping from the base up, knock-out drum, liquid seal, flare stack, gas seal, burner tip, pilot burners, steam jets, ignition system, and controls. Figure 7.1 is a diagram of a steam-assisted elevated smokeless flare system showing the usual components that are included.

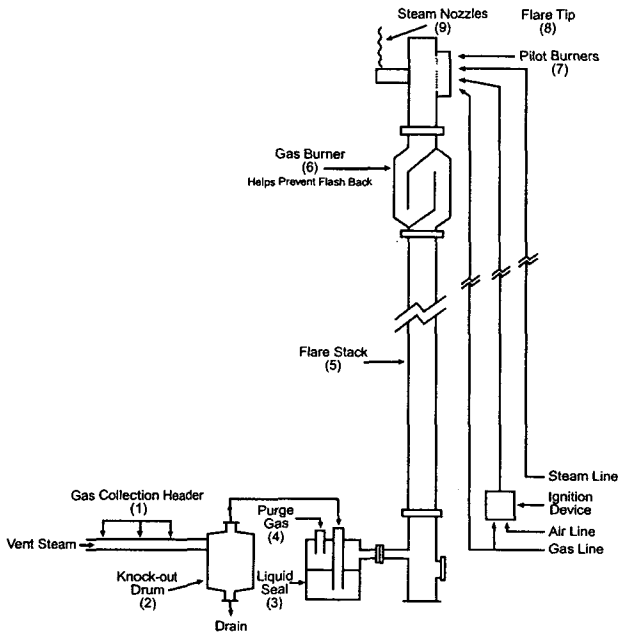


Figure 1.1: Steam-Assisted Elevated Flare System

1.2.1 Gas Transport Piping

Process vent streams are sent from the facility release point to the flare location through the gas collection header. The piping (generally schedule 40 carbon steel) is designed to minimize pressure drop. Ducting is not used as it is more prone to air leaks. Valving should be kept to an absolute minimum and should be "car-sealed" (sealed) open. Pipe layout is designed to avoid any potential dead legs and liquid traps. The piping is equipped for purging so that explosive mixtures do not occur in the flare system either on start-up or during operation.

1.2.2 Knock-out Drum

Liquids that may be in the vent stream gas or that may condense out in the collection header and transfer lines are removed by a knock-out drum. (See Figure 1.2.) The knock-out or disentrainment drum is typically either a horizontal or vertical vessel located at or close to the base of the flare, or a vertical vessel located inside the base of the flare stack. Liquid in the vent stream can extinguish the flame or cause iregular combustion and smoking. \in addition, flaring liquids can generate a spray of burning chemicals that could reach group level and create a safety hazard. For a flare system designed to handle emergency process upsets this drum must be sized for worst-case conditions (e.g., loss of cooling water or total unit depressuring) and is usually quite large. For a flare system devoted only to vent stream VOC control, the sizing of the drum is based primarily on vent gas fow rate with consideration given to liquid entrainment.

1.2.3 Liquid Seal

Process vent streams are usually passed through a liquid seal before going to the flare stack. The liquid seal can be downstream of the knockout drum or incorporated into the same vessel. This prevents possible flame flashbacks, caused when air is inadvertently introduced into the flare system and the flame front pulls down into the stack. The liquid seal also serves to maintain a positive pressure on the upstream system and acts as a mechanical damper on any explosive shock wave in the flare stack. Other devices, such as flame arresters and check valves, may sometimes replace a liquid seal or be used in conjunction wit hit. Purge gas (as discussed in Section 1.3.4) also helps to prevent flashback in the flare stack cause by low vent gas flow.

1.2.4 Flare Stack

For safety reasons a stack is used to elevate the flare. The flare must be located so that it does not present a hazard to surrounding personnel and facilities. Elevated flares can be self-supported (free-standing), guyed, or structurally supported by a derrick. Examples of these three types of elevated flares are shown in Figures 1.3, 1.4, and 1.5 for self-supported, derrick supported, and guy-supported flares, respectively. Self-supporting flares are generally used for lower flare tower heights (30-100 feet) but can be designed for up to 250 feet. Guy towers are designed for over 300 feet, while derrick towers are designed for above 200 feet.[4, 6, 7, 8, 9, 10]

Free-standing flares provide ideal structural support. However, for very high units the costs increase rapidly. In addition, the foundation required and nature of the soil must be considered.

Derrick-supported flares can be built as high as required since the system load is spread over the derrick structure. This design provides for differential expansion between the stack, piping, and derrick. Derrick-supported flares are the most expensive design for a given flare height.

The guy-supported flare is the simplest of all the support methods. However, a considerable amount of land is required since the guy wires are widely spread apart. A rule of thumb for space required to erect a guy-supported flare is a circle on the ground with a radius equal to the height of the flare stack.[6]

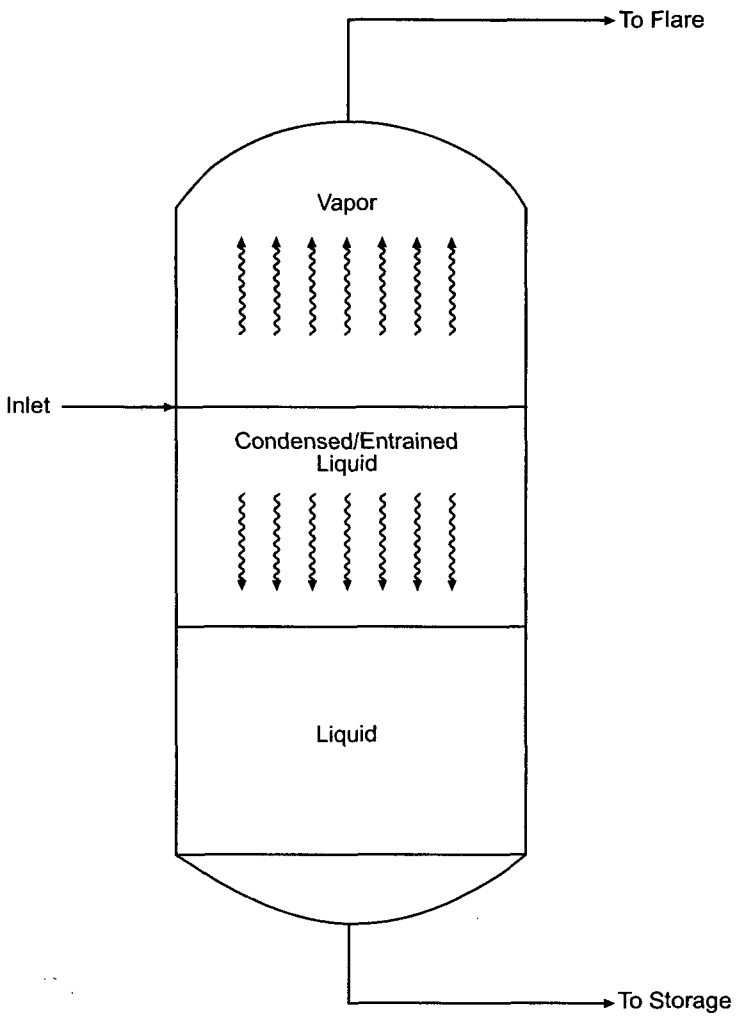


Figure 1.2: Typical Vertical Knock-out Drum

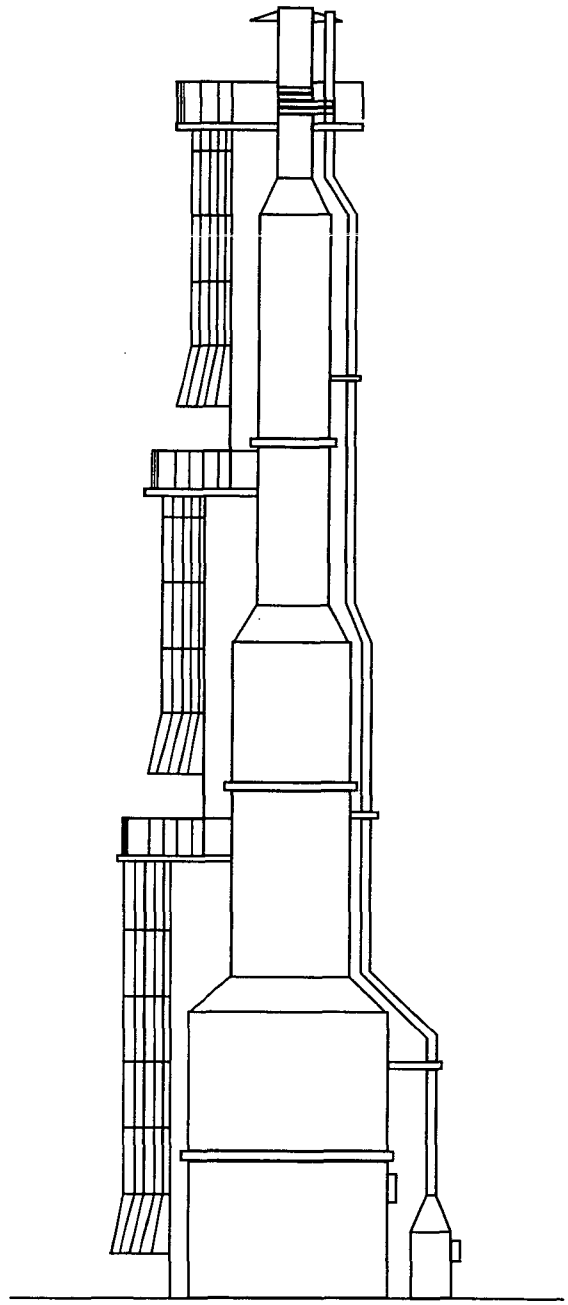


Figure 1.3: Self-Supported Elevated Flare

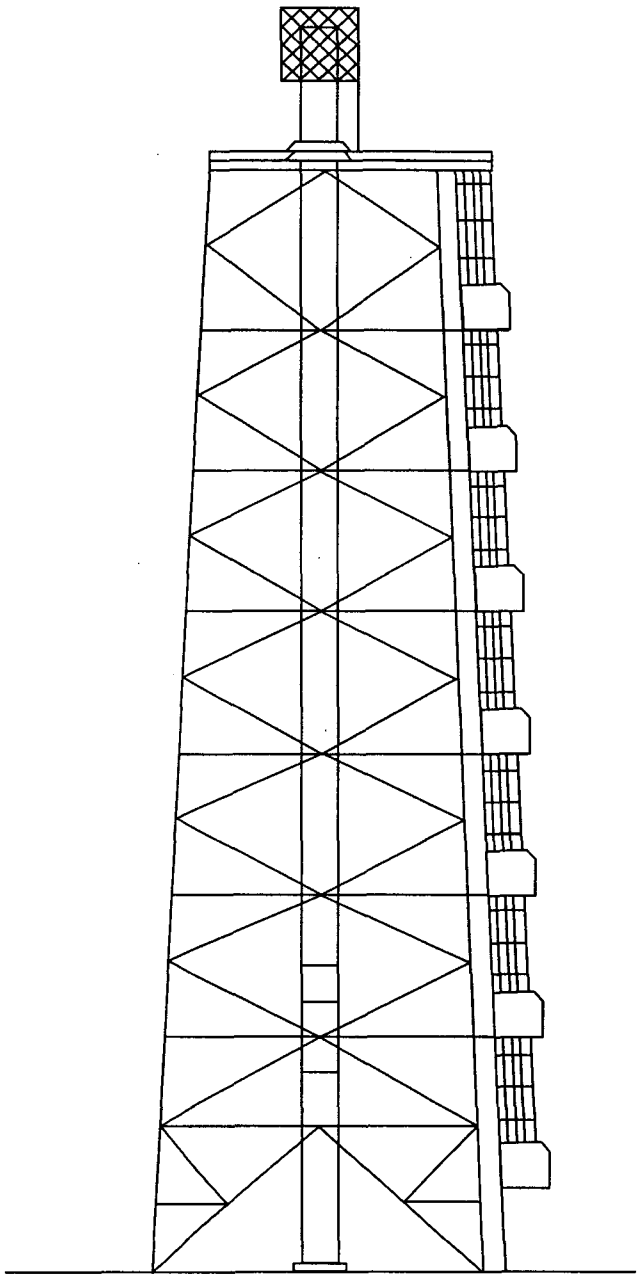


Figure 1.4: Derrick-Supported Elevated Flare

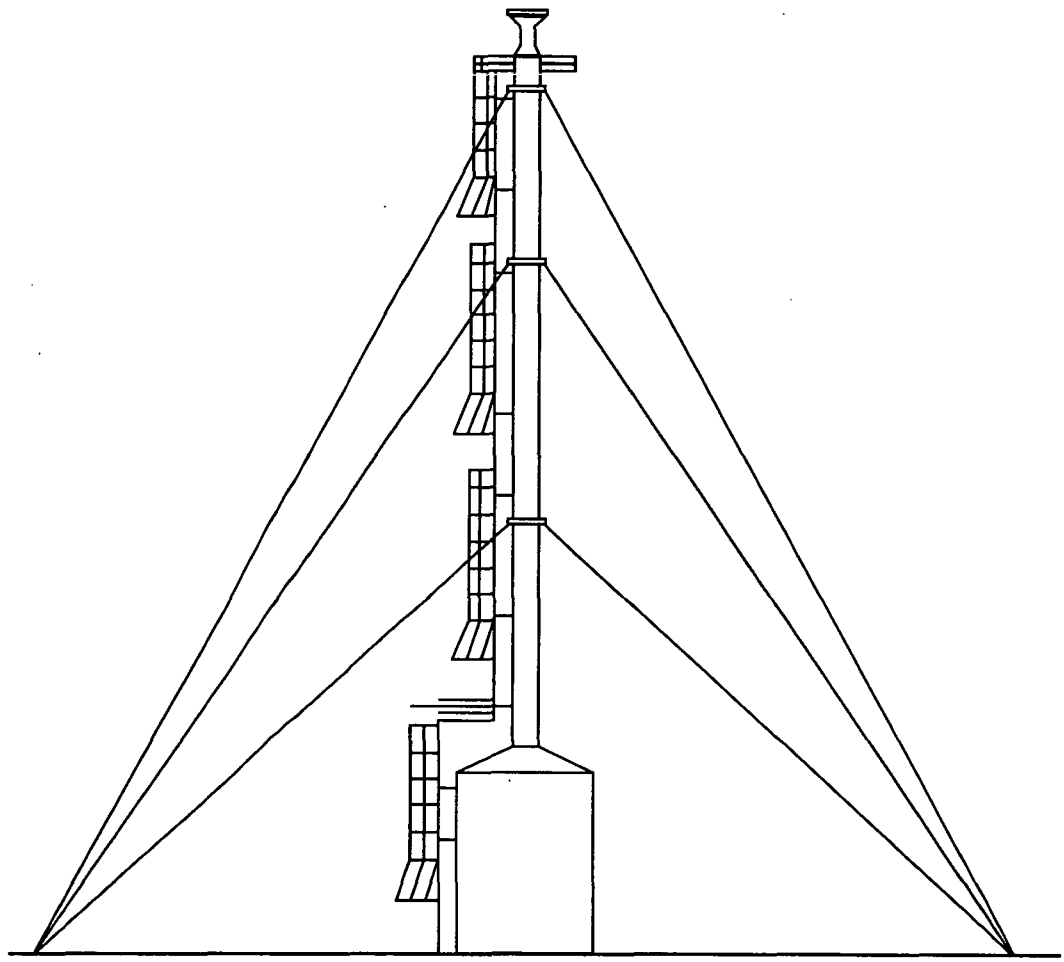


Figure 1.5: Guy-Supported Elevated Flare

1.2.5 Gas Seal

Air may tend to flow back into a flare stack due to wind or the thermal contraction of stack gases and create an explosion potential. To prevent this, a gas seal is typically installed in the flare stack. One type of gas seal (also referred to as a flare seal, stack seal, labyrinth seal, or gas barrier) is located below the flare tip to impede the flow of air back into the flare gas network. There are also "seals" which act as orifices in the top of the stack to reduce the purge gas volume for a given velocity and also interfere with the passage of air down the stack from the upper rim. These are known by the names "internal gas seal, fluidic-seal, and arrestor seal".[5] These seals are usually proprietary in design, and their presence reduces the operating purge gas requirements.

1.2.6 Burner Tip

The burner tip, or flare tip, is designed to give environmentally acceptable combustion of the vent gas over the flare system's capacity range. The burner tips are normally proprietary in design. Consideration is given to flame stability, ignition reliability, and noise suppression. The maximum and minimum capacity of a flare to burn a flared gas with a stable flame (not necessarily smokeless) is a function of tip design. Flame stability can be enhanced by flame holder retention devices incorporated in the flare tip inner circumference. Burner tips with modern flame holder designs can have a stable flame over a flare gas exit velocity range of 1 to 600 ft/sec.[2] The actual maximum capacity of a flare tip is usually limited by the vent stream pressure available to overcome the system pressure drop. Elevated flares diameters are normally sized to provide vapor velocities at maximum throughput of about 50 percent of the sonic velocity of the gas subject to the constraints of CFR 60.18.[1]

1.2.7 Pilot Burners

EPA regulations require the presence of a continuous flame. Reliable ignition is obtained by continuous pilot burners designed for stability and positioned around the outer perimeter of the flare tip. The pilot burners are ignited by an ignition source system, which can be designed for either manual or automatic actuation. Automatic systems are generally activated by a flame detection device using either a thermocouple, an infra-red sensor or, more rarely, (for ground flare applications) an ultra-violet sensor.[4]

1.2.8 Steam Jets

A diffusion flame receives its combustion oxygen by diffusion of air into the flame from the surrounding atmosphere. The high volume of fuel flow in a flare may require more combustion air at a faster rate than simple gas diffusion can supply. High velocity steam injection nozzles, positioned

around the outer perimeter of the flare tip, increase gas turbulence in the flame boundary zones, drawing in more combustion air and improving combustion efficiency. For the larger flares, steam can also be injected concentrically into the flare tip.

The injection of steam into a flare flame can produce other results in addition to air entrainment and turbulence. Three mechanisms in which steam reduces smoke formation have been presented.[1] Briefly, one theory suggests that steam separates the hydrocarbon molecule, thereby minimizing polymerization, and forms oxygen compounds that burn at a reduced rate and temperature not conducive to cracking and polymerization. Another theory claims that water vapor reacts with the carbon particles to form CO, CO_2, and H_2, thereby removing the carbon before it cools and forms smoke. An additional effect of the steam is to reduce the temperature in the core of the flame and suppress thermal cracking.[5] The physical limitation on the quantity of steam that can be delivered and injected into the flare flame determines the smokeless capacity of the flare. Smokeless capacity refers to the volume of gas that can be combusted in a flare without smoke generation. The smokeless capacity is usually less than the stable flame capacity of the burner tip.

Significant disadvantages of steam usage are the increased noise and cost. Steam aggravates the flare noise problem by producing high-frequency jet noise. The jet noise can be reduced by the use of small multiple steam jets and, if necessary, by acoustical shrouding. Steam injection is usually controlled manually with the operator observing the flare (either directly or on a television monitor) and adding steam as required to maintain smokeless operation. To optimize steam usage infrared sensors are available that sense flare flame characteristics and adjust the steam flow rate automatically to maintain smokeless operation. Automatic control, based on flare gas flow and flame radiation, gives a faster response to the need for steam and a better adjustment of the quantity required. If a manual system is used, steam metering should be installed to significantly increase operator awareness and reduce steam consumption.

1.2.9 Controls

Flare system control can be completely automated or completely manual. Components of a flare system which can be controlled automatically include the auxiliary gas, steam injection, and the ignition system. Fuel gas consumption can be minimized by continuously measuring the vent gas low rate and heat content (Btu/scf) and automatically adjusting the amount of auxiliary fuel to maintain the required minimum of 300 Btu/scf for steam-assisted flares. Steam consumption can likewise be minimized by controlling flow based on vent gas flow rate. Steam flow can also be controlled using visual smoke monitors. Automatic ignition panels sense the presence of a flame with either visual or thermal sensors and reignite the pilots when flameouts occur.

1.3 Design Procedures

Flare design is influenced by several factors, including the availability of space, the characteristics of the flare gas (namely composition, quantity, and pressure level) and occupational concerns. The sizing of flares requires determination of the required flare tip diameter and height. The emphasis of this section will be to size a steam-assisted elevated flare for a given application.

1.3.1 Auxiliary Fuel Requirement

The flare tip diameter is a function of the vent gas flow rate plus the auxiliary fuel and purge gas flow rate. The purge gas flow rate is very small relative to the vent gas and fuel flow rates, so it may be ignored when determining the tip diameter. The flow rate of the auxiliary fuel, if required, is significant, and must be calculated before the tip diameter can be computed.

Some flares are provided with auxiliary fuel to combust hydrocarbon vapors when a lean flare gas stream falls below the flammability range or heating value necessary to sustain a stable flame. The amount of fuel required, F, is calculated based on maintaining the vent gas stream net heating value at the minimum of 300 Btu/scf required by rules defined in the Federal Register (see next section):

$$Q\,B_v + F\,B_f = (Q + F)\left(300\ \frac{Btu}{scf}\right) \qquad (1.2)$$

where

Q = the vent stream flow rate, scfm

B_v and B_f are the Btu/scf of the vent stream and fuel, respectively.

Rearranging gives:

$$F\ (scfm) = Q\left(\frac{300 - B_v}{B_f - 300}\right) \qquad (1.3)$$

The annual auxiliary fuel requirement, F_a, is calculated by:

$$F_a\left(\frac{\text{M scf}}{\text{yr}}\right) = F(\text{scfm})\ 60\left(\frac{\text{min}}{\text{hr}}\right) 8760\left(\frac{\text{hr}}{\text{yr}}\right) = 526\ F\ \frac{\text{scfm}}{\text{yr}} \qquad (1.4)$$

Typical natural gas has a net heating value of about 1,000 Btn/scf. Automatic control of the auxiliary fuel is ideal for processes with large fluctuations in VOC compositions. These flares are used for the disposal of such streams as sulfur tail gases and ammonia waste gases, as well as any low Btu vent streams.[2]

1.3.2 Flare Tip Diameter

Flare tip diameter is generally sized on a velocity basis, although pressure drop must also be checked. Flare tip sizing for flares used to comply with EPA air emission standards is governed by rules defined in the Federal Register (see 40 CFR 60.18). To comply with these requirements, the maximum velocity of a steam-assisted elevated flare is given in Table 1.1:

Table 1.1: Maximum Velocity of Steam-Assisted Elevated Flare

Net Heating Value of Vent Stream B_v *(Btu/scf)*	**Maximum Velocity V_{max} *(ft/sec)***
300 300 - 1,000 > 1,000	$log_{10}(V_{max}) = \frac{(B_v + 1{,}214)}{852}$

By determining the maximum allowed velocity, V_{max} (ft/sec), and knowing the total volumetric flow rate, Q_{tot} (acfm), including vent stream and auxiliary fuel gas, a minimum flare tip diameter, D_{min} (in), can be calculated. It is standard practice to size the flare so that the design velocity of flow rate Q_{tot}, is 80 percent of V_{max}, i.e.:

$$D_{min}\ (in) = 12\sqrt{\frac{\frac{4}{\pi}\ \frac{Q_{tot}}{60\left(\frac{sec}{min}\right)}}{0.8\ V_{max}}} = 1.95\sqrt{\frac{Q_{tot}}{V_{max}}} \qquad (1.5)$$

where

Q_{tot} = $Q + F$ (measured at stream temperature and pressure)

The flare tip diameter, D, is the calculated diameter, $D = D_{min}$, rounded up to the next commercially available size. The minimum flare size is 1 inch; larger sizes are available in 2-inch increments from 2 to 24 inches and in 6-inch increments above 24 inches. The maximum size commercially available is 90 inches.[5]

A pressure drop calculation is required at this point to ensure that the vent stream has sufficient pressure to overcome the pressure drop occurring through the flare system at maximum flow conditions. The pressure drop calculation is site specific but must take into account losses through the collection header and piping, the knock-out drum, the liquid seal, the flare stack, the gas seal, and finally the flare tip. Piping size should be assumed equal to the flare tip diameter. Schedule 40 carbon steel pipe is typically used. If sufficient pressure is not available, the economics of either a larger flare system (pressure drop is inversely proportional to the pipe diameter) or a mover such as a fan or compressor must be weighed. (Refer to Section 1.3.8 for typical pressure drop relationships.)

1.3.3 Flare Height

The height of a flare is determined based on the ground level limitations of thermal radiation intensity, luminosity, noise, height of surrounding structures, and the dispersion of the exhaust gases. In addition, consideration must also be given for plume dispersion in case of possible emission ignition failure. Industrial flares are normally sized for a maximum heat intensity of 1,500-2,000 Btu/hr-ft^2 when flaring at their maximum design rates.[1,2] At this heat intensity level, workers can remain in the area of the flare for a limited period only. If, however, operating personnel are required to remain in the unit area performing their duties, the recommended design flare radiation level excluding solar radiation is 500 Btu/hr-ft^2.[1] The intensity of solar radiation is in the range of 250-330 Btu/hr-ft^2.[1] Flare height may also be determined by the need to safely disperse the vent gas in case of flameout. The height in these cases would be based on dispersion modeling for the particular installation conditions and is not addressed here. The minimum flare height normally used is 30 feet.[5] Equation (1.6) by Hajek and Ludwig may be used to determine the minimum distance, L, required from the center of the flare flame and a point of exposure where thermal radiation must be limited.[1]

$$L^2\ (\text{ft}^2) = \frac{\tau\ f\ R}{4\ \pi\ K} \qquad (1.6)$$

where

J	=	fraction of heat intensity transmitted
f	=	fraction of heat radiated
R	=	net heat release (Btu/hr)
K	=	allowable radiation (500 Btu/hr-ft^2)

The conservative design approach used here ignores wind effects and calculates the distance assuming the center of radiation is at the base of the flame (at the flare tip), not in the center. It is also assumed that the location where thermal radiation must be limited is at the base of the flare. Therefore, the distance, L, is equal to the required flare stack height (which is a minimum of 30 feet). The f factor allows for the fact that not all the heat released in a flame can be released as radiation. Heat transfer is propagated through three mechanisms: conduction, convection, and radiation. Thermal radiation may be either absorbed, reflected, or transmitted. Since the atmosphere is not a perfect vacuum, a fraction of the heat radiated is not transmitted due to atmospheric absorption (humidity, particulate matter). For estimating purposes, however, assume all of the heat radiated is transmitted (i.e., $r = 1$). Table 1.2 is a summary of heat radiated from various gaseous diffusion flames:[1]

Table 1.2: Heat from Various Gaseous Difusion Flames

Gas	Flare Tip Diameter (in)	Fraction of Heat Radiated (f)
Hydrogen	<1	.10
	1.6	.11
	3.3	1.6
	8.0	1.5
	16.0	1.7
Butane		
	<1	.29
	1.6	.29
	3.3	.29
	8.0	.28
	16.0	.30
Methane		
	<1	.16
	1.6	.16
	3.3	.15
Natural Gas		
	8.0	.19
	16.0	.23

In general, the fraction of heat radiated increases as the stack diameter increases. If stream-specific data are not available, a design basis of $f = 0.2$ will give conservative results.[4] The heat release, R, is calculated from the flare gas flow rate, W, and the net heating value, B_v, as follows:

$$R\left(\frac{\text{Btu}}{\text{hr}}\right) = W\left(\frac{\text{lb}}{\text{hr}}\right) B_v\left(\frac{\text{Btu}}{\text{lb}}\right) \tag{1.7}$$

1.3.4 Purge Gas Requirement

The total volumetric flow to the flame must be carefully controlled to prevent low flow flashback problems and to avoid flame instability. Purge gas, typically natural gas, N_2, or CO_2, is used to maintain a minimum required positive flow through the system. If there is a possibility of air in the flare manifold, N_2, another inert gas, or a flammable gas must be used to prevent the formation of an explosive mixture in the flare system. To ensure a positive flow through all flare components, purge gas injection should be at the farthest upstream point in the flare transport piping.

The minimum continuous purge gas required is determined by the design of the stack seals, which are usually proprietary devices. Modern labyrinth and internal gas seals are stated to require a gas velocity of 0.001 to 0.04 ft/sec (at standard conditions).[6, 7, 8, 9, 10] Using the conservative value of 0.04 ft/sec and knowing the flare diameter (in), the annual purge gas volume, F_{pu}, can be calculated:

$$\begin{aligned} F_{pu}\left(\frac{M\text{scf}}{\text{yr}}\right) &= \left(0.04\ \frac{\text{ft}}{\text{sec}}\right)\left(\frac{\frac{\pi D^2}{4}}{144}\text{ft}^2\right)\left(3{,}600\ \frac{\text{sec}}{\text{hr}}\right)\left(8{,}760\ \frac{\text{hr}}{\text{yr}}\right) \\ &= 6.88D^2\left(\frac{M\text{scf}}{\text{yr}}\right) \end{aligned} \tag{1.8}$$

There is another minimum flare tip velocity for operation without burn lock or instability. This minimum velocity is dependent on both gas composition and diameter and can range from insignificant amounts on small flares to 0.5 ft/sec on greater than 60-inch diameter units.[5]

Purge gas is also required to clear the system of air before startup, and to prevent a vacuum from pulling air back into the system after a hot gas discharge is flared. (The cooling of gases within the flare system can create a vacuum.) The purge gas consumption from these uses is assumed to be minor.

1.3.5 Pilot Gas Requirement

The number of pilot burners required depends on flare size and, possibly, on flare gas composition and wind conditions. Pilot gas usage is a function of the number of pilot burners required to ensure positive ignition of the flared gas, of the design of the pilots, and of the mode of operation. The average pilot gas consumption based on an energy-efficient model is 70 scf/hr (of typical 1000 Btu per scf gas) per pilot burner.[6, 7, 8, 9, 10] The number of pilot burners, N, based on flare size is:[6, 7, 8, 9, 10]

Table 1.3: Number of Burners by Flam Tip Diameter

Flare Tip Diameter (in)	Number of Pilot Burners (N)
1-10	1
12-24	2
30-60	3
>60	4

The annual pilot gas consumption, F_{pi} is calculated by:

$$F_{pi}\left(\frac{\text{Mscf}}{\text{yr}}\right) = \left(70\ \frac{\text{scf}}{\text{hr}}\right)(N)\left(8{,}760\ \frac{\text{hr}}{\text{yr}}\right) = \left(613\ \frac{\text{scf}}{\text{yr}}\right) N \qquad (1.9)$$

1.3.6 Steam Requirement

The steam requirement depends on the composition of the vent gas being flared, the steam velocity from the injection nozzle, and the flare tip diameter. Although some gases can be flared smokelessly without any steam, typically 0.01 to 0.6 pound of steam per pound of flare gas is required.[6, 7, 8, 9, 10] The ratio is usually estimated from the molecular weight of the gas, the carbon-to-hydrogen ratio of the gas, or whether the gas is saturated or unsaturated. For example, olefins, such as propylene, require higher steam ratios than would paraffin hydrocarbons to burn smokelessly.[2]

In any event, if a proprietary smokeless flare is purchased, the manufacturer should be consulted about the minimum necessary steam rate. A small diameter flare tip (less than 24 inches) can use steam more effectively than a large diameter tip to mix air into the flame and promote turbulence.[2] For a typical refinery, the average steam requirement is typically 0.25 lb/lb, with this number increasing to 0.5 lb/lb in chemical plants where large quantities of unsaturated hydrocarbons are flared.[10]

For general consideration, the quantity of steam required, S, can be assumed to be 0.4 pounds of steam per pound of flare gas, W Using a 0.4 ratio, the amount of steam required is:

$$S\left(\frac{\text{lbs}}{\text{yr}}\right) = \left(0.4\frac{\text{lb steam}}{\text{lb flare gas}}\right)\left(W\frac{\text{lb}}{\text{yr}}\right)\left(8{,}760\frac{\text{hr}}{\text{yr}}\right) \qquad (1.10)$$

Operating a flare at too high a steam-to-gas ratio is not only costly, but also results in a lower combustion efficiency and a noise nuisance. The capacity of a steam-assisted flare to burn smokelessly may be limited by the quantity of steam that is available.

1.3.7 Knock-out Drum

As explained previously, the knock-out drum is used to remove any liquids that may be in the vent stream. Two types of drums are used: horizontal and vertical. The economics of vessel design influences the choice between a horizontal and a vertical drum. When a large liquid storage vessel is required and the vapor flow is high, a horizontal drum is usually more economical. Vertical separators are used when there is small liquid load, limited plot space, or where ease of level control is desired. It is assumed here that the drum is not sized for emergency releases and that liquid flow is minimal. Flares designed to control continuous vent streams generally have vertical knockout drums, whereas emergency flares typically have horizontal vessels. The procedure described below applies to vertical drums exclusively. A typical vertical knock-out drum is presented in Figure 1.2.

Liquid particles will separate when the residence time of the vapor is greater than the time required to travel the available vertical height at the dropout velocity of the liquid particles, *i.e.*, the velocity is less than the dropout velocity. In addition, the vertical gas velocity must be sufficiently low to permit the liquid droplets to fall. Since flares are designed to handle small-sized liquid droplets, the allowable vertical velocity is based on separating droplets from 300 to 600 micrometers in diameter.[1] The dropout velocity, U, of a particle in a stream, or the maximum design vapor velocity, is calculated as follows:[11]

$$U\left(\frac{\text{ft}}{\text{sec}}\right) = G\sqrt{\frac{\rho_l - \rho_v}{\rho_v}} \qquad (1.11)$$

where

G = design vapor velocity factor
p_l and p_v = liquid and vapor densities, lb/ft^3

Note that in most cases,

$$\frac{\rho_l - \rho_v}{\rho_v} \approx \frac{\rho_l}{\rho_v} \tag{1.11a}$$

The design vapor velocity factor, G, ranges from 0.15 to 0.25 for vertical gravity separators at 85% of flooding.[11]

Once the maximum design vapor velocity has been determined the minimum vessel cross-sectional area, A, can be calculated by:

$$A\ (\text{ft}^2) = \frac{Q_a \left(\frac{\text{ft}}{\text{min}}\right)}{\left(60\ \frac{\text{sec}}{\text{min}}\right)\left(U\ \frac{\text{ft}}{\text{sec}}\right)} \tag{1.12}$$

where Q_a is the vent stream flow in actual ft^3/min, or Q adjusted to the vent stream temperature and pressure.

The vessel diameter, d_{min}, is then calculated by:

$$d_{min} = \sqrt{\frac{4}{\pi} A} \tag{1.13}$$

In accordance with standard head sizes, drum diameters in 6-inch increments are assumed so:

$$d = d_{min}\ \left(rounded\ to\ the\ next\ largest\ size\right) \tag{1.14}$$

Some vertical knockout drums are sized as cyclones and utilize a tangential inlet to generate horizontal separating velocities. Vertical vessels sized exclusively on settling velocity (as in the paragraph above) will be larger than those sized as cyclones.[5]

The vessel thickness, t, is determined from the diameter as showen in Table 1.4 [15]. Proper vessel height, h, is usually determined based on required liquid surge volume. The calculated height is then checked to verify that the height-to-diameter ratio is within the economic range of 3 to 5.[11] For small volumes of liquid, as in the case of continuous VOC vent control, it is necessary to provide more liquid surge than is necessary to satisfy the h/d > 3 condition. So for purposes of flare knock-out drum sizing:

$$h = 3d \tag{1.15}$$

Table 1.4: Vessel Thickness based on Diameter

Diameter, *d* (inches)	Thickness, *t* (inches)
d<36	0.25
36<d<72	0.37
72<d<108	50.5
108<d<144	0.75
d>144	1.0

1.3.8 Gas Mover System

The total system pressure drop is a function of the available pressure of the vent stream, the design of the various system components, and the flare gas flow rate. The estimation of actual pressure drop requirements involves complex calculations based on the specific system's vent gas properties and equipment used. For the purposes of this section, however, approximate values can be used. The design pressure drop through the flare tip can range from 0.1 to 2 psi with the following approximate pressure drop relationships:[5] The total system pressure drop ranges from about 1 to 25 psi.[5]

Table 1.5: Design Pressure Losses through the Flare Tip

Equipment	Approximate Pressure Loss
Gas seal:	1 to 3 times flare tip pressure drop
Stack:	0.25 to 2 times flare tip pressure drop
Liquid seal and Knock out drum:	1 to 1.5 times flare tip pressure drop *plus* pressure drop due to liquid depth in the seal, which is normally 0.2 to 1.5 psi.
Gas collection system:	calculated based on diameter, length, and flow. System is sized by designer to utilize the pressure drop available and still leave a pressure at the stack base of between 2 and 10 psi.

1.4 Estimating Total Capital Investment

The capital costs of a flare system are presented in this section and are based on the design/sizing procedures discussed in Section 7.3. The costs presented are in September 2000 dollars.[1] The capital costs for this Chapter were updated through vender contacts in the summer of 2000. The costs were updated by sending vendors tables and graphs of previous cost equations and asking for updated information.

Vendors reported that costs had not increased subastantially since 1990, the primary reasons cited for stable prices ere increased competitions and lower steel prices. One vendor reported slight price increases over the period and another reported slight price decreases. Vendors agreed that the costs developed in 1990 reflected current market conditions. Items not such as platforms and ladders could result in some variation around these prices. Based on the information supplied by the vendors, the 1990 prices are carried forward for the year 2000 and are presented in Tables 1.6 to 1.8 and Figures 1.5 to 1.7. [2][7]

Total capital investment, TCI, includes the equipment costs, EC, for the flare itself, the cost of auxiliary equipment, the cost of taxes, freight, and instrumentation, and all direct and indirect installation costs.

The capital cost of flares depends on the degree of sophistication desired (i.e., manual vs automatic control) and the number of appurtenances selected, such as knock-out drums, seals, controls, ladders, and platforms. The basic support structure of the flare, the size and height, and the auxiliary equipment are the controlling factors in the cost of the flare. The capital investment will also depend on the availability of utilities such as steam, natural gas, and instrument air.

The total capital investment is a battery limit cost estimate and does not include the provisions for bringing utilities, services, or roads to the site, the backup facilities, the land, the research and development required, or the process piping and instrumentation interconnections that may be required in the process generating the waste gas. These costs are based on a new plant installation; no retrofit cost considerations such as demolition, crowded construction working conditions, scheduling construction with production activities, and long interconnecting piping are included. These factors are so site-specific that no attempt has been made to provide their costs.

1.4.1 Equipment Costs

Flare vendors were asked to provide budget estimates for the spectrum of commercial flare sizes. These quotes [6, 7, 8, 9, 10] were used to develop the equipment cost correlations for flare units, while the cost equations for the auxiliary equipment were based on references [12] and

[1]For information on escalating these prices to more current dollars, refer to the EPA report Escalation Indexes for Air Pollution Control Costs and updates thereto, all of which are installed on the OAQPS Technology Transfer Network (CTC Bulletin Board).

[13] (knock-out drums) and [14] and [15] (piping). The expected accuracy of these costs is ± 30% (i.e., "study" estimates). Keeping in mind the height restrictions discussed in Section 7.2.4, these cost correlations apply to flare tip diameters ranging from 1 to 60 inches and stack heights ranging from 30 to 500 feet. The standard construction material is carbon steel except when it is standard practice to use other materials, as is the case with burner tips.

The flare costs, C_F presented in Equations 1.16 through 1.18 are calculated as a function of stack height, L (ft) (30 ft minimum), and tip diameter, D (in), and are based on support type as follows:

Self Support Group:

$$C_F\ (\$) = (78.0 + 9.14D + 0.749L)^2 \qquad (1.16)$$

Guy Support Group:

$$C_F\ (\$) = (103 + 8.68D + 0.470L)^2 \qquad (1.17)$$

Derrick Support Group:

$$C_F\ (\$) = (76.4 + 2.72D + 1.64L)^2 \qquad (1.18)$$

The equations are least-squares regression of cost data provided by different vendors. It must be kept in mind that even for a given flare technology (i.e., elevated, steam-assisted), design and manufacturing procedures vary from vendor to vendor, so that costs may vary. Once a study estimate is completed, it is recommended that several vendors be solicited for more detailed cost estimates.

Each of these costs includes the flare tower (stack) and support, burner tip, pilots, utility (steam, natural gas) piping from base, utility metering and control, liquid seal, gas seal, and galvanized caged ladders and platforms as required. Costs are based on carbon steel construction, except for the upper four feet and burner tip, which are based on 310 stainless steel.

The gas collection header and transfer line requirements are very site specific and depend on the process facility where the emission is generated and on where the flare is located. For the purposes of estimating capital cost it is assumed that the transfer line will be the same diameter as the flare tip[6] and will be 100 feet long. Most installations will require much more extensive piping, so 100 feet is considered a minimum.

The costs for vent stream piping, C_p, are presented separately in Equation 1.19 or 1.20 and are a function of pipe, or flare, diameter D.[15]

$$C_p\ (\$) = 127D^{1.21}\ (where\ 1'' < D < 24'') \qquad (1.19)$$

$$C_p\ (\$) = 139D^{1.07}\ (where\ 30'' < D < 60'') \qquad (1.20)$$

The costs, C_p, include straight, Schedule 40, carbon steel pipe only, are based on 100 feet of piping, and are directly proportional to the distance required.

The costs for a knock-out drum, C_K, are presented separately in Equation 7.22 and are a function of drum diameter, d (in), and height, h (in).[12, 13]

$$C_K\ (\$) = 14.2\ [dt\ (h + 0.812d)\ ]^{0.737} \qquad (1.21)$$

where t is the vessel thickness, in inches, determined based on the diameter.

Flare system equipment cost, EC, is the total of the calculated flare, knock-out drum, and piping costs.

$$EC\ (\$) = C_F + C_K + C_p \qquad (1.22)$$

Purchased equipment costs, PEC, is equal to equipment cost, EC, plus factors for ancillary instrumentation (i.e., control room instruments) (.10), sales taxes (0.03), and freight (0.05) or,

$$PEC\ (\$) = EC\ (1 + 0.10 + 0.03 + 0.05) = 1.18\ EC \qquad (1.23)$$

1.4.2 Installation Costs

The total capital investment, TCI, is obtained by multiplying the purchased equipment cost, PEC, by an installation factor of 1.92.

$$TCI\ (\$) = 1.92\ PEC \qquad (1.24)$$

These costs were determined based on the factors in Table 1.6. The bases used in calculating annual cost factors are given in Table 1.5. These factors encompass direct and indirect installation costs. Direct installation costs cover foundations and supports, equipment handling and erection, piping, insulation, painting, and electrical. Indirect installation costs cover engineering, construction and field expenses, contractor fees, start-up, performance testing, and contingencies. Depending

on the site conditions, the installation costs for a given flare could deviate significantly from costs generated by these average factors. Vatavuk and Neveril provide some guidelines for adjusting the average installation factors to account for other-than-average installation conditions .[1]

The use of steam as a smoke suppressant can represent as much as 90% or more of the total direct annual costs.

Table 1.6: Self-Supporting Flare Costs

D (Diameter in Inches)	Hf (Hight in Feet)	Year 2000 (In Dollars)
12	30	$44,163
12	40	$47,367
12	50	$50,684
12	60	$54,112
12	70	$57,653
12	80	$61,306
12	90	$65,071
12	100	$68,948
24	30	$102,291
24	40	$107,138
24	50	$112,098
24	60	$117,169
24	70	$122,353
24	80	$127,649
24	90	$133,057
24	100	$138,578

Figure 1.5: Capital Costs of Self-supporting Flares for 12 in. and 24 in. Diameters

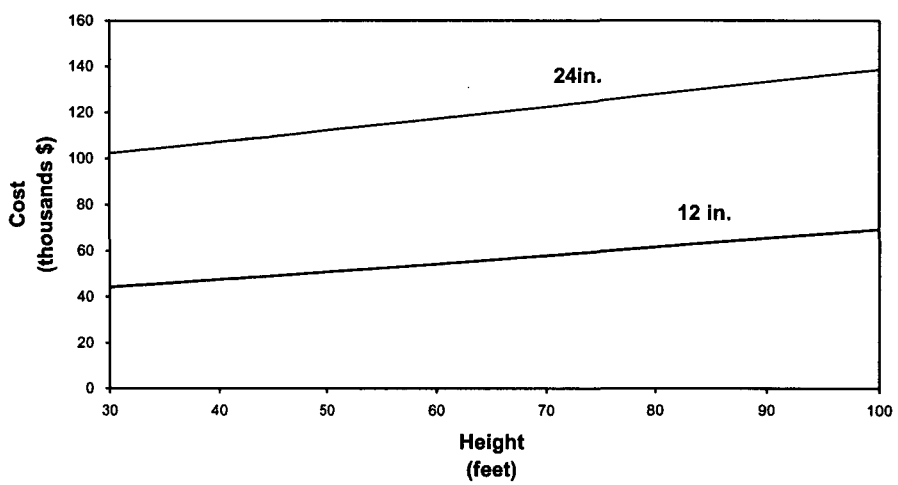


Table 1.7: Guy-Supported Flare Costs

Df (Diameter in Inches)	Hf (Height in Feet)	Year 2000 (In Dollars)
24	50	$112,104
24	100	$128,393
24	150	$145,787
24	200	$164,284
24	250	$183,887
24	300	$204,593
24	250	$183,887
24	400	$249,320
24	450	$273,341
48	50	$295,001
48	100	$321,081
48	150	$348,265
48	200	$376,554
48	250	$405,947
48	300	$436,445
48	250	$405,947
48	300	$436,445
48	250	$405,947
48	400	$500,754
48	450	$534,566

Figure 1.6: Captial Costs of Guy-Supported Flares for 24 in. and 48 in. Diameters

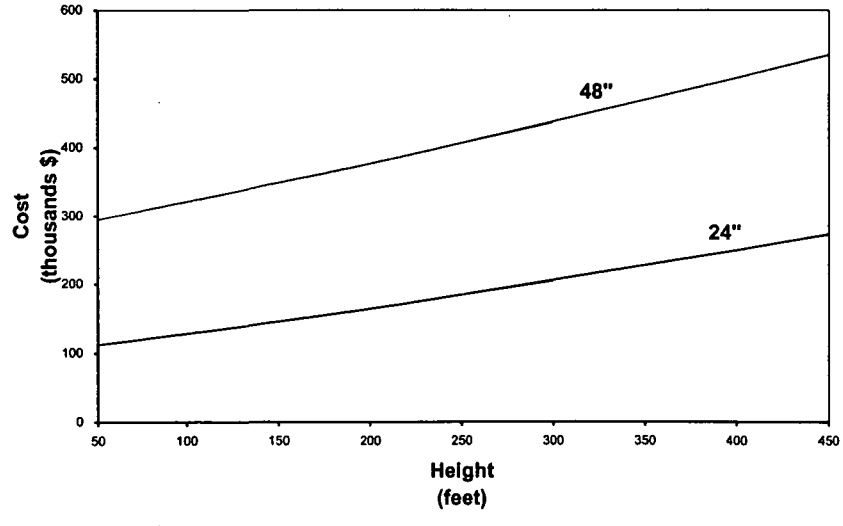
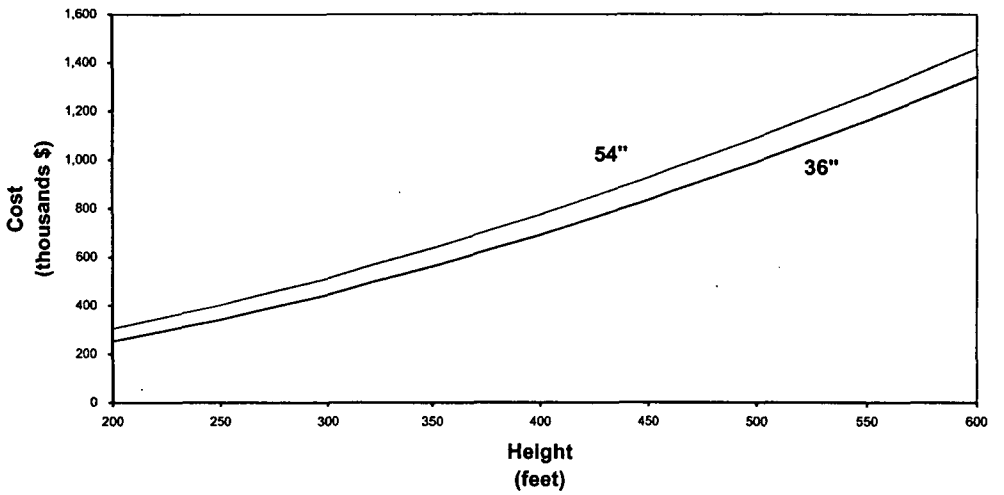


Table 1.8: Derrick Supported Flare Costs

Df (Diameter in Inches)	Hf (Height in Feet)	Year 2000 (In Dollars)
36	200	$252,325
36	250	$341,430
36	300	$443,982
36	350	$559,983
36	400	$689,431
36	450	$832,328
36	500	$988,672
36	550	$1,158,465
36	600	$1,341,705
54	200	$303,910
54	250	$401,044
54	300	$511,625
54	350	$635,655
54	400	$773,133
54	450	$924,059
54	500	$1,088,433
54	550	$1,266,255
54	600	$1,457,525

Figure 1.7: Captial Costs of Derrick-Supported Flares for 36 in. and 54 in. Diameters



1.5 Estimating Total Annual Costs

The total annual cost, TAC, is the sum of the direct and indirect annual costs. The bases used in calculating annual cost factors are given in Table 1.2

1.5.1 Direct Annual Costs

Direct annual costs include labor (operating and supervisory), maintenance (labor and materials), natural gas, steam, and electricity. Unless the flare is to be dedicated to one vent stream and specific on-line operating factors are known, costs should be calculated based on a continuous operation of 8,760 hr/yr and expressed on an annual basis. Flares serving multiple process units typically run continuously for several years between maintenance shutdowns.

Operating labor is estimated at 630 hours annually.[3] A completely manual system could easily require 1,000 hours. A standard supervision ratio of 0.15 should be assumed. Maintenance labor is estimated at 0.5 hours per 8-hour shift. Maintenance materials costs are assumed to equal maintenance labor costs. Flare utility costs include natural gas, steam, and electricity.

Flare systems can use natural gas in three ways: in pilot burners that fire natural gas, in combusting low Btu vent streams that require natural gas as auxiliary fuel, and as purge gas. The total natural gas cost, C_f to operate a flare system includes pilot, C_{pi}, auxiliary fuel, C_a, and purge costs, C_{pu}:

$$C_f \left(\frac{\$}{yr}\right) = C_{pi} + C_a + C_{pu} \tag{1.25}$$

where, C_{pi} is equal to the annual volume of pilot gas, F_{pi}, multiplied by the cost per scf, $Cost_{fuel}$:

$$C_{pi} \left(\frac{\$}{\text{yr}}\right) = \left(F_{pi} \frac{\text{scf}}{\text{yr}}\right)\left(Cost_{fuel} \frac{\$}{\text{scf}}\right) \tag{1.26}$$

C_a and C_{pu} are similarly calculated.

Steam cost (C_s) to eliminate smoking is equal to the annual steam consumption, multiplied by the cost per lb, $Cost_{Stream}$:

$$C_s \left(\frac{\$}{\text{yr}}\right) = \left(8{,}760 \frac{\text{hr}}{\text{yr}}\right)\left(S \frac{\text{lb}}{\text{hr}}\right)\left(Cost_{steam} \frac{\text{lb}}{\text{hr}}\right) \tag{1.27}$$

1.5.2 Indirect Annual Costs

The indirect (fixed) annual costs include overhead, capital recovery, administrative (G & A) charges, property taxes, and insurance. Suggested indirect annual cost factors are presented in Table 1.9.

Overhead is calculated as 60% of the total labor (operating, maintenance, and supervisory) and maintenance material costs. Overhead cost is discussed in Section 1 of this Manual.

Table 1.9: Capital Cost Factors for Flare Systems

Cost Item	Factor
Direct Costs	
Purchased equipment costs	
Flare system, EC	As estimated, A
Instrumentation	0.10 A
Sales taxes	0.03 A
Freight	0.05 A
Purchased equipment cost, PEC	B = 1.18 A
Direct installation costs	
Foundations & supports	0.12 B
Handling & erection	0.40 B
Electrical	0.01 B
Piping	0.02 B
Insulation	0.01 B
Painting	0.01 B
Direct installation costs	0.57 B
Site preparation	
As required, SP	
Buildings	
As required, Bldg.	
Total Direct Costs, DC	1.57 B + SP + Bldg.
Indirect Annual Costs, DC	
Engineering	0.10 B
Construction and Field expenses	0.10 B
Contractor fees	0 B
Start-up	0.01 B
Performance test	0.01 B
Contingencies	0.03 B
Total Indirect Costs, IC	0.35 B
Total Capital Investment = DC + IC	1.92 B + SP + Bldg.

The system capital recovery cost, CRC, is based on an estimated 15-year equipment life. (See Section 1 of this Manual for a thorough discussion of the capital recovery cost and the variables that determine it.) For a 15-year life and an interest rate of 7%, the capital recovery factor is 0.1098. The system capital recovery cost is the product of the system capital recovery factor, CRF, and the total capital investment, TCI, or:

$$CRC\left(\frac{\$}{yr}\right) = CRF \times TCI = 0.1098 \times TCI \qquad (1.28)$$

As shown in Table 1.10, G & A, taxes, and insurance can be estimated at 2%, 1%, and 1% of the total capital investment, TCI, respectively.

Table 1.10: Suggested Annual Cost Factors for Flare Systems

Cost Item	Factor
Direct Annual Costs, DC	
Operating labor{3}	
Operator	630 man-hours/year
Supervisor	15% of operator
Operating materials	-
Maintenance	
Labor	½ hour per shift
Material	100% of maintenance labor
Utilities	
Electricity	All utilities equal to:
Purge gas	(Consumption rate) x
Pilot gas	(Hours/yr) x (unit cost)
Auxiliary fuel	
Steam	
Indirect Annual Costs, IC	
Overhead	60% of total labor and material costs
Administrative charges	2% of Total Capital Investment
Property tax	1% of Total Capital Investment
Insurance	1% of Total Capital Investment
Capital recovery[a]	0.1315 x Total Capital Investment
Total Annual Cost	Sum of Direct and Indirect Annual Costs

[a]See Chapter 2.

1.6 Example Problem

The example problem described in this section shows how to apply the flare sizing and costing procedures to the control of a vent stream associated with the distillation manufacturing of methanol.

1.6.1 Required Information for Design

The first step in the design procedure is to determine the specifications of the vent gas to be processed. The minimum information required to size a flare system for estimating costs are the vent stream:

Volumetric or mass flow rate
Heating value or chemical composition
Temperature
System pressure
Vapor and liquid densities

In addition the following are needed to calculate direct annual costs.

Labor costs
Fuel costs
Steam costs

Vent stream parameters and cost data to be used in this example problem are listed in Table 1.11.

1.6.2 Capital Equipment

The first objective is to properly size a steam-assisted flare system to effectively destroy 98% of the VOC (methanol) in the vent gas stream. Using the vent stream parameters and the design procedures outlined in Section 1.3, flare and knock-out drum heights and diameters can be determined. Once equipment has been specified, the capital costs can be determined from equations presented in Section 1.4.1.

Table 1.11: Example Problem Data

Vent Stream Parameters	
Flow rate	63.4 acfm[a]
	399.3 lb/hr
Heat content	449 Btu/scf[b]
System pressure	10 psig[c]
Temperature	90 °F
Liquid density[17]	49.60 lb/ft^{3}[d]
Vapor density[17]	0.08446 lb/ft^{3}[d]
Cost Data (March 1990)[18,19]	
Operating hours	8,760 hrs/yr
Natural gas	3.03 $/1000 scf
Steam	4.65 $/1000 lbs
Operating labor	15.64 $/hr
Maintenance labor	17.21 $/hr

[a]Measured at *flare tip*. Flow rate has been adjusted to account for drop in pressure from 10 psig at source to 1 psig at flare tip.
[b]Standard conditions: 77°F, 1 atmosphere.
[c]Pressure at source (gas collection point). Pressure at flare tip is lower: 1 psig.
[d]Measured at standard conditions.

1.6.2.1 Equipment Design

The first step in flare sizing is determining the appropriate flare tip diameter. Knowing the net (lower) heating value of the vent stream, the maximum allowed velocity can be calculated from the Federal Register requirements. Since the heating value is in the range of 300 to 1,000 Btu/scf. the maximum velocity, V_{max}, is determined by Equation 1.1.

$$\log_{10} V_{max} = \frac{449 \frac{Btu}{scf} + 1{,}214}{852} = 1.95$$

$$V_{max} = 89.5 \frac{ft}{sec}$$

Because the stream heating value is above 300 Btu/scf, no auxiliary fuel is required. Hence, Q_{tot} equals the vent stream flow rate. Based on Q_{tot} and V_{max}, the flare tip diameter can be calculated using Equation 1.5.

$$D_{min} = 1.95 \sqrt{\frac{63.4\ scfm}{89.5\left(\frac{ft}{sec}\right)}} = 1.64\ in$$

The next largest commercially available standard size of 2 inches should be selected for D.

The next parameter to determine is the required height of the flare stack. The heat release from the flare is calculated using Equation 1.7.

$$R\ \frac{Btu}{hr} = W\ \frac{lb}{hr}\ B_v\ \frac{Btu}{lb}$$

First the heat of combustion, or heating value, must be converted from Btu/scf to Btu/lb. The vapor density of the vent stream at standard temperature and pressure is 0.08446 lb/scf.

So,

$$B_v = \frac{449\left(\frac{Btu}{scf}\right)}{0.08446\left(\frac{lb}{scf}\right)} = 5316\left(\frac{Btu}{lb}\right)$$

and,

$$R = \left(399.3\frac{\text{lb}}{\text{hr}}\right)\left(5{,}316\frac{\text{Btu}}{\text{lb}}\right) = 2{,}123{,}000\frac{\text{Btu}}{\text{hr}}$$

Substituting R and appropriate values for other variables into Equation 1.6:

$$L^2 = \frac{(1)\ (0.2)\left(2{,}123{,}000\ \frac{Btu}{hr}\right)}{4\pi\left(500\ \frac{Btu}{hr\text{-}ft^2}\right)} = 68\ ft^2$$

Resulting in:

$$L = 8.2\text{ft}$$

Assuming the smallest commercially available flare is 30 feet, the flare height is set to this value, L = 30 ft.

Next the knock-out drum must be sized. Assuming a design vapor velocity factor, G, of 0.20, and substituting the vapor and liquid densities of methanol into Equation 1.11 yields a maximum velocity of:

$$U = 0.20\sqrt{\frac{49.60 - 0.08446}{0.08446}} = 4.84\frac{\text{ft}}{\text{sec}}$$

Given a vent gas flow rate of 63.4 scfm, the minimum vessel cross-sectional, diameter is calculated by Equation 1.12:

$$A = \frac{63.4\text{acfm}}{(60)\frac{\text{sec}}{\text{min}}\,(4.84)\frac{\text{ft}}{\text{sec}}} = 0.218\ \text{ft}^2$$

This results in a minimum vessel diameter of:

$$d_{min} = 12\frac{in}{ft}\sqrt{\frac{4}{\pi}(0.218\,ft^2)} = 6.3\ in$$

The selected diameter, d, rounded to the next largest 6 inches is 12 inches. Using the rule of the height to diameter ratio of three gives a vessel height of 36 inches, or 3 feet.

1.6.2.2 Equipment Costs

Once the required flare tip diameter and stack height have been determined the equipment costs can be calculated. Since the height is 30 feet, the flare will be self-supporting. The costs are determined from Equation 1.16.

$$C_F = [78.0 + 9.14\ (2\ \text{inches}) + 0.749\ (30\ \text{ft})\ ]^2$$
$$= \$14{,}100$$

Knock-out drum costs are determined using Equation 7.21, where t is determined from the ranges Presented in Section 7.3.7. Substituting 0.25 for t:

$$C_K = 14.2 \left[(12) (0.25) (36 + 0.812 (12)) \right]^{0.737} = \$530$$

Transport piping costs are determined using Equation 1.19.

$$C_p = 127 (2)^{1.21} = \$290$$

The total auxiliary equipment cost is the sum of the knock-out drum and transport piping costs, or: $530 + $290 = $820.

The total capital investment is calculated using the factors given in Table 1.9. The calculations are shown in Table 1.12. Therefore:

$$PEC = 1.18 \times (14{,}920) = \$17{,}610$$

And:

$$TCI = 1.92 \times (17{,}610) = \$33{,}800$$

1.6.3 Operating Requirements

Operating labor is estimated at 630 hours annually with supervisory labor at 15% of this amount. Maintenance labor is estimated at 1/2 hour per shift. Maintenance material costs are assumed to be equal to maintenance labor costs.

As stated in Table 1.11, since the heat content of the example stream is above 300 Btu/scf no auxiliary fuel is needed. Natural gas is required, however, for purge and pilot gas. Purge gas requirements are calculated from Equation 1.8.

$$F_{pu} = 6.88 (2\ in)^2 = 27.5 \frac{Mscf}{yr}$$

Table 1.12: Capital Cost for Flare Systems - Example Problem

Cost Item	Cost
Diret Costs	
Purchased Equipment Costs	$14,100
Flare system (Self-supporting)	820
Auxiliary Equipment*	$14,920
Sum = A	
Instrumentation	1,490
Sales taxes	450
Freight	750
Purchased equipment cost, PEC	$17,610
Direct Installation Costs	
Foundations & Supports	2,110
Handling & Erection	7,040
Electrical	180
Piping	350
Insulation	180
Painting	180
Direct Installation Costs	$10,040
Site Preparation	
Facilities and Buildings	
Total Direct Costs	**$27,650**
Indirect Annual Costs, DC	
Engineering	1,760
Construction and Field Expenses	1,760
Contractor Fees	1,760
Start-Up	180
Performance Test	180
Contingencies	530
Total Indirect Costs, IC	$6,170
Total Capital Investment = DC + IC	**$33,800**

* Asumed to be 6% of the flare system cost. For more information on the costing of auxiliary equipment, refer to Section 2 of the Manual.

Since the flare tip diameter is less than 10 inches, pilot gas requirements are based on one pilot burner, (see Section 1.3.5) and are calculated by Equation 1.9.

$$F_{pu} = 6.88\ (2\ in)^2 = 27.5\ \frac{Mscf}{yr}$$

When N = 1,

$$F_{pi} = 613\ (1) = 613\ \frac{Mscf}{yr}$$

Steam requirements are calculated from Equation 1.10. Inserting the methanol mass flow rate of 399.3 lb/hr yields:

$$S = \left(0.4\ \frac{lb\ steam}{lb\ flare\ gas}\right)\left(8760\frac{hr}{yr}\right)\left(399.3\frac{lb}{hr}\right) = 1{,}400\ \frac{lb}{hr}$$

1.6.4 Total Annual Costs

The sum of the direct and indirect annual costs yields a total annual cost of $61,800. Table 1.13 shows the calculations of the direct and indirect annual costs for the flare system as calculated from the factors in Table 1.10. Direct costs include labor, materials, and utilities. Indirect costs are the fixed costs allocated to the project, including capital recovery costs and such costs as overhead, insurance, taxes, and administrative charges.

Electrical costs of a mover system (fan, blower, compressor) would have to be included if the vent stream pressure was not sufficient to overcome the flare system pressure drop. In this example case, the pressure is assumed to be adequate.

Table 1.13: Annual Costs for Flare System Example problem

Cost Item	Calculations	Cost
Direct Annual Costs, DC		
Operating Labor		
Operator	$\frac{630\ h}{year} \times \frac{\$15.64}{h}$	$ 9,850
Supervisor	15% of operator = 0.15 x 9,850	1,480
Operating materials		
Maintenance		
Labor	$\frac{0.5\ h}{shift} \times \frac{shift}{8\ h} \times \frac{8,760h}{yr} \times \frac{\$17.21}{h}$	9,420
Maintenance Material	100% of maintenance labor	9,420
Utilities		
Electricity		
Purge gas Inserting the methanol mass flow rate of 399.3 lb/hr yields:		
	$\frac{27.5\ Mscf}{yr} \times \frac{\$3.03}{Mscf}$	80
Pilot gas	$\frac{613\ Mscf}{yr} \times \frac{\$3.03}{Mscf}$	1,860
Steam	$\frac{1,400 \times 10^3 lb}{yr} \times \frac{\$4.65}{10^3 lb}$	6,510
Total DC (rounded)		$38,600
Indirect Annual Costs, IC		
Overhead	60% of total labor and material costs = 0.6(9,850 + 1,480 + 9,420 + 9,420)	18,100
Administrative charges	2% of Total Capital Investment = 0.02 ($33,800)	680
Property tax	1% of Total Capital Investment = 0.01 ($33,800)	340
Insurance	1% of Total Capital Investment = 0.01 ($33,800)	340
Capital recovery[a]	0.1098 x $33,800	3,710
Total IC (rounded)		23,200
Total Annual Cost (rounded)		$61,800

[a]The capital recovery cost factor, CRF, is a function of the flare equipment life and the opportunity cost of the capital (i.e. interest rate). For example, for a 15 year equipment life and 7% interest rate, CRF = 0.1098.

1.7 Acknowledgments

The authors gratefully acknowledge the following companies for contributing data to this chapter:

- Flaregas Corporation (Spring Valley, NY)
- John Zink Company (Tulsa, OK)
- Kaldair Incorporated (Houston, TX)
- NAO Incorporated (Philadelphia, PA)
- Peabody Engineering Corporation (Stamford, CT)
- Piedmont HUB, Incorporated (Raleigh, NC)

References

[1] *Guide for Pressure-Relieving and Depressurizing Systems*, Refining Department, API Recommended Practice 521, Second Edition, September 1982.

[2] Kalcevic, V. (IT Enviroscience), "Control Device Evaluation Flares and the Use of Emissions as Fuels", *Organic Chemical Manufacturing Volume 4; Combustion Control Devices*, U.S. Environmental Protection Agency, Research Triangle Park, NC, Publication no. EPA-450/3-80-026, December 1980, Report 4.

[3] *Reactor Processes in Synthetic Organic Chemical Manufacturing Industry-Background Information for Proposed Standards*, U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards, Research Triangle Park, NC, Preliminary Draft, EPA 450/3-90-016a, June 1990.

[4] Letter from J. Keith McCartney (John Zink Co., Tulsa, OK) to William M. Vatavuk (U.S. Environmental Protection Agency, Research Triangle Park, NC), November 19, 1990.

[5] Letter from David Shore (Flaregas Corp., Spring Valley, NY) to William M. Vatavuk (U.S. Environmental Protection Agency, Research Triangle Park, NC), October 3, 1990.

[6] Letter from Pete Tkatschenko (NAO, Inc., Philadelphia, PA) to Diana Stone (Radian, Research Triangle Park, NC), May 2, 1990.

[7] Letter to Gary Tyler (Kaldair, Inc., Houston, TX) to Diana Stone (Radian, Research Triangle Park, NC), April 10, 1990.

[8] Letter from Zahir Bozai (Peabody Engineering Corp., Stamford, CT) to Diana Stone (Radian, Research Triangle Park, NC), May 7, 1990.

[9] Letter from James Parker (John Zink Co., Tulsa, OK) to Diana Stone (Radian, Research Triangle Park, NC), April 17, 1990.

[10] Letter from Nick Sanderson (Flaregas Corp., Spring Valley, NY) to Diana Stone (Radian, Research Triangle Park, NC), May 2, 1990.

[11] Wu, F.H., "Drum Separator Design, A New Approach," *Chemical Engineering*, April 2, 1984, pp. 74-81.

[12] Mulet, A., "Estimate Costs of Pressure Vessels Via Correlations," *Chemical Engineering*, October 5, 1981, pp. 145-150.

[13] *Process Plant Construction Estimating Standards*, Richardson Engineering Services, Inc., Volume 4, 1988 Edition.

[14] Peters, Max S. and Klaus D. Timmerhaus, Plant Design and *Economics for Chemical Engineers*, Third Edition, McGraw-Hill, 1980.

[15] Cost Information from Piedmont HUB, Incorporated, Raleigh, NC. August 1990.

[16] Vatavuk, W.M., and R. Neveril,"Estimating Costs of Air Pollution Control Systems, Part II: Factors for Estimating Capital and Operating Costs,: *Chemical Engineering*, November 3, 1980, pp. 157-162.

[17] *Handbook of Chemistry and Physics*, 55th Edition, CRC Press, 1974-1975.

[18] Green, G.P. and Epstein, R.K., Employment and Earnings, Department of Labor, Bureau of Labor Statistics, Volume 37, No.4, April 1990.

[19] *Monthly Energy Review*, Energy Information Administration, Office of Energy Markets and End Use, U.S. Department of Energy, DOE-EIA-0035(90/12), February 1990.

EPA/452/B-02-001

Section 3

VOC Controls

EPA/452/B-02-001

Section 3.2

VOC Destruction Controls

EPA/452/B-02-001

Chapter 2

Incinerators

William M. Vatavuk
Innovative Strategies and Economics Group, OAQPS
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

Donald R. van der Vaart & James J. Spivey
Research Triangle Institute
Research Triangle Park, NC 27709

September 2000

Contents

2.1 Introduction 2.-3

2.2 Process Description 2-3
2.2.1 Thermal Incinerators 2-6
2.2.2 Catalytic Incinerators 2-10
2.2.3 Other Considerations: Packaged versus Field-Erected Units, Auxiliary Equipment 2-13
2.2.4 Technology Comparison 2-15

2.3 General Treatment of Material and Energy Balances 2-16

2.4 Design Procedures 2-18
2.4.1 Steps Common to Thermal and Catalytic Units 2-18
2.4.2 Steps Specific to Thermal Units 2-23
2.4.3 Steps Specific to Catalytic Units 2-29

2.5 Cost Analysis 2-35
2.5.1 Estimating Total Capital Investment 2-35
2.5.1.1 Equipment Costs, EC 2-35
2.5.1.2 Installation Costs 2-40
2.5.2 Estimating Total Annual Cost 2-41
2.5.2.1 Direct Annual Costs 2-41
2.5.2.2 Indirect Annual Costs 2-47
2.5.3 Cost Comparison for Example Case 2-47

2.6 Acknowledgements 2-48

References 2-49

2.1 Introduction

Incineration, like carbon adsorption, is one of the best known methods of industrial gas waste disposal. Unlike carbon adsorption, however, incineration is an ultimate disposal method in that the objectionable combustible compounds in the waste gas are converted rather than collected. On the other hand, carbon adsorption allows recovery of organic compounds which may have more value as chemicals than just their heating value. A major advantage of incineration is that virtually any gaseous organic stream can be incinerated safely and cleanly, provided proper engineering design is used.

The particular application of both thermal and catalytic incineration to gaseous waste streams containing volatile organic compounds (VOCs) is discussed here. The U.S. Environmental Protection Agency defines any organic compound to be a VOC unless it is specifically determined not to be a VOC. Indeed, a number of organics (e.g., methane) are specified as not being VOCs. Although both VOC and non-VOC organic compounds are combustible and are therefore important in the design of the incinerator, this distinction is important since it is the control of VOCs that is regulated.

2.2 Process Description

Seldom is the waste stream to be combusted a single organic compound. Rather, it is common to have a complex mixture of organic compounds. This mixture is typically analyzed for carbon, hydrogen, oxygen, and other elements; and an empirical formula is developed which represents the mixture. Combustion of such a mixture of organic compounds containing carbon, hydrogen, and oxygen is described by the overall exothermic reaction:

$$C_x H_y O_z + \left[x + \frac{y}{4} - \frac{z}{2}\right] O_2 \Rightarrow xCO_2 + \frac{y}{2} H_2O \qquad (1.2)$$

The complete combustion products CO_2 and H_2O are relatively innocuous, making incineration an attractive waste disposal method. When chlorinated sulfur-containing compounds are present in the mixture, the products of complete combustion include the acid components HCl or SO_2, respectively, in addition to H_2O and CO_2. In general, these streams would require removal of the acid components by a scrubber unit, which could greatly affect the cost of the incineration system. (The sizing and costing of these scrubbers is covered in the "Wet Scrubbers" chapter of this Manual.)

The heart of an incinerator system is a combustion chamber in which the VOC-containing waste stream is burned. Since the inlet waste gas stream temperature is generally much lower than

that required for combustion, energy must be supplied to the incinerator to raise the waste gas temperature. Seldom, however, is the energy released by the combustion of the total organics (VOCs and others) in the waste gas stream sufficient to raise its own temperature to the desired levels, so that auxiliary fuel (e.g., natural gas) must be added.

The combustion of the waste gases may be accomplished in a thermal incinerator or in a catalytic incinerator. In the catalytic incinerator a catalyst is used to increase the rate of the combustion reaction, allowing the combustion to occur at lower temperatures. Because the catalytic process operates at a lower temperature than the thermal process, less auxiliary fuel may be required in the catalytic process to preheat the waste gas.

Auxiliary fuel requirements may also be decreased, and energy efficiency improved, by providing heat exchange between selected inlet streams and the effluent stream. The effluent stream containing the products of combustion, along with any inerts that may have been present in or added to the inlet streams, can be used to preheat the incoming waste stream, auxiliary air, or both via a "primary", or recuperative, heat exchanger. It is useful to define the fractional energy recovery by the preheater or primary heat exchanger as follows:

$$\text{Fractional Energy Recovery} = \frac{\text{Energy actually recovered flue gas}}{\text{Maximum energy recoverable if flue gas approaches lowest temperature available to heat exchanger}} \qquad (2.2)$$

The energy actually recovered, the numerator of Equation 2.2, is the increase in sensible heat of the gas, e.g., waste gas or waste gas plus dilution air, being heated. The maximum energy recoverable would be the decrease in sensible heat of the flue gas, if it were cooled to the temperature of the incoming waste gas. While this maximum energy recovery would be attained only with a very large heat exchanger, the concept of fractional energy recovery is useful in expressing the extent of the improvement in energy efficiency using a "primary" heat exchanger.

Energy efficiency can be further improved by placing another ("secondary") exchanger downstream of the primary exchanger to recover additional energy from the effluent stream (e.g., to generate low pressure process steam or hot water). However, secondary energy recovery is generally not used, unless there is a specific on site use for the steam or hot water.

The majority of industrial gases that contain VOCs are dilute mixtures of combustible gases in air. In some applications, such as air oxidation processes, the waste gas stream is very deficient in oxygen. Depending on the oxygen content of the waste stream, auxiliary air may be required to combust the total organic content of the waste gas as well as any auxiliary fuel that has been used.

The concentration of combustible gas in the waste gas stream plays an integral role in the design and operation of an incinerator. From a cost standpoint, the amount of air in excess of the stoichiometric amounts should be minimized. For safety reasons, however, any mixture within the flammability limits, on either the fuel-rich or fuel-lean side of the stoichiometric mixture, presents an unacceptable fire hazard as a feed stream to the incinerator. The lower, or fuel-lean, explosive limit (LEL) of a given organic compound defines the minimum concentration of that compound in air that can produce more energy than is needed to raise its own temperature to the ignition point (e.g., ignite). Similarly, the upper, or fuel-rich, explosive limit (UEL) represents the highest concentration of the organic in air that is ignitable. In the latter case, air is limiting the reaction. Both the LEL and the UEL are measured at ambient conditions. Empirically, it has been found that mixtures of hydrocarbons in air at their LEL have a heating value of approximately 50 Btu/scf.

Since the majority of industrial waste gases that contain VOCs are dilute mixtures of combustible gases in air, their heating value is low and their oxygen content exceeds that required to combust both the waste organics (VOCs and others) and the auxiliary fuel. If a waste gas above 50 percent LEL (about 25 Btu/scf) is encountered, it must be diluted to satisfy fire insurance regulations. Generally, the streams are brought to below 25 percent LEL, although concentrations from 25 percent to 50 percent are permitted provided the waste stream is continuously monitored by LEL monitors. Because air is the usual diluent gas, care must be taken with preheating the diluted stream so that it remains below about 1200°F. (See discussion below on preheating.) A table showing LEL, UEL, and heats of combustion for selected organic compounds is given in Appendix A.

The goal of any incineration system is to control the amount of VOCs released to the environment. Performance of a control device such as an incinerator can be described by a control efficiency in percent (%), defined according to the following equation:

$$Control\ Eff. = \left[\frac{Inlet\ mass\ rate\ VOC - Outlet\ mass\ rate\ VOC}{Inlet\ mass\ rate\ VOC}\right] \times 100 \qquad (2.3)$$

It is important to note, however, that incomplete combustion of the inlet VOCs could result in the formation of other VOCs not originally present. For example, the incomplete oxidation of dichloroethane can yield vinyl chloride. Both of these compounds are VOCs. The definition given in Equation 2.3 would still be meaningful, however, as long as the newly formed VOC (e.g., vinyl chloride) is detected. This situation necessitates the complete chemical analysis of the inlet and outlet gas streams to confirm compliance with State and Federal regulations.

Performance of an incinerator can also be measured solely by the *outlet* VOC concentration, usually in parts per million by volume (ppmv).

There are a number of different incinerator designs. These designs can be broadly classified as thermal systems and catalytic systems. Thermal systems may be direct flame incinerators with no energy recovery, flame incinerators with a recuperative heat exchanger, or regenerative systems which operate in a cyclic mode to achieve high energy recovery. Catalytic systems include fixed-bed (packed-bed or monolith) systems and fluid-bed systems, both of which provide for energy recovery. The following sections discuss design aspects of these systems.

2.2.1 Thermal Incinerators

The heart of the thermal incinerator is a nozzle-stabilized flame maintained by a combination of auxiliary fuel, waste gas compounds, and supplemental air added when necessary (see Figure 2.1). Upon passing through the flame, the waste gas is heated from its inlet temperature (e.g., 100°F) to its ignition temperature. The ignition temperature varies for different compounds and is usually determined empirically. It is the temperature at which the combustion reaction rate (and consequently the energy production rate) exceeds the rate of heat losses, thereby raising the temperature of the gases to some higher value. Thus, any organic/air mixture will ignite if its temperature is raised to a sufficiently high level.

The organic-containing mixture ignites at some temperature between the preheat temperature and the reaction temperature. That is, ignition, as defined in this section, occurs at some point during the heating of a waste stream as it passes through the nozzle-stabilized flame regardless of its concentration. The mixture continues to react as it flows through the combustion chamber.

The required level of VOC control of the waste gas that must be achieved within the time that it spends in the thermal combustion chamber dictates the reactor temperature. The shorter the residence time, the higher the reactor temperature must be. The nominal residence time of the reacting waste gas in the combustion chamber is defined as the combustion chamber volume divided by the volumetric flow rate of the gas. Most thermal units are designed to provide no more than 1 second of residence time to the waste gas with typical temperatures of 1,200 to 2,000°F. Once the unit is designed and built, the residence time is not easily changed, so that the required reaction temperature becomes a function of the particular gaseous species and the desired level of control. Table 2.1 illustrates the variability in (theoretical) reactor temperatures that is required to destroy 99.99 percent of the inlet mass of various noxious compounds with excess air for a 1-second reactor residence time [1].

These temperatures cannot be calculated a priori, although incinerator vendors can provide guidelines based on their extensive experience. In practice, most streams are mixtures of compounds,

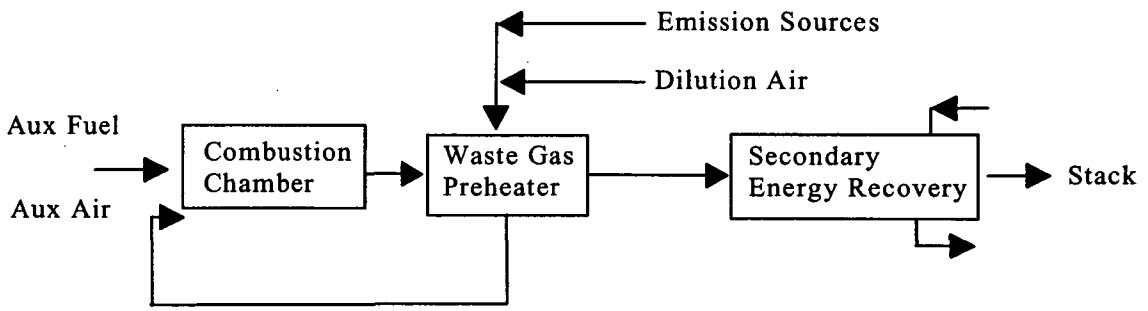


Figure 2.1: Thermal Incinerator - General Case

thereby further complicating the prediction of this temperature. Other studies [2,3,4], which are based on actual field test data, show that commercial incinerators should generally be run at 1600°F with a nominal residence time of 0.75 seconds to ensure 98% destruction of non-halogenated organics. In some States the reactor temperature and residence time of the unit are specified rather than attempting to measure actual levels of VOC control. The selected temperature must be maintained for the full, selected residence time for combustion to be complete.

These three studies also conclude that mixing is a critical factor in determining the destruction efficiency. Even though it cannot be measured, mixing is a factor of equal or even greater importance than other parameters, such as temperature. The most feasible and efficient way to improve the mixing in an incinerator is to adjust it after start-up. The 98% control level discussed in the previous paragraph presumes such an adjustment.

Ultimately, once the unit is built, it is the responsibility of the user to operate and maintain the incinerator to insure compliance with applicable regulations.

Direct Flame Incinerators

Many configurations of thermal incinerators exist with the same goal—to raise the VOC-containing stream to the desired reaction temperature and hold it there for the given reaction time

Table 2.1: Theoretical Reactor Temperatures Required for 99.99 Percent Destruction by Thermal Incineration for a 1-Second Residence Time [1]

Compound	Temperature, °F
acrylonitrile	1,344
allyl chloride	1,276
benzene	1,350
chlorobenzene	1,407
1,2-dichloroethane	1,368
methyl chloride	1,596
toluene	1,341
vinyl chloride	1,369

to achieve the required destruction efficiency. The simplest example of such a system is the direct flame incinerator. With reference to Figure 2.1, the direct flame incinerator is comprised only of the combustion chamber. The waste gas preheater and the secondary energy recovery heat exchanger are energy recovery devices and are not included as part of the direct flame incinerator.

Recuperative Incinerators

Recuperative incinerators have improved energy efficiency as a result of placing heat exchangers in the hot outlet gas streams. With reference to Figure 2.1, the recuperative incinerator is comprised of the combustion chamber, the waste gas preheater, and, if appropriate, the secondary, energy recovery heat exchanger.

Primary Energy Recovery (Preheating Inlet Streams) - Considerable fuel savings can be realized by using the exit (product) gas to preheat the incoming feed stream, combustion air, or both via a heat exchanger, as shown in Figure 2.1 in the so-called "recuperative" incinerator. These heat exchangers can recover up to 70% of the energy (enthalpy) in the product gas.

The two types of heat exchangers most commonly used are plate-to-plate and shell-and-tube. Plate-to-plate exchangers offer high efficiency energy recovery at lower cost than shell-and-tube designs. Also, because of their modular configuration, plate-to-plate units can be built to achieve a variety of efficiencies. But when gas temperatures exceed 1000°F, shell-and-tube exchangers usually have lower purchase costs than plate-to-plate designs. Moreover, shell-and-tube exchangers offer better long-term structural reliability than plate-to-plate units.[5] In any case, because most incinerators installed are packaged units, the design (and cost) of the recuperative heat exchangers have already been incorporated.

Most heat exchangers are not designed to withstand high temperatures, so that most of the energy needed to reach ignition is supplied by the combustion of fuel in the combustion chamber and only moderate preheat temperatures are sought in practice (<1200°F).

Secondary Energy Recovery (Additional Waste Energy Recovery) - It should be noted, however, that at least some of the energy added by auxiliary fuel in the traditional thermal units (but not recovered in preheating the feed stream) can still be recovered. Additional heat exchangers can be added to provide process heat in the form of low pressure steam or hot water for on-site application. Obviously, an in-plant use for such low level energy is needed to realize these savings.

The need for this higher level of energy recovery will be dependent upon the plant site. The additional heat exchanger is often provided by the incineration unit vendor. The cost of this additional heat exchanger may be estimated via standard heat exchanger correlations and should be added to the costs estimated using the cost correlations in this section.

Regenerative Incinerators

A distinction in thermal incinerators can now be made on the basis of this limitation in the preheat temperature. The traditional approach to energy recovery in the units (shown schematically in Figure 2.1) still requires a significant amount of auxiliary fuel to be burned in the combustion chamber when the waste gas heating values are too low to sustain the desired reaction temperature at the moderate preheat temperature employed. Additional savings can, under these conditions, be realized in units with more complete transfer of exit-stream energy. This is the concept behind the so-called excess-enthalpy or regenerable burner systems. These systems use direct contact heat exchangers constructed of a ceramic material that can tolerate the high temperatures needed to achieve ignition of the waste stream.

The operation of the regenerative system is illustrated in Figure 2.2. The inlet gas first passes through a hot ceramic bed thereby heating the stream (and cooling the bed) to its ignition temperature. If the desired temperature is not attainable, a small amount of auxiliary fuel is added in the combustion chamber. The hot gases then react (releasing energy) in the combustion chamber and while passing through another ceramic bed, thereby heating it to the combustion chamber outlet temperature. The process flows are then switched, now feeding the inlet stream to the hot bed. This cyclic process affords very high energy recovery (up to 95%).

The higher capital costs associated with these high-performance heat exchangers and combustion chambers may be offset by the increased auxiliary fuel savings to make such a system economical. The costs of these regenerative units will be given separately in the cost correlations presented in Section 2.4. Regenerative incinerators are not packaged units but are field-erected only. Accordingly, the costs given in Section 2.4 for regenerative units are for field-erected units.

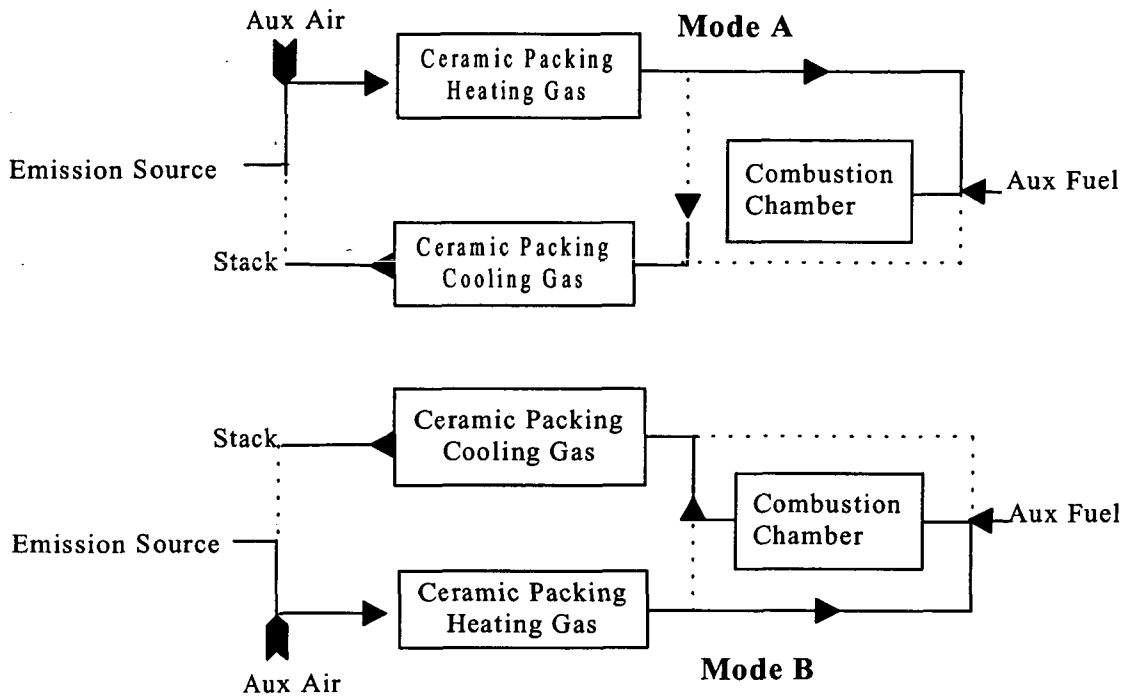


Figure 2.2: Regenerable-Type Thermal Incinerator

2.2.2 Catalytic Incinerators

Catalytic incinerators employ a bed of active material (catalyst) that facilitates the overall combustion reaction given in Equation 2.1. The catalyst has the effect of increasing the reaction rate, enabling conversion at lower reaction temperatures than in thermal incinerator units. Nevertheless, the waste stream must be preheated to a temperature sufficiently high (usually from 300 to 900°F) to initiate the oxidation reactions. The waste stream is preheated either directly in a preheater combustion chamber or indirectly by heat exchange with the incinerator's effluent or other process heat or both (Figure 2.3). The preheated gas stream is then passed over the catalyst bed. The chemical reaction (combustion) between the oxygen in the gas stream and the gaseous pollutants takes place at the catalyst surface. Catalytic incineration can, in principle, be used to destroy essentially any oxidizable compound in an air stream. However, there are practical limits to the types of compounds that can be oxidized due to the poisoning effect some species have on the catalyst. These limits are described below. In addition, most configurations require a low heating value of the inlet gas and a particulate content which is less than some small value.

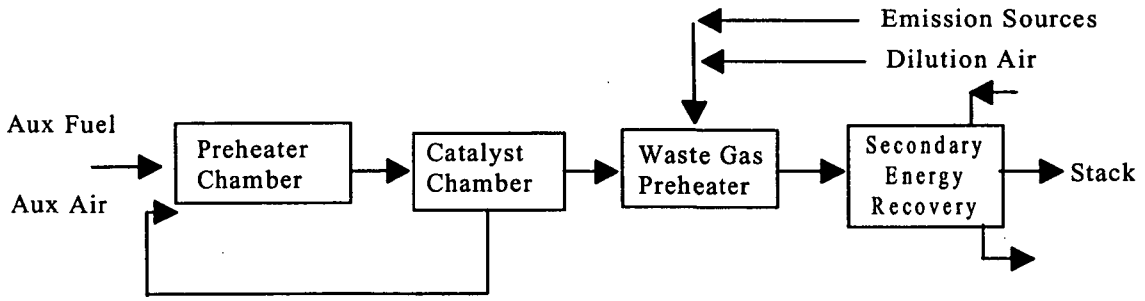


Figure 2.3: Catalytic Incinerator

Until recently, the use of catalytic oxidation for control of gaseous pollutants has really been restricted to organic compounds containing only carbon, hydrogen, and oxygen. Gases containing compounds with chlorine, sulfur, and other atoms that may deactivate the supported noble metal catalysts often used for VOC control were not suitably controlled by catalytic oxidation systems. Catalysts now exist, however, that are tolerant of such compounds. Most of these catalysts are single or mixed metal oxides, often supported by a mechanically strong carrier such as alumina. Perhaps most of the development of poison-tolerant catalysts has focused on the oxidation of chlorine-containing VOCs. These compounds are widely used as solvents and degreasers and are often the subject of concern in VOC control. Catalysts such as chromia/alumina [6,7], cobalt oxide [8], and copper oxide/manganese oxide [9] have been used for oxidation of gases containing chlorinated compounds. Platinum-based catalysts are active for oxidation of sulfur containing VOCs, although they are rapidly deactivated by the presence of chlorine. Compounds containing atoms such as lead, arsenic, and phosphorous should, in general, be considered poisons for most oxidation catalysts. Nevertheless, their concentration may be sufficiently low so that the rate of deactivation and therefore, the catalyst replacement costs, could be low enough to consider catalytic oxidation.

As was the case for thermal units, it is impossible to predict a priori the temperature and residence time (e.g., inverse space velocity) needed to obtain a given level of conversion of a VOC mixture in a catalytic oxidation system. For example, Table 2.2 from Pope *et al.* [3] shows the temperature needed for 80% conversion of a number of VOCs over two oxidation catalysts in a specific reactor design. This table shows that the temperature required for this level of conversion of different VOCs on a given catalyst and of the same VOC on different catalysts can vary significantly.

Particulate matter, including dissolved minerals in aerosols, can rapidly blind the pores of catalysts and deactivate them over time. Because essentially all the active surface of the catalyst is

contained in relatively small pores, the particulate matter need not be large to blind the catalyst. No general guidelines exist as to particulate concentration and particulate size that can be tolerated by catalysts because the pore size and volume of catalysts vary greatly.

The volumetric gas flow rate and the concentration of combustibles in the gas flowing to the catalytic incinerator should be constant for optimal operation. Large fluctuations in the flow rate will cause the conversion of the VOCs to fluctuate also. Changes in the concentration or type of organics in the gas stream can also affect the overall conversion of the VOC contaminants. These changes in flow rate, organics concentration, and chemical composition are generally the result of upsets in the manufacturing process generating the waste stream. It may be uneconomical to change the process for the sake of making the operation of the catalytic incinerator feasible. In such cases, thermal incinerators (discussed earlier in this chapter) or carbon adsorption (discussed in Section 3.1 of this Manual) should be evaluated as alternative control technology.

The method of contacting the VOC-containing stream with the catalyst serves to distinguish catalytic incineration systems. Both fixed-bed and fluid-bed systems are used.

Fixed-Bed Catalytic Incinerators

Fixed-bed catalytic incinerators may use a monolith catalyst or a packed-bed catalyst. Each of these is discussed below.

Monolith Catalyst Incinerators - The most widespread method of contacting the VOC-containing stream with the catalyst is the catalyst monolith. In this scheme the catalyst is a porous solid block containing parallel, non-intersecting channels aligned in the direction of the gas flow.

Table 2.2: Catalyst Temperatures Required for Oxidizing 80% of Inlet VOC to CO_2, °F for Two Catalyss

	Temperature, °F	
Compound	**CO_3O_4**	**Pt - Honeycomb**
acrolein	382	294
n-butanol	413	440
n-propylamine	460	489
toluene	476	373
n-butyric acid	517	451
1, 1, 1-trichloroethane	661	>661
dimethyl sulfide	-	512

Monoliths offer the advantages of minimal attrition due to thermal expansion/ contraction during startup/shutdown and low overall pressure drop.

Packed-Bed Catalytic Incinerators - A second contacting scheme is a simple packed-bed in which catalyst particles are supported either in a tube or in shallow trays through which the gases pass. The first scheme is not in widespread use due to its inherently high pressure drop, compared to a monolith, and the breaking of catalyst particles due to thermal expansion when the confined catalyst bed is heated/cooled during startup/shutdown. However, the tray type arrangement, where the catalyst is pelletized is used by several industries (e.g., heat-set web-offset printing). Pelletized catalyst is advantageous where large amounts of such contaminants as phosphorous or silicon compounds are present. [10]

Fluid-Bed Catalytic Incinerators

A third contacting pattern between the gas and catalyst is a fluid-bed. Fluid-beds have the advantage of very high mass transfer rates, although the overall pressure drop is somewhat higher than for a monolith. An additional advantage of fluid-beds is a high bed-side heat transfer as compared to a normal gas heat transfer coefficient. This higher heat transfer rate to heat transfer tubes immersed in the bed allows higher heat release rates per unit volume of gas processed and therefore may allow waste gas with higher heating values to be processed without exceeding maximum permissible temperatures in the catalyst bed. In these reactors the gas phase temperature rise from gas inlet to gas outlet is low, depending on the extent of heat transfer through imbedded heat transfer surfaces. The catalyst temperatures depend on the rate of reaction occurring at the catalyst surface and the rate of heat exchange between the catalyst and imbedded heat transfer surfaces.

As a general rule, fluid-bed systems are more tolerant of particulates in the gas stream than either fixed-bed or monolithic catalysts. This is due to the constant abrasion of the fluidized catalyst pellets, which helps remove these particulates from the exterior of the catalysts in a continuous manner. A disadvantage of a fluid-bed is the gradual loss of catalyst by attrition. Attrition-resistant catalysts have been developed to overcome this disadvantage.[76]

2.2.3 Other Considerations: Packaged versus Field-Erected Units, Auxiliary Equipment

Packaged vs. Field-Erected Units

With the exception of regenerative incinerators, the equipment cost correlations included in this chapter are for packaged units only. They are not valid for field-erected units. For regenerative incinerators, the correlations are valid for field-erected units only. Packaged units are units that have been shop fabricated and contain all elements necessary for operation, except for connection

to facilities at the site, e.g., utilities. The elements include the combustion chamber, preheater, instrumentation, fan, and the necessary structural steel, piping, and electrical equipment. This equipment is assembled and mounted on a "skid" to facilitate installation on a foundation at the plant site. Tie-in to the local emission source is not part of the packaged unit. Units are usually sized to handle flow rates of < 20,000 scfm, but can be built to accommodate flow rates up to 50,000 standard cubic feet per minute (scfm). The cost correlations in this chapter are valid to 50,000 scfm for packaged units, except for fluid-bed units which are valid to 25,000 scfm.

Conversely, field-erected units may be built to any desired size. The combustion chamber, preheater, and other equipment items are designed and fabricated individually, and assembled at the site. However, both the equipment and installation costs of field-erected units are typically higher than those for equivalent-sized packaged units because the factors that improve efficiency of shop-fabrication, such as uniform working environment, availability of tools and equipment, and more efficient work scheduling, are generally not available in the field.

Acid Gas Scrubbers

The final outlet stream of any incineration system may contain certain pollutants that must be removed. The combustion of sulfur-containing compounds results in SO_2, while chlorinated compounds yield Cl_2 and HCl in the product stream. These acid gases must be removed from the gas stream if they are present at significant concentrations (regulations for limits on these gases vary from state to state). This removal can be effected in, for instance, a packed-bed gas absorber (vertical scrubber) in which the flue gas is contacted with a caustic scrubbing liquid. For fluid-bed catalytic reactors, venturi scrubbers are often used because they provide for particulate removal as well as acid gas scrubbing. In most cases adding a scrubber or absorber significantly increases the cost of the incineration unit, sometimes by a factor of two. Costing of absorbers is discussed in the "Gas Absorbers" chapter of this Manual.

If chlorinated VOCs are present in the waste gas, heat exchangers may require special materials of construction. This added expense is not included in the costing procedures outlined in this section.

Heat Exchangers (Preheaters and Other Waste Energy Recovery Units)

For the thermal and catalytic units having some degree of energy recovery, the cost of the primary heat exchanger is included in the cost, and its design is usually done by the incineration unit vendor. The cost correlations presented in this chapter include units both with and without energy recovery. Secondary energy recovery, if desired, requires an additional heat exchanger, which is also often provided by the incineration unit vendor. Costing procedures for secondary energy recovery are not included in this section.

Other Auxiliary Equipment

Additional auxiliary equipment such as hoods, ductwork, precoolers, cyclones, fans, motors, and stacks are addressed separately in other sections of this *Manual.*

2.2.4 Technology Comparison

Both the thermal and catalytic incineration systems are designed to provide VOC control through combustion at a level in compliance with applicable state and federal requirements. Given the wide range of options available, however, it is obvious that not every incinerator will fulfill these requirements at the same cost. This section presents a first step toward deciding how best to deal with VOC emission abatement using incinerators considering some qualitative factors pertinent to the types of incinerators described in this chapter. It is the intent of the remainder of this section to provide a method by which the cost of VOC control for a particular application can be calculated.

A summary of the principal types of incinerators is presented in Table 2.3. From the earlier discussions, the following factors relating to the presence of contaminants should be considered by potential users [12]:

- The fouling of the catalyst in a catalytic system is a possibility. Poisons to the system include heavy metals, phosphorous, sulfur and most halogens, although catalysts have been developed that are chlorine resistant.

- The possibility of process upsets that could release any of the above poisons or cause fluctuations in the heating value to the incinerator would favor a thermal system.

- Except for the No.2 grade, fuel oil should not be considered as auxiliary fuel to a catalytic system due to the sulfur and vanadium it may contain.[10]

All of the above factors would serve to increase the operating expense of a catalytic unit through replacement costs of the catalyst. An additional factor relates to relative energy efficiency of the various types of incinerators:

- Thermal units generally require more auxiliary fuel than catalytic units and operate at temperatures that are roughly 1000°F higher. This difference in fuel requirement increases as the heating value of the waste stream decreases.

In general, a trade-off exists between the higher capital costs of catalytic incinerators and the higher operating costs of thermal incinerators. This difference will be illustrated by a design example presented in Section 2.2.4 which treats both technologies.

Table 2.3: Principal VOC Incineration Technologies

Thermal Systems	Catalytic Systems
- Direct Flame Incinerator	- Fixed-Bed
- Recuperative Incinerator (Direct Flame with Recuperative Heat Exchanger)	- Monolith - Packed-Bed
- Regenerative Incinerator Operating in a Cycle Mode	- Fluid-Bed

2.3 General Treatment of Material and Energy Balances

In the sizing and costing of the incinerator and the calculation of the auxiliary fuel requirements, it is necessary to make material and energy balances around the entire incinerator unit and around certain parts of the unit, such as the combustion chamber or the preheater. This section presents a general approach to making these balances.

These balances are based on the law of conservation of mass and energy. They can be stated in general equation form as

$$In - Out + Generation = Accumulation \tag{2.4}$$

Because the incineration process is a steady-state process, the accumulation term is zero and the equation becomes:

$$In - Out + Generation = 0 \tag{2.5}$$

For ***mass*** balances it is useful to restrict the balances to be made on the mass of each atomic species so that for mass balances the generation term becomes zero. However, because the combustion reaction liberates energy, the energy balances around equipment where combustion takes place would include a generation term. Thus, the simplified equations are

$$In - Out = 0, \quad for\ steady-state\ mass\ balances \tag{2.6}$$

$$In - Out + Generation = 0, \quad for\ steady-state\ energy\ balances \tag{2.7}$$

For the incineration process the two terms *In* and *Out* are generally mass terms (for a mass balance) of the form,

$$\dot{m} = \rho\, q \tag{2.8}$$

where

ρ = density (mass per unit volume)
Q = volumetric flow rate (volume per unit time)

or sensible heat terms (for an energy balance) of the form,

$$\dot{Q} = \dot{m}\, \Delta h = \rho\, q\, p \left(T - T_{ref}\right) \tag{2.9}$$

where

C_ρ = heat capacity
T = temperature

The reference temperature, T_{ref}, is often taken to be zero or the temperature of a convenient stream, e.g., the inlet gas stream, in whatever units T is in, so the T_{ref} term may not appear in the equations. When the reference temperature is taken as zero, the sensible heat terms become

$$\dot{Q} = r\, Q\, C_p\, T \tag{2.10}$$

Energy losses, H_L, are also part of the *Out* term and, for the incinerator process, are taken here to be 10% of the total energy input to the incinerator.

For the incineration process, the generation term for energy balances accounts for the energy released through the combustion reactions. This term is generally of the form

$$\dot{Q} = r\, Q\, (-\Delta h_c) \tag{2.11}$$

where

$(-\Delta h_c)$ = heat of combustion.

2.4 Design Procedures

The following procedure is designed to provide parameters for use in developing a study cost estimate (accuracy ± 30%). The principal parameters of interest are

- flue gas flow rate, upon which all the equipment cost correlations are based.
- auxiliary fuel requirement, which is important in estimating annual operating costs.

For applications which involve control of waste gas streams that are dilute mixtures of VOCs in air (>20% oxygen in the waste gas stream), the flue gas flow rate is greater than the inlet waste gas flow rate by the amount of auxiliary fuel and the increase in the moles of gas as a result of the combustion reaction. Because these two factors usually cause only small increases in flow rate, a number of simplifying assumptions can be made in the design calculations. For applications where diluent air must be used to adjust the combustible concentration in the waste gas to 25% LEL and where auxiliary fuel and auxiliary combustion air are needed, more complete mass and energy balances must be made.

The design procedure illustrated below is for waste gas streams that are dilute mixtures of VOCs in air (>20% oxygen in the waste gas stream). In this discussion the design procedure will be illustrated by a sample problem that will be solved step-by-step.

2.4.1 Steps Common to Thermal and Catalytic Units

Step 1 - Establish design specifications The first step in the design procedure is to determine the specifications of the incinerator and the waste gas to be processed. The following parameters of the waste gas stream at the emission source must be available:

- Volumetric flow rate, scfm—Standard conditions are normally 77°F and 1 atm. pressure
- Temperature
- Oxygen content
- Chemical composition of the combustibles
- Inerts content
- Heating value—In cases the heating value may act as a surrogate for the chemical composition of the combustibles. This is particularly true for dilute mixtures of combustibles in air.

- Particulate content—The particulate content is important if catalytic incinerators are to be coated. An upstream filter may suffice if particulate content is too high. Fluid-bed catalytic incinerators can tolerate higher particulate contents than fixed-bed catalytic incinerators.

The following parameters must be specified for the incinerator:

- Desired control efficiency—This efficiency should be based on requirements dictated by relevant state and federal regulations.
- Combustion chamber outlet temperature—This temperature may also be based on requirements of a regulation or on recommendations developed during regulatory development.
- Desired percent energy recovery—The desired percent energy recovery should be the result of a process optimization in which costs of incinerators with several different levels of energy recovery are estimated and the minimum cost design selected. The tradeoff is between the capital cost of the energy recovery equipment and the operating (fuel) cost.

Specifications for the sample problem are given in Table 2.4.

Step 2 - Verify that the oxygen content of the waste gas exceeds 20% There must be sufficient oxygen in the waste gas to support the combustion of the waste organics (including VOCs) and the auxiliary fuel, if auxiliary fuel is needed. It may be necessary to add auxiliary air if the oxygen content is less than about 20%. This example is based on streams that contain >20% oxygen, as shown below for 1000 ppmv of benzene and methylchloride:

$$Air\ Content,\ Vol.\ \% = 100 - \frac{1{,}000\ (ppm)}{10^6} \times 100 - \frac{1{,}000\ (ppm)}{10^6} \times 100 = 99.8\% \qquad (2.12)$$

This gives the oxygen content in percent as:

$$Oxygen\ Content = 98.8\% \times 0.209 = 20.86\% \qquad (2.13)$$

where 0.209 is the concentration of oxygen in air.

Step 3 - Calculate the LEL and the Percent of the LEL of the gas mixture Note: If the waste stream contains a significant amount of inerts in addition to the nitrogen associated with the

Tabla 2.4: Especificaciones para el Problema de Muestra

Variable	Valor
RelaciPreheater Inlet Wast Gas Vol. Flow Rate, Q_{wi}, scfm	20,000
Preheater Inlet Waste Gas Temp., T_{wi}, °F	100
Composition	
- Benzene Cointent, ppmv	1000
- Methyl Chloride Content, ppmv	1000
- Air Content	Balance
Particulate Content	Negligible
Moisture Content	Negligible
Desired Control Efficiency, %	98
Desired Percent Energy Recovery, HR%	70

oxygen in air, the calculation of LEL (and UEL) loses meaning since the LEL (and UEL) is measured in mixtures of organic with air only. A complete chemical analysis is necessary to complete the design procedure in such a case.

The example chosen here is typical, in that there is more than one VOC component in the gas stream. An approximate method to calculate the LEL of a mixture of compounds, LEL_{mix}, is given by Grelecki [13] as

$$LEL_{mix} = \left[\sum_{j=1}^{n} \frac{x_j}{\left(\sum_{i=1}^{n} x_i \right) LEL_j} \right]^{-1} \qquad (2.14)$$

where

x_i = volume fraction of combustible component i
LEL_j = lower explosive limits of combustible component j (ppmv)
n = number of combustible components in mixture

For the example case,

$$\sum_{i=1}^{n} x_i = (1{,}000 + 1{,}000) \times 10^{-6} = 2{,}000 \times 10^{-6}$$

From standard references [13] or from Appendix A,

LEL_{Bz} = 14,000 ppmv for benzene
LEL_{MC} = 82,5000 ppmv for methyl chloride

$$LEL_{mix} = \left[\frac{1{,}000}{2{,}000 \times 14{,}000} + \frac{1{,}000}{2{,}000 \times 82{,}500}\right]^{-1} = 23{,}938 \ ppmv$$

$$\%\ LEL_{mix} = \frac{total\ combustible\ conc.\ in\ mixture}{LEL_{mix}} \times 100 \qquad (2.15)$$

$$\%\ LEL_{mix} = \frac{2{,}000}{23{,}938} \times 100 = 8.4\%$$

The percent *LEL* of the mixture is therefore 8.4%. Because this is well below 25%, no dilution air is needed in this example. If the mixture had been above 25% *LEL*, sufficient dilution air would have been needed to bring the concentration of the mixture to less than 25% to satisfy fire insurance regulations.

Step 4 - Calculate the volumetric heat of combustion of the waste gas streams, (- Δh_{c_w}), Btu/scf The energy content of the gas stream, expressed in terms of the heat of combustion, is calculated as follows:

$$\left(-\Delta h_{c_w}\right) = \sum_{i=1}^{n} \left(-\Delta h_{c_i}\right) x_i \qquad (2.16)$$

where

$(\&\Delta h_{cw})$	= heat of combustion of the waste stream (Btu/scf)
$(\&\Delta h_{ci})$	= volumetric heat of combustion of component i at 25°C (Btu/scf)
x_i	= volume fraction of component *i* in the waste gas
n	= number of combustible components in the waste gas

The heat of combustion that should be used in these calculations is the "lower" heat of combustion, i.e., with gaseous water, rather than liquid water, as a reaction product since water leaves the incinerator in the vapor state. From Appendix 3F or standard references [79,80] with appropriate conversion of units, the volumetric heat of combustion at 25°C for the two components is calculated to be as follows:

$(-\Delta h_{cBc})$ = 3,475 Btu/scf for Benzene

$(-\Delta h_{cMC})$ = 705 Btu/scf for mrthyl chloride

The compositions specified earlier as ppmv are converted to volume fractions as follows:

χ_{Bz} = 1,000 ppmv × 10^{-6} = 10^{-3} for benzene
χ_{MC} = 1,000 ppmv × 10^{-6} = 10^{-3} for methyl chloride

Using these values of heat of combustion and composition, the heat of combustion of the waste gas stream per standard cubic foot (scf) of incoming gas is

$$\left(-\Delta h_{c_w}\right) = (3{,}475)\left(10^{-3}\right) + (705)\left(10^{-3}\right) = 4.18 \ \frac{Btu}{scf}$$

Assuming the waste gas is principally air, with a molecular weight of 28.97 and a corresponding density of 0.0739 lb/scf, the heat of combustion per pound of incoming waste gas is:

$$(-\Delta h_{cw}) = 56.6 \ Btu/lb$$

The negative heat of combustion, by convention, denotes an exothermic reaction. Also by convention, if one refers to heat of reaction rather than heat of combustion, then a positive value denotes an exothermic reaction.

Empirically, it has been found that 50 Btu/scf roughly corresponds to the *LEL* of organic/ air mixtures. Insurance codes require a value below 25% *LEL*, which corresponds to about 13 Btu/scf. However, if *LEL* sensors and monitors are installed, one can incinerate a waste gas with a combustible organic content between 25 and 50% *LEL*, which corresponds to 13 to 25 Btu/scf.

For catalytic applications the heat of combustion must normally be less than 10 Btu/scf (for VOCs in air) to avoid excessively high temperatures in the catalyst bed. This is, of course, only an approximate guideline and may vary from system to system.

After Step 4, determination of the ($-\Delta h_{cw}$) design procedure for thermal and catalytic incinerators is discussed separately, beginning with Step 5 for each type of incinerator.

2.4.2 Steps Specific to Thermal Units

Figure 2.1 shows a generic thermal incinerator with the appropriate streams labeled.

Step 5t - Establish the temperature at which the incinerator will operate As mentioned in Section 2.2.1, both the reactor temperature and residence time of the waste gas in the reactor determine the level of VOC destruction. In general, state and local regulations specify the required level of destruction that the customer must meet. In this example a destruction efficiency of 98 % is specified. Studies by Mascone [2,3,4] show that this destruction efficiency can be met in a thermal incinerator operated at a temperature, T_{fi}, of 1600°F and a residence time of 0.75 second. (Note: This higher efficiency level is the minimum achievable by any new properly designed and operated incinerator. Many incinerators can achieve destruction efficiencies of 99% or higher.)

Step 6t - Calculate the waste gas temperature at the exit of the preheater The extent of the heat exchange to be carried out in the preheater is the result of a technical and economic optimization procedure that is not illustrated in this example. As the VOC stream temperature leaving the heat exchanger, T_{wo}, increases, the auxiliary fuel requirement decreases, but at the expense of a larger heat exchanger. However, there are several important limits on T_{wo}. First, T_{wo} must not be close to the ignition temperature of the organic-containing gas to prevent damaging temperature excursions inside the heat exchanger should the gas ignite. Second, for gases containing halogens, sulfur, and phosphorous (or other acid-forming atoms), the flue gas temperature after the heat exchanger, T_{fo}, must not drop below the acid dew point. Both limitations limit the amount of heat exchange and thus the maximum value of T_{wo}. The calculation of the acid dew point is not simple. It is recommended that vendor guidance be sought to ensure that the dew point is not reached. Condensation of acid gases will result in corrosion of many of the metals used in heat exchangers. As an example, fuel sulfur contents of 1 to 2 percent can give acid dew points of about 200 to 270°F. Increasing the sulfur content to 4 percent can raise the dew to about 290°F. Chlorine and phosphorous have a much smaller effect on acid dew elevation.

With the following assumptions, one can estimate T_{w_o} using Equation 2.2, the definition of fractional energy recovery for a heat exchanger:

- The fractional energy recovery is specified.
- The amount of auxiliary fuel, Q_{af}, and auxiliary combustion air, Q_a, are small relative to the waste gas, Q_w, so that the mass flow rates of gases, $p_w Q_w$ and $p_f Q_f$, on both sides of the preheater are approximately the same, or:

$$\rho_w \, Q_w = \rho_f \, Q_f \qquad (2.17)$$

- The heat capacities of the gases on both sides of the preheater are approximately the same, regardless of composition. This is true for waste streams which are dilute mixtures of organics in air, the properties of the streams changing only slightly on combustion.
- The mean heat capacities above the reference temperature of the gases on both sides of the preheater are approximately the same regardless of temperature.

With these assumptions, the equation for fractional energy recovery for a heat exchanger becomes:

$$Fractional\ Enery\ Recovery = \frac{T_{w_o} - T_{w_i}}{T_{f_i} - T_{w_i}}$$

For this example with a fractional energy recovery of 0.70, an incinerator operating temperature, T_{f_i}, of 1600°F, and a waste gas inlet temperature, T_{w_i}, of 100°F, the waste gas temperature at the end of the preheater becomes:

$$T_{wo} = 1,150^{o}F \tag{2.18}$$

The temperature of the exhaust gas, T_{f_o}, can be determined by an energy balance on the preheater, which, with the same assumptions as used in deriving Equation 2.18 regarding the mass flow rates and average heat capacities of the gases involved, results in the following equation:

$$T_{f_i} - T_{f_o} = T_{w_o} - T_{w_i} \tag{2.19}$$

e.g., the temperature rise in the waste gas is approximately equal to the temperature decrease in the flue gas with which it is exchanged. For this example, this results in

$$T_{fo} = 550^{o}F \tag{2.20}$$

This value of T_{f_o} should be well above the acid dew point of the flue gas stream.

It should be remembered that T_{w_o} should be well below the ignition temperature of the VOC stream to prevent unwanted temperature excursions in the preheater. This must be verified even if the stream is well below the *LEL* because flammability limits can be expanded by raising the reactant stream temperature. A sufficiently high preheat temperature, T_{w_o}, could initiate reaction (with heat release) in the preheater. This would ordinarily be detrimental to the materials of construction in the heat exchanger. The one exception is the thermal incinerator of the regenerable type described in Section 2.1 The 95-percent energy recovery, obtainable in regenerable systems would result in this example in a T_{w_o} of 1,525°F. The significant reaction rate that would occur at this temperature in the ceramic packing of the heat exchanger/reactor is by design.

Step 7t - Calculate the auxiliary fuel requirement, Q_{af}. Auxiliary fuel will almost invariably be needed for startup of the unit. However, at steady state, if the energy released by combustion of the organics present in the waste stream is sufficient to maintain the reactor temperature (1600°F in the example), only a small amount of auxiliary fuel (< 5% of the total energy input) is needed to stabilize the flame. In most cases, however, more fuel than just this stabilizing fuel will be required to maintain the reactor temperature.

With the following assumptions, one can estimate Q_{af} using a mass and energy balance around the combustion chamber and following the principles discussed in Section 2.2, with reference to Figure 2.1.

- The reference temperature, T_{ref}, is taken as the inlet temperature of the auxiliary fuel, T_{af}.

- No auxiliary air, Q_a, is required.

- Energy losses, H_L, are assumed to be 10% of the total energy input to the incinerator above ambient conditions.[16,17] Thus, if the reference temperature is near ambient conditions,

$$H_L = 0.1\ \rho_{fi}\ Q_{fi}\ C_{pm_{fi}}\left(T_{fi} - T_{ref}\right) \tag{2.20}$$

- The heat capacities of the waste gases entering and leaving the combustion chamber are approximately the same, regardless of *composition*. This is true for waste streams which are dilute mixtures of organics in air, the properties of the streams changing only slightly on combustion.

- The mean heat capacities above the reference temperature of the waste gases entering and leaving the combustion chamber are approximately the same regardless of temperature. Thus the mean heat capacity for the waste gas stream entering or leaving the combustion chamber should be evaluated at the average of T_{w_o} and T_{fi}. For air this assumption introduces an error of, at most, 5% over the temperatures of interest.

With these assumptions, the mass and energy balance around the combustion chamber reduces to the following equation:

$$Total\ Energy\ Input = r_{fi}\ Q_{fi}\ C_{pm_{fi}}\left(T_{fi} - T_{ref}\right) \tag{2.21}$$

Input data for this equation are summarized below:

The waste stream is essentially air so that

$$\rho_{wo} = \rho_{\omega i} = 0.0739 \text{ lb/scf, air at } 77^{\circ}\text{F, 1 atm}$$

$C_{pm_{air}} = 0.255$Btu/lb F, the mean heat capacity of air between 77°F and 1375°F (the average temperature of the waste gas entering and leaving the combustion chamber)

Other input data to Equation 2.21 include:

$$
\begin{aligned}
Q_{wo} &= Q_{wi} = 20{,}000 \text{ scfm} \\
(\&\Delta h_{caf}) &= 21{,}502 \text{ Btu/lb, for methane} \\
T_{af} &= T_{ref} = 77°\text{F, assume ambient conditions} \\
\rho_{af} &= 0.0408 \text{ lb/ft}^3\text{, methane at } 77°\text{F} \\
T_{fi} &= 1{,}600°\text{F, Step 5t} \\
T_{wo} &= 1{,}150°\text{F, Step 6t} \\
(\&\Delta h_{cwo}) &= 56.6 \text{ Btu/lb, Step 4}
\end{aligned}
$$

Substituting the above values into Equation 2.21 results in:

$$Q_{af} = 167 \text{ scfm} \qquad (2.21a)$$

The values of the parameters in the energy balance are summarized in Table 2.5.

It is instructive to examine the magnitude of the various terms in the energy balance around the combustor for the sample problem. This is done in Table 2.6. The energy balance shown does not quite add to zero due to round-off-error and simplifying assumptions. Table 2.6 shows that the largest inlet term is the sensible heat of the incoming waste gas. The heat of combustion of the organics contained in the waste gas stream is somewhat smaller than that of the auxiliary methane because of the relatively small amount of organics in the waste gas stream. The largest term in the outlet stream is the sensible heat of the outgoing waste stream. The overall energy losses are based on an assumption, but are relatively small. Because the sensible heat contents of the entering and leaving waste stream are so large, it is apparent that energy recovery is an important factor in achieving energy efficiency. In fact, with zero energy recovery in the sample problem, the auxiliary fuel requirements would be 605 scfm, about four times the energy requirements based on 70% energy recovery.

Step 8t - Verify that the auxiliary fuel requirement is sufficient to stabilize the burner flame Only a small amount of auxiliary fuel (< 5% of the total energy input) is needed to stabilize the burner flame. In general, more fuel than just this stabilizing fuel will be required to maintain the reactor temperature. It is wise to verify that the auxiliary fuel requirement calculated in Step 7t is sufficient for stabilization. If it is insufficient, then a minimum amount of auxiliary fuel must be used, and the amount of energy recovery, specified earlier must be reduced to avoid exceeding the specified temperature at which the incinerator will operate (Step 5t).

This check is made by calculating 5% of the total energy input to the incinerator and comparing it with the auxiliary fuel energy input. The total energy input is given as follows:

Table 2.5: Summary of Example Problem Variable Valuation T_{ref} = 77°F

Stream	Subscript, *j*		ρ_j, lb/scf	Q_j, scfm C_{pmj},	Btu/lb °F	T_j, °F
IN	Sensible Heat					
	Auxiliary Air	a	na	na	na	na
	Auxiliary Fuel	af	0.0408	167	1	77
	Waste Gas	w_o	0.0739	20,000	0.255	1,150
OUT	Sensible Heat					
	Waste Stream	f_i	0.0739	20,167	0.255	1,600

($-\Delta h_c$), waste gas = 56.6 Btu/lb
($-\Delta h_c$), auxiliary fuel = 21,502 Btu/lb
na = Not Applicable
1 = Not used becaused reference temperature is taken equal to auxiliary fuel temperature.

Table 2.6: Terms in Energy Balance Around Combuster - Example Problem

Stream		Subscript, j	Value, Btu/min
IN	Sensible Heat, $\rho_j Q_j C_{pmj}(T_i - T_{ref})$		
	Auxiliary Air	*a*	0
	Waste Gas	w_o	404,403
OUT	Sensible Heat, $\rho_j Q_j C_{pmj}(T_i - T_{ref})$		
	Waste Stream	f_i	578,796
OUT	Losses		
	10% of total energy input		57,8800
GENERATION			
	Heat Combustion, $\rho_j Q_j(-\Delta h_{cf})$		
	Waste Gas	w_o	83,655
	Auxiliary Fuel	*af*	146,506

$$Total\ Energy\ Input = r_{fi}\ Q_{fi}\ C_{pm_{fi}}\ \left(T_{fi} - T_{ref}\right) \qquad (2.22)$$

$$Auxiliary\ Fuel\ Energy\ Input = r_{af}\ Q_{af}\ \left(-\Delta h_{c_{af}}\right) \qquad (2.23)$$

The auxiliary fuel used in the design, Q_{af}, should be the larger of 5% of the total energy input (28,900 Btu/min.) and the auxiliary fuel energy input (146,500 Btu/min.). The auxiliary fuel used easily meets this criterion.

Step 9t - Calculate the total volumetric flow rate of gas through the incinerator, Q_{fi} The total volumetric flow rate of gas leaving the incinerator is referred to as the flue gas flow rate, Q_{fi}, and is the gas rate on which the incinerator sizing and cost correlations are based. The flue gas flow rate measured at the standard conditions of 77°F and 1 atmosphere, where the increase in volumetric throughput due to an increase in the number of moles of gas as a result of combustion is neglected, is the sum of the inlet streams to the incinerator.

$$\begin{aligned} Q_{f_i} &= Q_{w_a} + Q_a + Q_{af} \\ &= 20{,}000 + 0 + 167 \\ &= 20{,}167\ scfm \end{aligned} \qquad (2.24)$$

This result conforms with the assumptions stated in Step 6t, e.g., the mass (and volume) flow rates on both sides of the preheater are approximately equal. Finally, it must be emphasized that steps 5t to 9t apply to thermal recuperative incinerators only. To calculate the auxiliary fuel requirements for other types of thermal incinerators (e.g., regenerative), a different procedure must be used. (See Appendix B.)

2.4.3 Steps Specific to Catalytic Units

Figure 2.3 shows a generic catalytic incinerator with the appropriate streams labeled. The approach used in the calculations on the catalytic incinerator is somewhat different than that used in the thermal incinerator. This difference arises because of additional constraints which are placed

on the catalytic incinerator. These constraints are as follows:

- The desired catalyst bed outlet temperature is typically 700 to 900°F. The maximum temperature to which the catalyst bed can be exposed continuously is limited to about 1200°F. Therefore, the combustible content of the waste gas is limited, and the amount of heat exchange that occurs in the primary heat exchanger may be limited.

- The inlet temperature to the catalyst bed itself must be above the catalytic ignition temperature required to give the desired destruction efficiency in the incinerator. Therefore, the combustible content of the waste gas is further limited to that which, when combusted, will raise the temperature in the catalyst bed no more than the ΔT between the required reactor bed inlet temperature, and the desired reactor bed outlet temperature.

- Auxiliary fuel, in combination with the preheat from the primary heat exchanger, is used to preheat the waste gas to the reactor inlet temperature. A minimum amount of auxiliary fuel (< 5% of the total energy input) must be used to stabilize the flame in the preheat combustion chamber. This has the effect of further limiting the combustible content of the waste gas stream and the amount of heat exchange permissible in the primary heat exchanger.

The steps outlined below represent one approach to recognizing these constraints and incorporating them into the calculation procedures.

Step 5c - Establish the desired outlet temperature of the catalyst bed, T_{fi} The energy released by the oxidation of the VOCs in the catalyst bed will raise the temperature of the gases by an amount, ΔT, as the gases pass through the catalyst bed. An outlet temperature from the catalyst, and thus from the reactor, must be specified that will ensure the desired level of destruction of the VOC stream. As in thermal incinerators, this value varies from compound to compound and also varies from catalyst to catalyst. Final design of the incinerator should be done by firms with experience in incinerator design. Guidelines given by Combustion Engineering [12] indicate that values from 300 to 900°F result in destruction efficiencies between 90 and 95 percent. To prevent deactivation of the catalyst a maximum bed temperature of 1200°F should not be exceeded. In the example problem the catalyst outlet temperature, T_{fi}, is selected to be 900°F.

Step 6c - Calculate the waste gas temperature at the exit of the preheater (primary) heat exchanger The waste gas temperature at the exit of the primary heat exchanger is estimated in the same manner as for the thermal incinerator. The equation for fractional energy recovery Equation 2.18, is used, with the same assumptions as used for the thermal incinerator. For the example problem with a fractional energy recovery of 0.70, a catalyst bed outlet temperature, T_{fi}, of 900°F, and a waste gas inlet temperature, T_{w_i}, of 100°F, the gas temperature at the exit of the preheater

becomes

$$T_{wo} = 660°F$$

The same considerations regarding the closeness of the temperature of the exhaust gas, T_{fa}, to its dew point apply to the catalytic incinerator as they did to the thermal incinerator.

Step 7c - Calculate the auxiliary fuel requirement, Q_{af}. The auxiliary fuel requirement, Q_{af}, is calculated by making mass and energy balances around the preheater combustion chamber and the catalyst chamber. The auxiliary fuel requirement calculated in this manner must be checked to insure that it falls within the constraints imposed by design considerations of the catalytic incinerator. These constraints are as follows:

- The auxiliary fuel requirement must be positive. A negative fuel requirement indicates that the heat of combustion of the waste gas, $(-\Delta h_c)$, is too high for the fractional energy recovery in the primary heat exchanger that was selected.

- The auxiliary fuel amount must be high enough to provide a stable flame in the preheater combustion chamber (See Step 8c below).

An energy balance around the preheater combustion chamber and the catalyst chamber, taken together, results in Equation 2.21, the same equation used in the thermal incinerator calculations. The input data for Equation 2.21 for the catalytic incinerator example problem are summarized below:

- The waste stream is essentially air so that

  ρ_{wo} = ρ_{wi} = 0.0739 lb/scf, air at 77°F, 1 atm

  C_{pmair} = 0.248 Btu/lb F, the mean heat capacity of air between 77°F and 780°F (the average of the preheater exit and catalyst bed outlet temperatures)

- Other input data to Equation 2.21 include:

$$Q_{wo} = Q_{wi} = 20{,}000 \text{ scfm}$$

$(\&\Delta h_{caf})$ = 21,502 Btu/lb, for methane
$T_{af} = T_{ref}$ = 77°F, assume ambient conditions
ρ_{af} = 0.0408 lb/ft^3, methane at 77°F
T_{fi} = 1,600°F, from Step 5c
T_{wo} = 1,150°F, from Step 6c
$(\&\Delta h_{cwo})$ = 56.6 Btu/lb, from Step 4

Substituting the above values into Equation 2.21 results in

$$Q_{af} = 40 \text{ scfm}$$

If the outlet temperature of the catalyst bed, T_{fi}, is 800°F, then Q_{af}, decreases to -6.7 scfm. In other words, no auxiliary fuel would, theoretically, be required at this bed temperature. However, as discussed above in Step 8t, a certain quantity of auxiliary fuel would be required to maintain burner stability.

At 70% energy recovery and 900°F outlet catalyst bed temperature, a waste gas with a heat of combustion, $(-\Delta h_{c_{wo}})$, of about 79.9 Btu/lb would cause the auxiliary fuel requirement, Q_{af}, to become negative, indicating the catalyst bed would exceed 900°F. At 70% energy recovery and 800°F outlet catalyst bed temperature, this same result occurs with a $(-\Delta h_{c_{wo}})$ of 52.7 Btu/lb. Both of these heats of combustion are relatively low for typical waste gases. These results are, of course, dependent on the assumption of energy losses from the combustion chamber. The lower the energy losses, the lower the allowable waste gas heat of combustion before overheating occurs in the catalyst bed.

Step 8c - Verify that the auxiliary fuel requirement is sufficient to stabilize the burner flame Only a small amount of auxiliary fuel (< 5% of the total energy input) is needed to stabilize the burner flame. In general, more fuel than just this stabilizing fuel will be required to maintain the reactor temperature. It is wise to verify that the auxiliary fuel requirement calculated in Step 7c is sufficient for stabilization. If it is insufficient, then a minimum amount of auxiliary fuel must be used and the amount of energy recovery specified earlier must be reduced to avoid exceeding the specified temperature at which the incinerator will operate (Step 5c).

This check is made in the same manner as that in Step 8t of the thermal incinerator calculation. The results of this check indicate that the auxiliary fuel requirement is more than sufficient to stabilize the burner flame.

Step 9c - Estimate the inlet temperature to the catalyst bed, T_{r_i} The inlet temperature to the catalyst bed must be calculated to ensure that the inlet temperature is above that necessary to ignite the combustible organic compounds in the catalyst that was selected for use.

The inlet temperature to the catalyst bed, T_{r_i}, should be such that, when the temperature rise through the catalyst bed, T, is added to it, the resulting temperature is T_{f_i}, 900°F. Thus,

$$\Delta T = T_{f_i} - T_{r_i} \tag{2.25}$$

The value of ΔT is determined by an energy balance around the preheater portion of the combustor. The preheater is required to heat the gases up to the catalyst bed inlet temperature using auxiliary fuel.[1] This energy balance is made with the assumptions made earlier in deriving Equation 2.21 and further assuming that only auxiliary fuel is combusted in the preheater portion. The resulting equation is very similar to Equation 2.21 except that (1) the terms with an f_i subscript become terms with r_i subscripts to denote a catalytic reactor inlet stream rather than a combustor outlet (flue gas inlet to the primary heat exchanger) and (2) the term for combustion of the waste gas organics does not appear. The resulting equation is as follows:

$$\rho_{af}\ Q_{af} = \frac{\rho_{w_o}\ Q_{w_o}\left[C_{pm_{air}}\left(1.1\ T_{r_i} - T_{w_o} - 0.1\ T_{ref}\right)\right]}{\left(-\Delta h_{c_{af}}\right) - 1.1\ C_{pm_{air}}\left(T_{r_i} - T_{ref}\right)} \tag{2.26}$$

This equation may be rearranged to solve for T_{r_i} explicitly. This produces an equation that is somewhat complex and non-intuitive.

$$T_{r_i} = \frac{\rho_{af}\ Q_{af}\left[\left(-\Delta h_{c_{af}}\right) + 1.1\ C_{pm_{air}}\ T_{ref}\right] + r_{w_o}\ Q_{w_o}\ C_{pm_{air}}\left(T_{w_o} + 0.1\ T_{ref}\right)}{1.1\ C_{pm_{air}}\left(\rho_{af}\ Q_{af} + \rho_{w_o}\ Q_{w_o}\right)} \tag{2.27}$$

After substituting the example problem parameters into Equation 2.27, we obtain a value for T_{r_i} of 693°F. Based on ignition temperatures shown in Table 2.2, this reactor inlet temperature should be satisfactory. Prior to a more definitive design, the ignition temperatures for the specific chemicals should be verified.

[1] At equilibrium, the temperature of the catalyst bed is maintained without requiring auxiliary fuel.

The temperature rise across the catalyst bed is thus (900 - 693) or 207°F. These temperatures are somewhat sensitive to the assumption for energy losses from the combustor. The assumption for energy losses is perhaps somewhat conservative, i.e., it causes a larger Q_{af} to be estimated than would a less conservative assumption, and becomes more conservative as the combustor size and insulation are increased.

Step 10c - Calculate the total volumetric flow rate of gas through the incinerator, Q_{fi} The total volumetric flow rate of gas leaving the incinerator is referred to as the flue gas flow rate, Q_{fi}, and is the gas rate on which the incinerator sizing and cost correlations are based. The flue gas flow rate measured at the standard conditions of 77°F and 1 atmosphere, where the increase in volumetric throughput due to an increase in the number of moles of gas as a result of combustion is neglected, is the sum of the inlet streams to the incinerator.

$$\begin{aligned} Q_{fi} &= Q_{wo} + Q_a + Q_{af} \\ &= 20{,}000 + 0/40 \\ &= 20{,}040\text{scfm} \end{aligned}$$

Step 11c - Calculate the volume of catalyst in the catalyst bed If the volumetric flow rate of gas through the catalyst bed, Q_{fi}, and the nominal residence time (reciprocal space velocity) in the catalyst bed are known, then the volume of catalyst can be estimated. There exists a complex set of relationships between the catalyst volume and geometry, overall pressure drop across the catalyst, conversion of the oxidizable components in the gas, gas temperature, and the reaction rate. These relationships are dependent on the catalyst and the type of compound being oxidized. It is beyond the scope if this Manual to discuss these relationships, even in an approximate way. For the purposes of cost estimation, the space velocity, in reciprocal time units, necessary to achieve the required level of destruction can be used to approximate the catalyst volume requirement. The space velocity is defined as

$$\phi = \frac{Q_{ft}}{V_{cat}} \qquad (2.28)$$

where

V_{cat} = Overall bulk volume of the catalyst bed, including interparticle voids (ft^3)

By petro-chemical industry convention, the space velocity is computed at the conditions of 60°F (not 77°F) and 1 atm. The volumetric flow rate, Q_{fi}, must be corrected to these conditions. The proper space velocity to achieve a desired level of conversion is based on experimental data for the system involved. For precious metal monolithic catalysts, the space velocity generally lies

between 10,000 h^{-1} and 60,000 h^{-1}. (Base metal catalysts operate at lower space velocities, ranging from 5,000 to 15,000 h^{-1}.)[75]

There are a number of catalyst bed parameters, such as catalyst configuration and bed design, that are not significant for study type cost estimates. Accordingly, design of these factors is not discussed here.

2.5 Cost Analysis

This section presents procedures and data for estimating capital and annual costs for four types of incinerators:(1) thermal-recuperative, (2) thermal regenerative, (3) fixed-bed catalytic, and (4) fluid-bed catalytic.

2.5.1 Estimating Total Capital Investment

Total capital investment, (TCI), includes the equipment cost, EC, for the incinerator itself, the cost of auxiliary equipment (e.g., ductwork), all direct and indirect installation costs, and costs for buildings, site preparation, offsite facilities, land, and working capital. However, the last five costs usually do not apply to incinerators. (See Section 1of this Manual for a detailed description of the elements comprising the TCI.) Although industry representatives were reluctant to provide updated costs in 1999 dollars, they did indicate costs have not significantly changed since 1988 [18,19,20]. In addition, SAIC obtained 11 vendor quotes for a specific configuration for three types of incinerator systems (recuperative, regenerative, fixed-bed catalytic). These quotes compared favorably to those generated from the cost equations.

2.5.1.1 Equipment Costs, EC

As discussed in Section 2.2.3, the EC given in this chapter apply to packaged incinerators, except for regenerative incinerators. For regenerative incinerators, the costs apply to field-erected units. The EC typically includes all flange-to-flange equipment needed to oxidize the waste gas, including the auxiliary burners, combustion chamber, catalyst, primary heat exchanger (except for the "zero heat recovery" cases), weathertight housing and insulation, fan, flow and temperature control systems, a short stack, and structural supports. Smaller units, e.g., typically less than 20,000 scfm, are typically preassembled skid-mounted [21]. The various available incineration systems are presented in four groups delineated according to their similarity of design. These groups are outlined in Table 2.7. With the exception of regenerative thermal and fluid-bed catalytic incinerators, the maximum size for which costs are given is 50,000 scfm. Although larger units of each technology can be built, applications are rare at flow rates above 50,000 scfm. Regenerative thermal incinerator costs are provided for flow rates from 10,000 to 100,000 scfm. Fluid-bed

catalytic incinerator costs are provided for flow rates from 2,000 to 25,000 scfm.

The cost curves are least-squares regressions of cost data provided by different vendors. It must be kept in mind that even for a given incineration technology, design and manufacturing procedures vary from vendor to vendor, so that costs may vary. As always, once the study estimate is completed, it is recommended that more than one vendor be solicited for a more

Table 2.7: Scope of Cost Correlations

Incinerator Type	Total (Flue) Gas Flowrate, scfm	Figure Number
Thermal – Recuperative	500[a] – 50,000	3.26
Thermal – Regenerative	10,000 – 100,000	3.27
Fixed-Bed Catalytic	2,000 – 50,000	3.28
Fluid-Bed Catalytic	2,000 – 25,000	3.29

[a] Although Figure 3.26 covers the 1,000 to 50,000 scfm range, the correlation is valid for the 500 to 50,000 scfm range.

detailed cost estimate.

The additional expense of acid gas clean-up or particulate control is not treated in this section. The equipment cost of a gas absorber to remove any acid gases formed in the incinerator can be quite large, sometimes exceeding the equipment cost of the incinerator itself even for simple packed tower scrubbers [22]. For more complex absorbers that include venturi scrubbers instead of, or in addition to, packed beds, the cost of the scrubber alone may be up to 4 times that of the incinerator [11]. These more complex absorbers are sometimes necessary when particulates, in addition to acid gases, must be removed from the flue gas.

Thermal Incinerators Among the thermal units, the direct flame (0% energy recovery) and recuperative systems are treated together because the various levels of energy recovery are achieved simply by adding heat exchanger surface area. Costs for these units were provided by several vendors [12,23,24]. The EC of these units are given as a function of total volumetric throughput, Q_{tot}, in scfm. "Q_{tot}", is the total volume of the gaseous compounds exiting the combustion chamber; it is identical to the term, "Q_{fi}," used in Figures 2.1 and 2.2. This includes the combustion products, nitrogen, unburned fuel and organics, and other constituents. (See Figure 2.4). Note that costs are given free on board (F.O.B.) in April 1988 dollars[12]. Based on a least-squares regression analysis, a log-log relationship between throughput and EC was found for a given level of energy

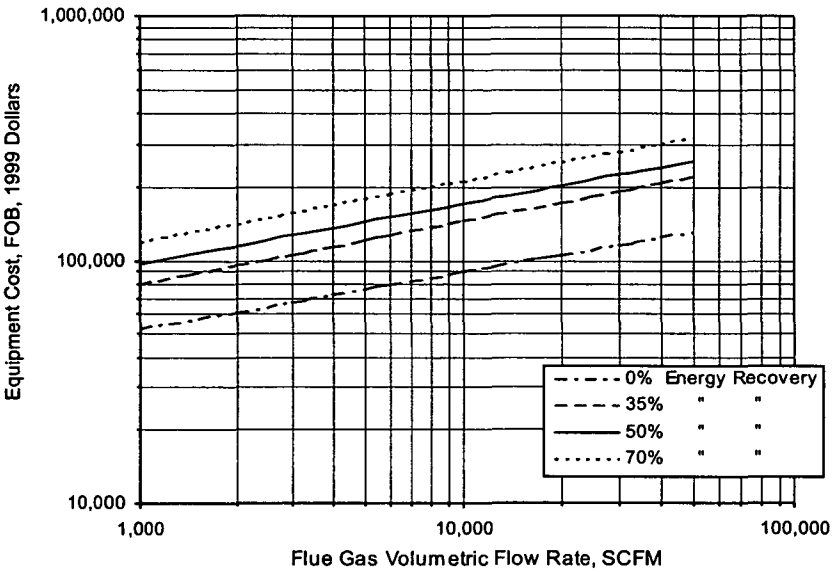


Figure 2.4: Equipment Costs of Thermal Incinerators, Recuperative

recovery (*HR*) over the flow rate range from 500 to 50,000 scfm. These relationships are as follows:

$$EC = 10{,}294\ Q_{tot}^{0.2355} \qquad HR = 0\% \tag{2.29}$$

$$EC = 13{,}149\ Q_{tot}^{0.2609} \qquad HR = 35\% \tag{2.30}$$

$$EC = 1{,}7056\ Q_{tot}^{0.2502} \qquad HR = 50\% \tag{2.31}$$

$$EC = 21{,}342\ Q_{tot}^{0.2500} \qquad HR = 70\% \tag{2.32}$$

The regenerative (or excess enthalpy) systems provide up to 95 percent heat recovery at the expense of higher capital costs. Their unique design [25,26], which combines the heat exchanger and reactor, is substantially different from traditional thermal units and is therefore treated separately in Figure 2.5. The ECs of these systems are given as an approximately linear function of total flow rate over a 10,000 to 100,000 scfm range by the following equation:

---

12 For information on escalating these and the other incinerator prices to more current dollars, refer to the EPA report *Escalation Indexes for Air Pollution Control Costs* and updates thereto, all of which are installed on the OAQPS Technology Transfer Network (CTC Bulletin Board).

$$EC = 2.204 \times 10^5 + 11.57\, Q_{tot} \qquad (2.33)$$

Again, the higher capital costs of these units can be substantially offset by the substantial savings in auxiliary fuel costs.

Catalytic Incinerator The EC for a catalytic incinerator is a function of the type of catalyst contacting pattern used and the total gas flow rate, Q_{tot}, for a given level of energy recovery. There are three types of contacting configurations used in catalytic systems: fixed-bed, catalytic monolith, and fluid-bed. The EC for the first two are generally comparable and are given in Figure 2.6. The data provided by several vendors [12,23,24,27] exhibited curvilinear relationships with Q_{tot} for *each* of the energy recovery rates. Least squares regressions of the data yielded the following correlations for total flow rates between 2,000 and 50,000 scfm:

$$EC = 1105\, Q_{tot}^{0.5471} \qquad HR = 0\% \qquad (2.34)$$

$$EC = 3623\, Q_{tot}^{0.4189} \qquad HR = 35\% \qquad (2.35)$$

$$EC = 1215\, Q_{tot}^{0.5575} \qquad HR = 50\% \qquad (2.36)$$

$$EC = 1443\, Q_{tot}^{0.5527} \qquad HR = 70\% \qquad (2.37)$$

Fluid-bed catalytic incinerators afford certain advantages over fixed-bed catalyst units in that they tolerate waste streams with (1) higher heating values, (2) particulate contents, and (3) chlorinated species. For this enhanced flexibility of feed streams, a higher capital cost is incurred,

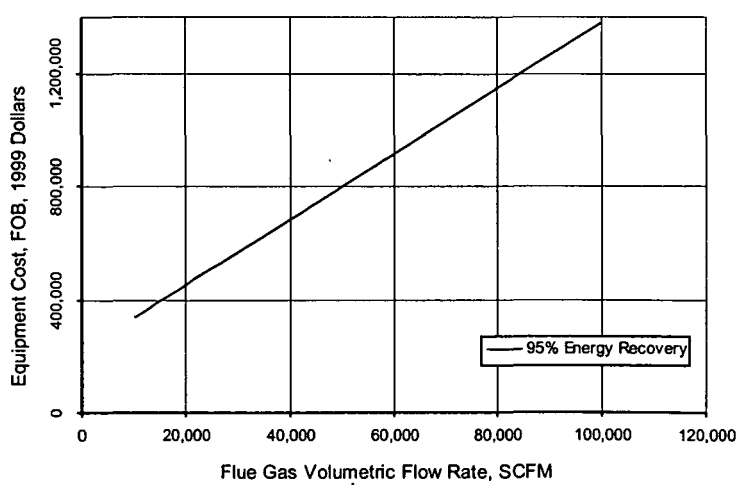


Figure 2.5: Equipment Costs of Thermal Incinerators, Regenerative

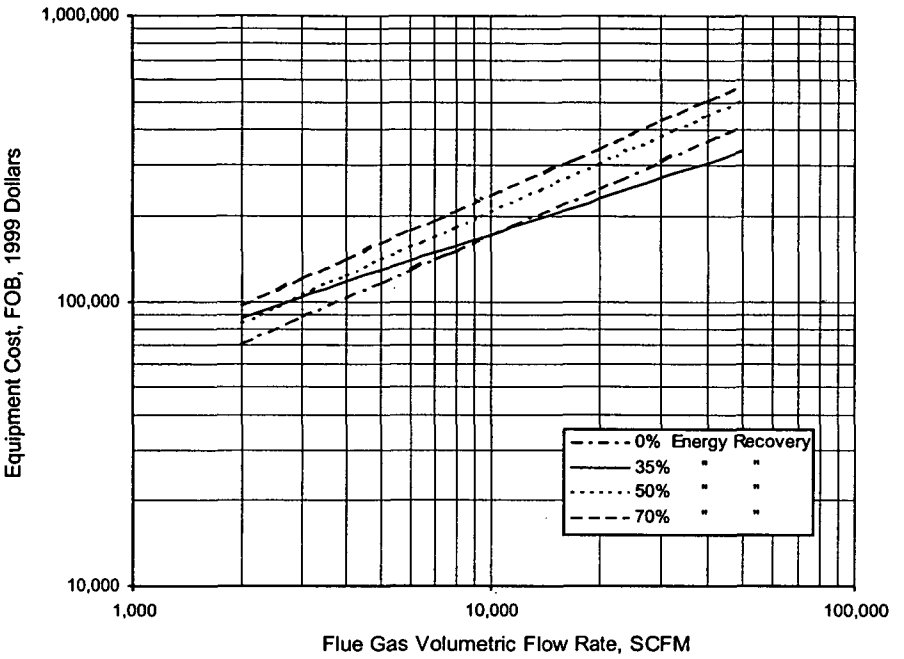


Figure 2.6: Equipment Cost of Catalytic Incinerators, Fixed-Bed

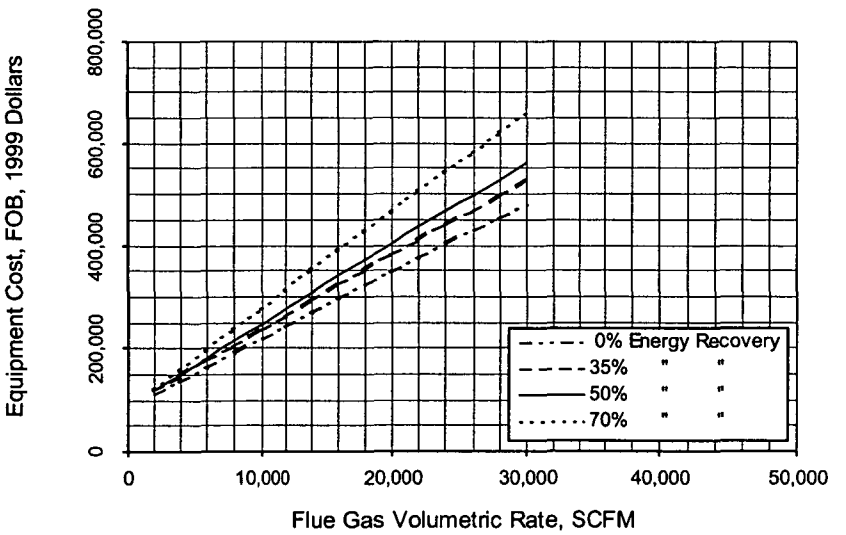


Figure 2.7: Equipment Costs of Catalytic Incinerators, Fluid-Bed

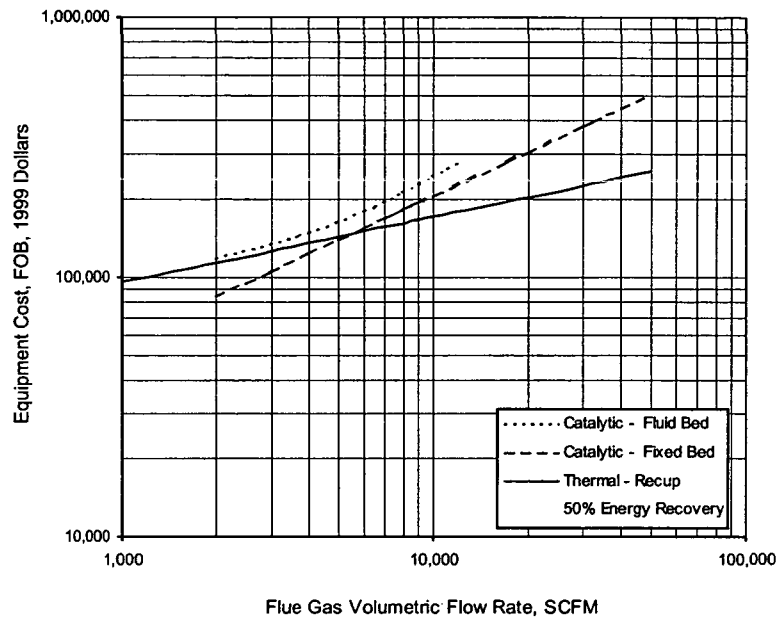


Figure 2.8: Equipment Costs Comparison of Incinerator Types

as indicated by the EC shown in Figure 2.7. The data shown were provided by vendors [1,22] and exhibited a linear relationship over the range of flow rates from 2,000 to 25,000 scfm. They can be approximated by the following equations:

$$EC = 8.48 \times 10^4 + 13.2\ Q_{tot} \qquad HR = 0\% \qquad (2.38)$$

$$EC = 8.84 \times 10^4 + 14.6\ Q_{tot} \qquad HR = 35\% \qquad (2.39)$$

$$EC = 8.66 \times 10^4 + 15.8\ Q_{tot} \qquad HR = 50\% \qquad (2.40)$$

$$EC = 8.39 \times 10^4 + 19.2\ Q_{tot} \qquad HR = 70\% \qquad (2.41)$$

A comparison of the thermal, catalytic fixed-bed, and catalytic fluid-bed systems with 50 percent energy recovery is shown in Figure 2.8.

2.5.1.2 Installation Costs

As explained in Section 1, the purchased equipment cost, PEC, is calculated by taking the

sum of the EC and the cost of auxiliary equipment (e.g., ductwork), taxes, freight, and instrumentation. Average values of direct and indirect installation factors [31] to be applied to the PEC are given in Table 2.8 for both recuperative thermal and fixed- and fluid-bed catalytic incinerators.

Table 2.9 shows the itemized installation costs that are obtained when these installation factors are applied to the PECs for the example incinerators. Depending on the site conditions, the installation costs for a given incinerator could deviate significantly from costs generated by these average factors. Vatavuk and Neveril [31] provide some guidelines for adjusting the average installation factors to account for other-than-average installation conditions. For units handling total gas flow rates lower than 20,000 scfm the installation costs are minimal, amounting normally to only utility tie-ins (electrical and, if necessary, combustion or dilution air). The installation costs for these smaller incinerators would be 20 to 25% of the PEC. Smaller units may be installed on the roofs of manufacturing buildings rather than at ground level. In such cases the installation factors could be as high as (or higher than) the factors shown in Table 2.8, even though the units would be "packaged."

2.5.2 Estimating Total Annual Cost

The total annual cost (TAC) is the sum of the direct and indirect annual costs. The TAC for both example systems is given in Table 2.10, alone with suggested factors for calculating them.

2.5.2.1 Direct Annual Costs

Direct annual costs for incinerators include labor (operating and supervisory), maintenance (labor and materials), fuel, electricity, and (in catalytic units) replacement catalyst. For thermal and catalytic units, the fuel usage rate is calculated as shown in Sections 2.3.2 and 2.3.3, respectively where natural gas (methane) is assumed to be the fuel. (Other fuels could be used for thermal units.)

The electricity costs are primarily associated with the fan needed to move the gas through the incinerator. The power (in kilowatts) needed to move a given inlet volumetric flow rate of air (Q_{w_i} per Sections 2.2.3 and 2.3.3) at a total flange-to-flange pressure drop of ΔP inches of water and combined motor/fan efficiency, ϵ, is adapted from Equation **2.7**, as follows:

$$Power_{fan} = \frac{1.17 \times 10^{-4} \, Q_{w_i} \, \Delta P}{\varepsilon} \qquad (2.42)$$

Fan efficiencies vary from 40 to 70 percent [31] while motor efficiencies are usually 90 percent.

The total pressure drop across an incinerator system depends on the number and types of

Table 2.8: Capital Cost Factors for Thermal and Catalytic Incinerators [10]

Cost Item	Factor
Direct Costs	
Purchased equipment costs	
Incinerator (EC) + auxiliary equipment[a]	As estimated, A
Instrumentation[b]	0.10 A
Sales taxes	0.03 A
Freight	0.05 A
Purchased equipment cost, PEC	B = 1.18 A
Direct installation costs	
Foundations & supports	0.08 B
Handling & erection	0.14 B
Electrical	0.04 B
Piping	0.02 B
Insulation for ductwork[c]	0.01 B
Painting	0.01 B
Direct installatoin costs	0.03 B
Site preparation	As required, SP
Buildings	As required, Bldg.
Total Direct Costs, DC	**1.30 B + SP + Bldg.**
Indirect Costs (installation)	
Engineering	0.10 B
Construction and field expenses	0.05 B
Contractor fees	0.10 B
Start-up	0.02 B
Performance test	0.01 B
Contingencies	0.03 B
Total Indirect Costs, IC	0.31 B
Total Capital Investment = DC + IC	**1.61 B + SP + Bldg.**

[a] Ductwork and any other equipment normally not included with unit furnished by incinerator vendor.
[b] Instrumentation and controls often furnished with the incinerator, and those often included in the EC.
[c] If ductwork dimensions have been established, cost may be estimated based on \$10 to \$12/ft^2 of surface for fluid application. (Alternatively, refer to Section 1.2 of this Manual. Fan housing and stacks may also be insulated.

equipment elements included in the system and on design considerations. The estimation of actual pressure drop requirements involves complex calculations based on the specific system's waste gas and flue gas conditions and equipment used. For the purposes of this section, however, the approximate values shown in Table 2.11 can be used.

For the example cases, we will assume 8,000 hours per year operation and a 60% efficiency for the fan and motor together. Using pressure drops of 4 and 8 inches of water, respectively, for the thermal and fluid-bed catalytic incinerators,[3] and adding the pressure drop of 15 inches of water for 70% heat recovery, the fan power requirements can be calculated as follows:

Thermal Incinerator[4]

$$Power_{fan} = \frac{1.17 \times 10^{-4} \left(20{,}900\ acfm\right)\left(19\ inches\ water\right)}{0.60} = 77.4\ kW$$

Catalytic Incinerator

$$Power_{fan} = \frac{1.17 \times 10^{-4} \left(20{,}900\ acfm\right)\left(23\ inches\ water\right)}{0.60} = 93.7\ kW$$

The annual electricity costs would be the products of these usages, the annual operating hours, and the electricity cost ($/kWh), or:

Electricity Cost (Thermal)	=	77.4 kW × 8,000 hours/yr × $0.059/kWh
	=	$36,500 per yr
Electricty Cost (Catalytic)	=	93.7 kW × 8,000 hours/yr × $0.059/kWh
	=	$44,200 per yr

The catalyst replacement costs and scheduling are highly variable and depend on the nature of the catalyst, the amount of "poisons" and particulates in the gas stream (including the auxiliary fuel), the temperature history of the catalyst, and the design of the unit. It is impossible to predict the costs in a general sense. However, noble metal monolith catalysts operating on pure hydrocarbon gases in air will last longer than fluid-bed base metal catalysts operating on chlorinated hydrocarbons in air. Noble metal catalysts are also more expensive than base metal oxide catalysts. The catalyst life for many field units is from 1 to 4 years. The cost, in April 1988 dollars, of the

Table 2.9: Capital Cost Factors for Thermal and Catalytic Incinerator [10] Example Problems

Cost Item	Cost, $	
	Thermal-Recuperative	Fluid-Bed
Catalytic		
Direct Costs		
Purchased equipment costs		
Incinerator (EC)	$254,200	$468,000
Auxiliary equipment	-	-
Sum = A	254,000	468,000
Instrumentation, 0.1A[a]	25,400	46,800
Sales taxes, 0.30 A	7,630	14,000
Freight. 0.05 A	12,700	13,400
Purchased equipment cost, B	$300,000	$552,400
Direct installation costs		
Foundations & supports, 0.08 B	24,000	44,200
Handling & erection, 0.14 B	42,000	77,300
Electrical, 0.04 B	12,000	22,100
Piping, .002 B	6,000	11,000
Insulation for ductwork, 0.01B	3,000	5,520
Painting, 0.01 B	3,000	5,520
Direct installation costs	$90,000	$165,000
Site preparation[a]	-	-
Buildings[a]	-	-
Total Direct	**$390,000**	**$718,000**
Indirect Costs (installation)		
Engineering, .010 B	30,000	55,200
Construction and field expenses, 0.05 B	15,000	27,600
Contractor fees, 0.10 B	30,000	55,200
Start-up, 0.02 B	6,000	11,000
Performance test, 0.01 B	3,000	5,520
Contingencies, 0.03 B	9,000	16,600
Total Indirect Costs	$93,000	$171,100
Total Capital Investment (rounded)	**$483,000**	**$889,000**

[a] None of these items is required.

Table 2.10: Annual Costs for Thermal and Catalytic Incinerators
Example Problem

Cost Item	Suggested Factor	Unit Cost[a]	Thermal	Fluid-Bed Catalyst
Direct Annual Costs[b], DC				
Operating Labor				
Operator	0.5 hr/shift	$12.95/hr.	6,480	6,480
Supervisor	15% of operator	-	972	972
Operating Materials	-			
Maintenance				
Labor	0.5 hr/shift	$14.95/hr.	7,130	7,130
Materials	100% of maintenance labor	-	7,130	7,130
Catalyst replacement	100% of catalyst replaced	$650/ft^3 for metal oxide	0	15,100
Utilities				
Natural Gas -		$3.30/kft^3	264,500	63,400
Electricity	-	$0059/kWh	36,500	44,200
Total DC			**$321,200**	**$144,400**
Indirect Annual Cost, IC				
Overhead	60% of sum of operating, supervisor, & maintenance labor & maintenance materials	-	13,000	17,800
Administrative Charges	2% TCI	-	9,650	17,800
Property Taxes	1% TCI	-	4,830	8,900
Insurance	1% TCI	-	4,830	122,700
Capital recovery[c]	CRF [TCI - 1.08 (cat. Cost)]	-	68,800	122,700
			$101,100	**$171,300**
Total Direct Cost (rounded)			**$422,000**	**$316,000**

[a] 1998 dollars
[b] Assumes 8,000 hr/yr
[c] The capital recovery cost factor, CRF, is a function of the catalyst or equipment life (typically, 2 and 10 years, respectively) and the opportunity cost of the capital (i.e., interest rate). For example, for a 10-year equipment life and a 7% interest rate, CRF = 0.1424

Table 2.11: Typical Pressure Drop Across Selected Equipment

Equipment Type	Energy Recovery, %	ΔP, in. H_2O
Thermal Incinerators	0	4
Catalytic Fixed-bed Incinerator	0	6
Catalytic Fluid-bed Incinerator	0	6-10
Heat Exchangers	35	4
Heat Exchangers	50	8
Heat Exchangers	70	15

replacement catalyst must be obtained from the vendor, but it may be estimated at $3,000/ft^3 for noble metal catalysts and $650/ft^3 for base metal oxide catalysts. For the example case, the catalyst is a base metal oxide because the waste gas contains a chlorinated compound. We will assume a two year catalyst life. Knowing that the catalyst volume is 39 ft^3 and using a cost of $650/ft^3 and a capital recovery factor of 0.5531 (2-year life at a 7% interest rate), the annual expense for catalyst replacement is

$$Power_{fan} = \frac{1.17 \times 10^{-4} \left(20{,}900\ acfm\right)\left(23\ inches\ water\right)}{0.60} = 93.7\ kW$$

(The " 1.08" factor covers the freight and sales tax for the replacement catalyst.)

To calculate the fuel or electricity annual cost, multiply the fuel usage rate (scfm) or the electricity usage rate (kW) by the total hours per year of operation (e.g., 333 d/yr x 24 h/d = 8,000 h/yr) and by the appropriate unit cost (e.g., $/scfm for fuel and $/kWh for electricity).

For the example cases, the fuel costs can be calculated from the fuel usage rates and the natural gas unit cost of $0.00330/scf. For the thermal incinerator example, the annual fuel cost is calculated as follows:

$$\text{Annual Fuel} = 0.00330\ \$/scf \times 167\ scf/min \times 60\ min/hr \times 8{,}000\ hr/yr$$

[3] A fluid-bed catalytic incinerator is used because the waste gas contains a chlorinated compound which would poison the catalyst in a fixed-bed incinerator.

[4] Computed from inlet waste-gas flow rate (20,000 scfm) at preheater inlet temperature (100°F).

Cost, Thermal = $264,500 per year

For the catalytic incinerator example. the annual fuel cost is found similarly:

Annual Fuel Cost, Catalytic = $63,400 per year

Operating and maintenance labor are estimated as 0.5 hours per 8-hour shift each, supervisory labor at 15% of operating labor, and maintenance material as 100% of maintenance labor.

2.5.2.2 Indirect Annual Costs

The indirect (fixed) annual costs include capital recovery, overhead, and property taxes, insurance, and administrative (G&A) charges. The last three of these can be estimated at 1%, 1%, and 2% of the total capital investment, respectively. The system capital recovery cost is based on an estimated 10-year equipment life. (See Section 1.2 for a thorough discussion of the capital recovery cost and the variables that determine it.) The system capital recovery cost is the product of the system capital recovery factor (CRF) and the total capital investment (TCI) less the purchased cost of the catalyst (C_{cat} x 1.08 where the 1.08 is for freight and sales tax). These values calculated for the example cases are given in Table 2.47.

2.5.3 Cost Comparison for Example Case

The example VOC stream defined in Section 2.2.4.1 serves to illustrate some typical characteristics of thermal and catalytic systems. The total annual costs shown in Table 2.47 show that the catalytic system's auxiliary fuel costs are significantly lower than those of the thermal unit. The disparity is enough to offset the higher capital costs of the catalytic incinerator over the assumed 10-year lifetime of the units. Two factors that should be noted in the comparison of these two systems are (1) the 98 percent level of destruction met by the thermal incinerator may be difficult to reach by the catalytic system (this may be important in some cases), and (2) the example waste stream is of particularly low heating value (4 Btu/scf) which favors the catalytic system due to the lower auxiliary fuel requirements.

2.6 Acknowledgements

The authors gratefully acknowledge the following companies for contributing data to this chapter:

- Peabody Engineering (Stamford, CT)
- Combustion Engineering - Air Preheater, Inc. (Wellsville, NY)
- TEC Systems, Inc. (DePere, WI)
- Air Research, Inc. (ARI) (Palatine, IL)
- Energy Development Associates (EDA) (Itasca, IL)
- Pillar Technologies, Inc. (Hartland, WI)
- Huntington Energy Systems, Inc. (Union, NJ)
- Regenerative Environmental Equipment Co. (REECO) (Morris Plains, NJ)
- Englehard Corp. (Edison, NJ)

References

[1] *Prudent Practices for Disposal of Chemicals from Laboratories*, National Academy Press, Washington, D.C., 1983.

[2] Memorandum from Mascone, D.C., EPA, OAQPS, to Farmer, J. R., OAQPS, EPA, June 11, 1980, Thermal Incinerator Performance for NSPS.

[3] Memorandum from Mascone, D.C., EPA, OAQPS, to Farmer, J. R., OAQPS, EPA, July 22, 1980, Thermal Incinerator Performance for NSPS, Addendum.

[4] Memorandum from Mascone, D.C., EPA, OAQPS, to Farmer, J. R., OAQPS, EPA, August 22, 1980, Thermal Incinerators and Flares.

[4] Letter from Thomas Schmidt (ARI International, Palatine, IL) to William M. Vatavuk (EPA, OAQPS, Research Triangle Park, NC), August 16, 1989.

[5] Weldon, J. and S. M. Senkan, *Combustion Sci. Technol.*, 1986, 47.

[6] Manning, P., *Hazard Waste*, 1984, 1(1).

[7] Pope, D., Walker, D. S., Moss, R. L., *Atmos. Environ.*, 1976, 10.

[8] Musick, J. K., and F. W. Williams, *Ind. Eng. Chem. Prod. Res. Dev.*, 1974, 13(3).

[9] Letter from Robert M. Yarrington (Englehard Corporation, Edison, NJ) to William M. Vatavuk (EPA, OAQPS, Research Triangle Park, NC), August 14, 1989.

[10] Personal Communication from Bill Sheffer (ARI, Inc., Palatine, IL) to Donald R. van der Vaart (RTI, Research Triangle Park, NC), March 30, 1988.

[11] Personal Communication from Ralph Stettenbenz (Combustion Engineering, Air Preheater, Inc., Wellsville, NY) to Donald R. van der Vaart (RTI, Research Triangle Park, NC), March 23, 1988.

[12] Grelecki, C., *Fundamentals of Fire and Explosion Hazards Evaluation*, AIChE Today Series, New York, 1976.

[13] Weast, R.C. (ed.), *CRC Handbook of Chemistry and Physics*, 49th ed., CRC Press, Cleveland, Ohio, 1968.

[14] Perry, R. H. and C. H. Chilton(eds.), *Chemical Engineers Handbook*, 5th ed., McGraw-Hill, New York, NY 1973.

[15] Personal Communication from Robert Yarrington (Englehard Corp., Edison, NJ) to William M. Vatavuk (EPA, OAQPS, Research Triangle Park, NC), June 6, 1989.

[16] Personal communication from Gary Masonick (Goal Line Environmental Technolgies) to Janet D. Groeber (SAIC. Cincinnati, OH), September 29,1999.

[17] Personal communication from Gary Raemheld (Goenergy, Kent, WA) to Janet D. Groeber (SAIC. Cincinnati, OH), September 28,1999.

[18] Personal communication from John Wilke (CVM Corp., Wilmington, DE) to Janet D. Groeber (SAIC. Cincinnati, OH), September 29,1999.

[19] Personal Communication from Thomas Schmidt (ARI International, Palatine, IL) to William M. Vatavuk (EPA, OAQPS, Research Triangle Park, NC), June 7, 1989.

[20] Githens, R. E. and D. M. Sowards, *Catalytic Oxidation of Hydrocarbon Fumes*, PB-299 132, National Technical Information Service, Springfield, VA.

[21] Personal Communication from Andrew Jones (Energy Development Associates, Itasca, IL) to Donald R. van der Vaart (RTI, Research Triangle Park, NC), March 4, 1988.

[22] Personal Communication from C. L. Bumford (Peabody Engineering, Stamford, CT) to Donald R. van der Vaart (RTI, Research Triangle Park, NC), March 28, 1988.

[23] Personal Communication from C. M. Martinson (TEC Systems, DePere, WI) to Donald R. van der Vaart (RTI, Research Triangle Park, NC), March 28, 1988.

[24] Personal Communication from Ronald J. Renko (Huntington Energy Systems, Inc., Union, NJ) to Donald R. van der Vaart (RTI, Research Triangle Park, NC), March 16, 1988.

[25] Personal Communication from James H. Mueller (Regenerative Environmental Equipment Co., Inc., Morris Plains, NJ) to Donald R. van der Vaart (RTI, Research Triangle Park, NC), January 13, 1988.

[26] Personal Communication from Robert Hablewitz (Pillar Technologies, Hartland, WI) to Donald R. van der Vaart (RTI, Research Triangle Park, NC), March 20, 1988.

[27] Kobe, K. A. and associates, "Thermochemistry for the Petrochemical Industry", *Petroleum Refiner*, Jan. 1949 through Nov. 1954.

[28] Letter from William M. Vatavuk (EPA, OAQPS, Research Triangle Park, NC) to Gerald Schrubba (Salem Engelhard, South Lyon, MI), September 22, 1992.

[29] Letter from William M. Vatavuk (EPA, OAQPS, Research Triangle Park, NC) to Rod Pennington (REECO/Research Cottrell, Somerville, NJ), September 22, 1992.

Appendix A

Properties of Selected Compounds

Table 2.12: Limits of Flammability of Combustible Organic Compounds in Air at Atmospheric Pressure, Room Temperature [79]

Compound	Molecular Weight	LEL[a] (volume %)	UEL[b] (volume %)
Methane	16.04	5.00	15.00
Ethane	30.07	3.00	12.50
Propane	44.09	2.12	9.35
Butane	58.12	1.86	8.41
Pentane	72.15	1.40	7.80
Hexane	86.17	1.18	7.40
Octane	114.23	0.95	
Nonane	128.25	0.83	
Decane	142.28	0.77	
Ethylene	28.05	2.75	28.60
Propylene	42.08	2.00	11.10
Acetylene	26.04	2.50	80.00
Cyclohexane	84.16	1.26	7.75
Benzene	78.11	1.40	7.10
Toluene	92.13	1.27	6.75

[a] Lower Explosive Limit
[b] Upper Explosive Limit

Table 2.13: Molar Heat Capabilities of Gases at Zero Pressure [25]

$$C_p = a + bT + cT^2 + dT^3 \; ; \; T \text{ in } °K$$

$$C_{pm} = \frac{\int_{T_1}^{T_2} C_p dT}{(T_2 - T_1)}$$

C_p *in calories/g-moles* °K *Btu/lb-mole* °R

Compound	a	b x 10^2	c x 10^6	d x 10^{10}	Temperature Range (°K)
Methane	4.750	1.200	0.3030	-2.630	273-1500
Ethane	1.648	4.124	-1.530	1.740	273-1500
Propane	-0.966	7.279	-3.755	7.580	273-1500
Butane	0.945	8.873	-4.380	8.360	273-1500
Pentane	1.618	10.85	-5.365	10.10	273-1500
Hexane	1.657	13.19	-6.844	13.78	273-1500
Cyclopentane	-12.957	13.087	-7.447	16.41	273-1500
Cyclohexane	-15.935	16.454	-9.203	19.27	273-1500
Benzene	-8.650	11.578	-7.540	18.54	273-1500
Toluene	-8.213	13.357	-8.230	19.20	273-1500
Nitrogen	6.903	-0.037553	0.1930	-0.6861	273-1500
Oxygen	6.085	0.3631	-0.1709	0.3133	273-1500
Air	6.713	0.04697	0.1147	-0.4696	273-1500
Carbon Dioxide	5.316	1.4285	-0.8362	1.784	273-1500

Table 2.14: Heats of Combustion of Selected Gaseous Organic Compounds [80]

Compound	Molecular Weight	(cal/g)	$-\Delta h_c$	(Btu per lb)
Methane	16.04	11,953.3		21,502
Ethane	30.07	11,349.6		20,416
Propane	44.09	11,079.2		19,929
Butane	58.12	10,932.3		19,665
Pentane	72.15	10,839.7		19,499
Hexane	86.17	10,780.0		19,391
Octane	114.23	10,737.2		19,256
Nonane	128.25	10,680.0		19,211
Decane	142.28	10,659.7		19,175
Ethylene	28.05	11,271.7		20,276
Propylene	42.08	10,942.3		19,683
Acetylene	26.04	10,476.7		19,001
Cyclohexane	84.16	9,698.4		19,846
Benzene	78.11	9,698.4		17,446
Toluene	92.13	9,784.7		17,601

Appendix B

Design Procedure for Non-Recuperative Thermal Incinerators

Not all thermal incinerators are equipped with recuperative heat exchangers to transfer energy from the flue gas stream to the incoming waste gas stream. These non-recuperative units use other mechanisms to recovery flue gas energy. One of these types is the *regenerative* incinerator. As discussed in a subsection of Section 2.1 entitled Regenerative Incinerators, a regenerative incinerator accomplishes energy recovery by conveying the flue gas through a ceramic bed which captures a portion of the stream's enthalpy. After a switching mechanism is engaged, the incoming waste gas passes through this hot bed and is warmed to its ignition temperature. This process is illustrated in Figure 2.2.

While we can determine the stream inlet and outlet temperatures for a recuperative heat exchanger fairly accurately, we cannot always do so for a regenerative incinerator bed. For one thing, these beds do not behave like typical heat exchangers. The bed temperature profiles are often difficult to predict. More importantly, because regenerative incinerators do not operate at steady state conditions, the temperatures within the beds and many other parts of the unit vary with time. For that reason, it is more convenient to view the entire regenerative incinerator as a "black box" into which waste gas and auxiliary fuel flow and from which flue gas emanates. Around this black box we may make mass and energy balances. In this way, we need not make any assumptions about what occurs inside the incinerator regarding temperatures, flowrates, or other stream parameters.

However, to determine the auxiliary fuel requirement for regenerative incinerators via the procedure shown in this appendix we have to make two key assumptions, viz.: (1) the temperatures and flowrates of all streams entering and leaving the incinerator are at steady state and (2) the combustion temperature (and by inference, the heat loss fraction) are constant as well. The other assumptions will be addressed in the following design steps:

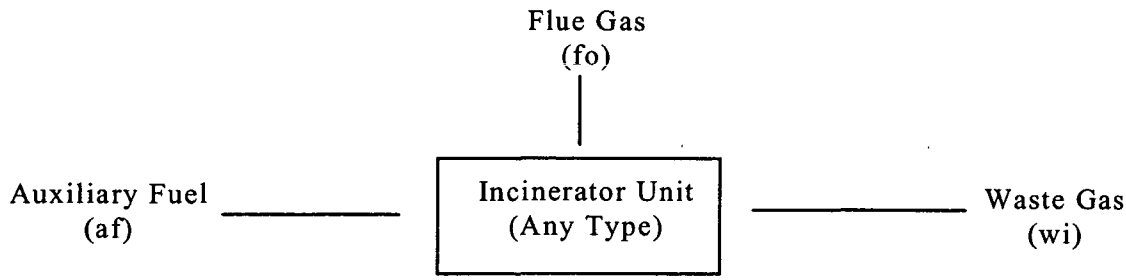
Steps 1 to 4: These are the same as those for thermal recuperative and catalytic incinerators. (See Section 2.4.1.)

Step 5t - Establish the incinerator operating temperature: Because their designs are more resistant to thermal stresses and because they can achieve very high heat recoveries, regenerative incinerators are usually operated at higher temperatures than recuperative units. Consequently, higher VOC destruction efficiencies are achieved. Operating temperatures of 1800 to 2000° F are typical.

Step 6t - Calculate the waste gas temperature at the exit of the preheater: As explained above, regenerative incinerators do not employ preheaters. The preheating is done by and within the ceramic beds. Moreover, because the mass and energy balances are made around the entire

unit, we do not need to know the temperature of the preheated waste gas to calculate the auxiliary fuel requirement.

Step 7t - Calculate the auxiliary fuel requirement, Q_{af}: Because a regenerative incinerator recovers nearly all of the energy from the combustion (flue) gas, its auxiliary fuel requirement is



usually lower than that for a recuperative incinerator. However, as discussed above, this fuel requirement is determined via mass and energy balances taken around the entire unit, not just the combustion chamber. Consider the following diagram:

Taking mass and energy balances around the incinerator, we obtain:

Mass balance:

Mass in = Mass out

Mass fuel + Mass waste gas = Mass flue gas

$$\rho_{af}Q_{af} + \rho_{wi}Q_{wi} = \rho_{fo}Q_{fo}$$

Energy balance:

Next, we take an energy balance around the incinerator unit:

Energy in - Energy out + Energy generated = 0

The terms of the energy balance equation are the inlet waste gas and outlet flue gas enthalpies (H_{wi} and H_{fo}, respectively), the energy loss (H_L), and the waste gas VOC and fuel (natural gas) heat contents (H_{cwi} and H_{caf}, in turn):

$$H_{wi} - (H_{fo} + H_L) + (H_{cwi} + H_{caf}) = 0$$

The variables comprising each of the terms in this energy balance equation are listed in Table2.6. They are:

$$H_{wi} = \rho_{wi} Q_{wi} C_{pmwi} (T_{wi} - T_{ref})$$

$$H_{fo} = \rho_{fo} Q_{fo} C_{pmfo} (T_{fo} - T_{ref})$$

$$H_L = \eta \rho_{fi} Q_{fi} C_{pmfi} (T_{fi} - T_{ref})$$

$$H_{cwi} = \rho_{wi} Q_{wi} (-\Delta h_{cwi})$$

$$H_{caf} = \rho_{af} Q_{af} (-\Delta h_{caf})$$

where

η	=	energy loss from combustion chamber (fractional)
T_{fi}	=	combustion temperature (°F)

We next substitute these variables into Equation 2.51 and solve for the fuel mass rate ($\rho_{af} Q_{af}$). When doing so, we make the following assumptions:

- The streams flowing to and from the incinerator are at steady state conditions.
- The auxiliary air requirements are zero.
- The ambient, reference, and fuel inlet temperatures are equal (77°F). (This assumption results in the inlet fuel stream having a zero enthalpy.)
- The heat capacities of the gas streams to and from the unit are approximately the same, regardless of composition.

- The *mean* heat capacities of the streams above the reference temperature (77°F) are approximately equal, regardless of *temperature*. Further, the mean heat capacity of the waste gas/flue gas stream entering/leaving the incinerator is evaluated at the average of the inlet (T_{wi}) and combustion (T_{fi}) temperatures. That is, $C_{pmwi} = C_{pmfi} = C_{pmfo} = C_{pm}$.

When we do all this, we get the following expression:

$$\underbrace{\rho_{wi}Q_{wi}C_{pm}(T_{wi} - T_{ref})}_{\text{(Energy in)}} - [\underbrace{\rho_{fo}Q_{fo}C_{pm}(T_{fo} - T_{ref})}_{\text{(Energy out)}} + \eta\rho_{fi}Q_{fi}C_{pm}(T_{fi} - T_{ref})] +$$

$$[\underbrace{\rho_{wi}Q_{wi}(-\Delta h_{cwi}) + \rho_{af}Q_{af}(-\Delta h_{caf})}_{\text{(Energy generated)}}] = 0$$

Substitution for $\rho_{fo}Q_{fo}$ per Equation 3.357 above yields:

$$[\rho_{wi}Q_{wi}C_{pm}(T_{wi}-T_{ref})] - [\eta C_{pm}(\rho_{af}Q_{af} + \rho_{wi}Q_{wi})(T_{fi}-T_{ref}) +$$
$$C_{pm}(\rho_{af}Q_{af} + \rho_{wi}Q_{wi})(T_{fo}-T_{ref})] + [\rho_{wi}Q_{wi}(-\Delta h_{cwi}) + \rho_{af}Q_{af}(-\Delta h_{caf})] = 0$$

Finally, solving for $\rho_{af}Q_{af}$, the auxiliary fuel mass rate (lb/min):

$$\rho_{wi}Q_{wi}\{C_{pm}[\eta(T_{fi} - T_{ref}) + (T_{fo} - T_{wi})] - (-\Delta h_{cwi})\}\rho_{af}Q_{af}$$
$$= \{(-\Delta h_{caf}) - C_{pm}[\eta(T_{fi} - T_{ref}) + (T_{fo} - T_{ref})]\}$$

Equation 3.361 provides the auxiliary fuel requirement for any type of thermal incinerator, as it is independent of any intermediate variables, such as the temperature of the preheated waste gas. Clearly, this equation can be used with regenerative incinerators, as long as the above-stated assumptions hold.

The heat loss fraction (η) will vary according to the incinerator type, how the incinerator components are configured in the unit, the construction materials, the type and amount of insulation, and other factors. For instance, for recuperative incinerators, η is approximately 0.10. The for regenerative incinerators is considerably lower, however. There are two reasons for this. First, the components of a regenerative incinerator—combustion chamber, ceramic beds, etc.—are housed in a single enclosure, while in a recuperative incinerator the combustion chamber, heat exchanger, and interconnecting ductwork are housed separately, thus offering more heat transfer area. Second, because regenerative units are lined with ceramic, they are better insulated than recuperative incinerators.

To gain an estimate of this heat loss fraction, we contacted two regenerative incinerator vendors. [94,95] Based on the heat loss data that they supplied, we calculated values ranging from **0.002 to 0.015 (0.2 to 1.5%)**. These values varied according to the incinerator configuration (vertical or horizontal), the waste gas flow rate, the ambient temperature, and the wind speed.

Step 8t - Verify that the auxiliary fuel requirement is sufficient to stabilize the burner flame: As explained in Section 2.4.2, only a small amount (< 5% of the total energy input) is needed to stabilize the burner flame. With recuperative incinerators, the auxiliary fuel requirement is usually much larger than the burner stabilization requirement, so that this constraint rarely comes into play. With regenerative incinerators, however, the auxiliary fuel requirement may be as low or lower than the fuel needed to stabilize the burner. Therefore, it is important to compare these two requirements. This comparison is made via Equations 3.326 and 3.327. If the auxiliary fuel is less, the minimum fuel requirement would be set at 5% of the total energy input.

Step 9t - Calculate the flue gas volumetric flow rate, Q_{fi}: As with thermal recuperative incinerators, the regenerative incinerator flue gas flow rate is the rate used to size and cost the unit. Measured at standard conditions (1 atmosphere and 77°F), Q_{fi} is the sum of the inlet waste gas (Q_{wi}) and fuel (Q_{af}) flow rates. But since Q_{af} for regenerative units is small compared to Q_{wi}, the waste gas and flue gas flows should be virtually identical.

EPA/452/B-02-001

Section 4

NO_x Controls

EPA/452/B-02-001

Section 4.2

NO_x Post- Combustion

Introduction

Nitrogen oxides (NO_x) are gaseous pollutants that are primarily formed through combustion process. While flue gas is within the combustion unit, about 95% of the NO_x exists in the form of nitric oxide (NO). The balance is nitrogen dioxide (NO_2), which is unstable at high temperatures. Once the flue gas is emitted into the atmosphere, most of the NO_x is ultimately converted to NO_2. NO_x in the atmosphere reacts in the presence of sunlight to form ozone (O_3), one of the criteria pollutants for which health-based National Ambient Air Quality Standards have been established. Since ozone formation requires sunlight and high temperatures, ozone formation is greatest in summer months.

NO_x is generated in one of three forms; fuel NO_x, thermal NO_x, and prompt NO_x. Fuel NO_x is produced by oxidation of nitrogen in the fuel source. Combustion of fuels with high nitrogen content such as coal and residual oils produces greater amounts of NO_x than those with low nitrogen content such as distillate oil and natural gas. Thermal NO_x is formed by the fixation of molecular nitrogen and oxygen at temperatures greater than 3600°F (2000°C). Prompt NO_x forms from the oxidation of hydrocarbon radicals near the combustion flame and produces an insignificant amount of NO_x.

Selective Noncatalytic Reduction (SNCR) and Selective Catalytic Reduction (SCR) are post-combustion control technologies based on the chemical reduction of nitrogen oxides (NO_x) into molecular nitrogen (N_2) and water vapor (H_2O). The primary difference between the two technologies is that SCR utilizes a catalyst to increase the NO_x removal efficiency, which allows the process to occur at lower temperatures. The technologies can be used separately or in combination with other NO_x combustion control technologies such as low NO_x burners (LNB) and natural gas reburn (NGR). SNCR and SCR can be designed to provide NO_x reductions year-round or only during summer months, when ozone concerns are greatest.

This section presents design specifications and a costing methodology for SNCR and SCR applications for large industrial boilers (greater than 250 MMBtu/hr). Its primary purpose is to provide study-level cost estimates. These estimates can be used to compare the approximate costs of SNCR, SCR, and alternative NO_x control technologies. The costing methodologies are based on cost estimations for SNCR and SCR applications to utility boilers, which were developed by the Office of Research and Development (ORD), U.S. Environmental Protection Agency (EPA), Research Triangle Park, NC.

As a prelude to the cost methodology, the section describes the process chemistry, performance parameters, and system components of SNCR and SCR. In addition, impacts to the boiler performance and facility operations resulting from the installation of SNCR and SCR are presented. The section also estimates important underlying design parameters including the normalized stoichiometric ratio, catalyst volume, and reagent consumption. Lastly, it presents

assumptions and equations for estimating capital costs, annual operation and maintenance costs, and annualized costs. SNCR and SCR are discussed in separate chapters, however, the chapter on SCR builds on the concepts discussed in the SNCR chapter.

Information on key aspects of the design of SNCR and SCR systems is considered proprietary by vendors, including methods for estimating certain design parameters and costs. This information is not available to develop cost methodologies for SNCR and SCR. In order to obtain estimates of the proprietary design parameters and costs, it is necessary to develop mathematical correlations from available data using regression and curve fitting techniques. These expressions are derived in EPA reports prepared by The Cadmus Group, Bechtel Power, Inc. and SAIC in SNCR and SCR References [1, 2 and 3] from documented comprehensive SNCR and SCR performance data and costs based on quotations provided by suppliers and facilities.

EPA/452/B-02-001

Chapter 1

Selective Noncatalytic Reduction

Daniel C. Mussatti
Innovative Strategies and Economics Group
Air Quality Strategies and Standards Division
Office of Air Quality Planning and Standards
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

Dr. Ravi Srivastava
Office of Research and Development
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

Paula M. Hemmer and
Randy Strait
E. H. Pechan & Associates, Inc.
3622 Lyckan Parkway, Suite 2002
Durham, NC 27707

October 2000

Contents

1.1 Introduction 1-3

1.2 Process Description 1-4
1.2.1 Reduction Chemistry 1-5
1.2.2 Reagents 1-5
1.2.3 SNCR Performance Parameters 1-7
1.2.4 SNCR System 1-13
1.2.5 Other Considerations 1-19

1.3 Design Parameters 1-21
1.3.1 Design Parameters for Study Level Estimates 1-21
1.3.2 Design Parameters for Detailed/Performance Specifications 1-27

1.4 Cost Analysis 1-30
1.4.1 Total Capital Investment 1-31
1.4.2 Total Annual Costs 1-33

1.5 Example Problem 1-39
1.5.1 Design Parameter Example 1-40
1.5.2 Cost Estimation Example 1-42

References 1-45

1.1 Introduction

SNCR is currently being used for NO_x emission control on industrial boilers, electric utility steam generators, thermal incinerators, and municipal solid waste energy recovery facilities. Its use on utility boilers has generally been limited to units with output of less than 3,100 MMBtu, but large-scale applications of SNCR have been successfully tested such as Cardinal Plant Unit 1, a 600 MW (5,700 MMBtu/hr) utility boiler located in Brilliant, Ohio [11]. SNCR can be applied as a stand-alone NO_x control or with other technologies such as combustion controls. The SNCR system can be designed for seasonal or year-round operations.

SNCR can achieve NO_x reduction efficiencies of up to 75 percent (%) in selected short-term demonstrations [5]. In typical field applications, however, it provides 30% to 50% NO_x reduction. Reductions of up to 65% have been reported for some field applications of SNCR in tandem with combustion control equipment such as low NO_x burners (LNB) [11]. SNCR systems applied to large combustion units (greater than 3,000 MMBtu/hr) typically have lower NO_x reduction efficiencies (less than 40%), due to mixing limitations [11]. Figure 1.1 presents the NO_x reduction efficiency of SNCR for various utility boiler sizes.

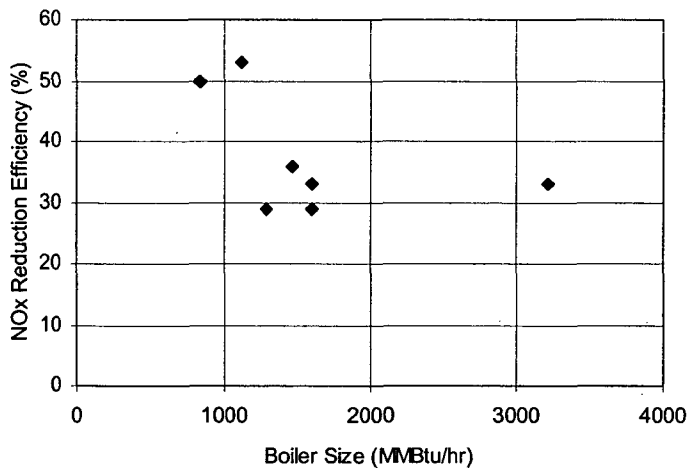


Figure 1.1: SNCR NO_x Reduction Efficiency for Various Boiler Sizes [8]

The hardware associated with an SNCR installation is relatively simple and readily available. Consequently, SNCR applications tend to have low capital costs compared to LNB and SCR. Installation of SNCR equipment requires minimum downtime. Though simple in concept, it is challenging in practice to design an SNCR system that is reliable, economical, simple to control, and meets other technical, environmental, and regulatory criteria. Practical application of SNCR is limited by the boiler design and operating conditions.

The costing algorithms in this report are based on retrofit applications of SNCR to existing coal-fired, dry bottom, wall-fired and tangential, balanced draft boilers. There is little difference between the cost of SNCR retrofit of an existing boiler and SNCR installation on a new boiler [10]. Therefore, the cost estimating procedure is suitable for retrofit or new boiler applications of SNCR on all types of coal-fired electric utilities and large industrial boilers.

1.2 Process Description

SNCR is based on the chemical reduction of the NO_x molecule into molecular nitrogen (N_2) and water vapor (H_2O). A nitrogen based reducing agent (reagent), such as ammonia or urea, is injected into the post combustion flue gas. The reagent can react with a number of flue gas components. However, the NO_x reduction reaction is favored over other chemical reaction processes for a specific temperature range and in the presence of oxygen, therefore, it is considered a selective chemical process.

The SNCR process occurs within the combustion unit which acts as the reaction chamber. Figure 1.2 shows a schematic of a boiler interior with SNCR. Reagent is injected into the flue gas through nozzles mounted on the wall of the combustion unit. The injection nozzles are generally located in the post-combustion area, the upper area of the furnace and convective passes. The injection causes mixing of the reagent and flue gas. The heat of the boiler provides the energy for the reduction reaction. The NO_x molecules are reduced and the reacted flue gas then passes out of the boiler. More detail on the SNCR process and equipment is provided in he following sections.

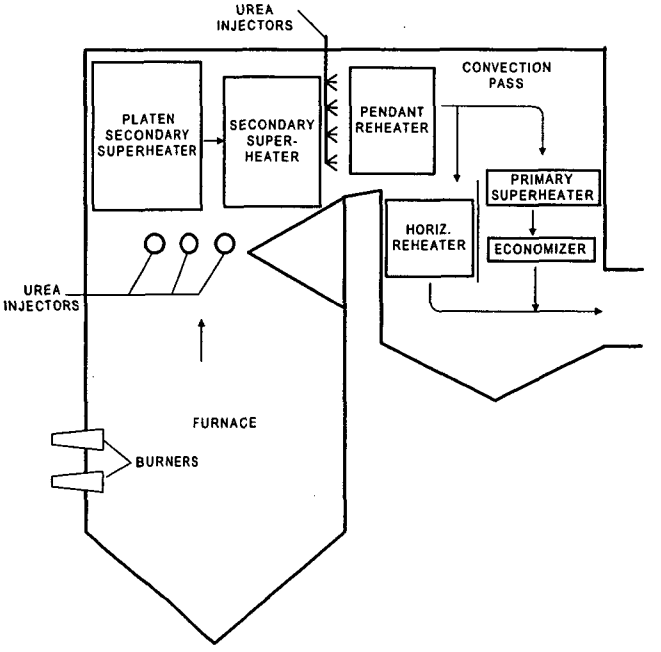


Figure 1.2: Boiler Gas Path Configuration

1.2.1 Reduction Chemistry

SNCR is a relatively simple chemical process. The process begins with an ammonia-based reagent, ammonia (NH_3) or urea ($CO(NH_2)_2$), being vaporized either before injection by a vaporizer or after injection by the heat of the boiler. Within the appropriate temperature range, the gas-phase urea or ammonia then decomposes into free radicals including NH_3 and NH_2. After a series of reactions, the ammonia radicals come into contact with the NO_x and reduce it to N_2 and H_2O. The global representation of these reactions is given below for both urea and ammonia. Note that the NO_x is represented as NO since it is the predominant form of NO_x within the boiler. The ammonia reaction equation is given by:

$$2NO + 2NH_3 + \tfrac{1}{2}O_2 \rightarrow 2N_2 + 3H_2O \qquad (1.1)$$

The reaction for urea is given by:

$$2NO + CO(NH_2)_2 + \tfrac{1}{2}O_2 \rightarrow 2N_2 + CO_2 + 2H_2O \qquad (1.2)$$

The primary byproduct formed during either ammonia or urea SNCR is nitrous oxide (N_2O). N_2O is an ozone depletor and a greenhouse gas but is currently not regulated. Urea based reduction generates significantly more N_2O than ammonia-based systems. Up to 30% of the NO_x can be transformed into N_2O. Proprietary additives are available for the urea-based SNCR process to reduce the formation of N_2O. [10]

1.2.2 Reagents

Both ammonia and urea have been successfully employed as reagents. The cost of the reagent is a large portion of the annual costs of operating an SNCR system. Ammonia is generally less expensive than urea. However, the choice of reagent is based not only on cost but on physical properties and operational considerations. The properties of urea and ammonia in aqueous solutions are shown in Table 1.1.

Ammonia can be utilized in either aqueous or anhydrous form. Anhydrous ammonia is a gas at normal atmospheric temperature. It must be transported and stored under pressure which presents safety issues. Aqueous ammonia is generally transported and stored at a concentration of 29.4% ammonia in water. At concentrations above 28%, storage of ammonia may require a permit, therefore some recent applications of SNCR are using a 19% solution [1]. Decreasing the concentration, however, increases the required storage volume. Ammonia is generally injected as a vapor. Providing sufficient ammonia vapor to the injectors requires a vaporizer, even though the 29.4% solution has substantial vapor pressure at normal air temperatures. The injection system equipment for vapor systems is more complicated and expensive than equipment for aqueous systems (see Section 1.2.4 SNCR System)

Table 1.1: Urea and Ammonia Reagent Properties, [6]

Property	Urea Solution	Aqueous Ammonia
Chemical formula	$CO(NH_2)_2$	NH_3
Molecular Weight of reagent	60.06	17.03
Liquid or gas at normal air temperature	Liquid	Liquid
Concentration of reagent normally supplied	50% by weight	29.4% by weight
Ratio of NH_3 to solution	28.3% by weight of NH_3	29.4% by weight of NH_3
Density of solution @ 60° F	71 lb/ft³	56 lb/ft³ (58 lb/ft³ for 19%)
Vapor pressure @ 80° F	< 1 psia	13.9 psia [8]
Crystallization temperature	64° F	- 110° F
Flamability limits in air	Non-flammable	Lower explosion limit = 16% NH_3 by volume; Upper explosion limit = 25% NH_3 by volume.
Threshold limit value (health effects)	Not Specified	25 ppm
Odor	Slight (ammonia-like)	Pungent odor @ 5 ppm or more
Acceptable materials for storage	Plastic, steel, or stainless steel (no copper or copper-based alloys or zinc/aluminum fittings)	Steel tank, capable of handling at least 25 psig pressure (no copper or copper-based alloys, etc.)

Urea is generally utilized in a 50% aqueous solution [1]. At this concentration, the urea solution must be heated and circulated in cold climates due to its low freezing point, 17.8EC (64EF). Higher-concentrations of urea solutions are available which decrease the storage volume but require extensive heating to prevent freezing. Urea is injected into the boiler as an aqueous solution and vaporized by the heat of the boiler. Urea can also be transported in pellet form which minimizes transportation requirements. However, to produce aqueous urea for use in the SNCR system, the urea must then be mixed with water at the facility. This mixing process is generally cost prohibitive except for remote sites, large facilities, or facilities where chemical mixing processes are already being performed [13].

Urea based systems have several advantages over ammonia based systems. Urea is a nontoxic, less volatile liquid that can be stored and handled more safely than ammonia. Urea solution droplets can penetrate farther into the flue gas when injected into the boiler. This enhances mixing with the flue gas which is difficult on large boilers [1]. Because of these advantages, urea is more commonly used than ammonia in large boiler applications of SNCR systems.

1.2.3 SNCR Performance Parameters

The rate of the reduction reaction determines the amount of NO_x removed from the flue gas. The important design and operational factors that affect the reduction of NO_x by an SNCR system include:

- Reaction temperature range;
- Residence time available in the optimum temperature range;
- Degree of mixing between the injected reagent and the combustion gases
- Uncontrolled NO_x concentration level;
- Molar ratio of injected reagent to uncontrolled NO_x; and
- Ammonia slip.

Temperature

The NO_x reduction reaction occurs within a specific temperature range where adequate heat is available to drive the reaction. At lower temperatures the reaction kinetics are slow and ammonia passes through the boiler (ammonia slip). At higher temperatures the reagent oxidizes and additional NO_x is generated. The temperature window is dependant on the reagent utilized. Figure 1.3 shows the NO_x reduction efficiency for urea and ammonia SNCR at various boiler temperatures. For ammonia, the optimum temperature is from 870EC to 1100EC (1,600EF to

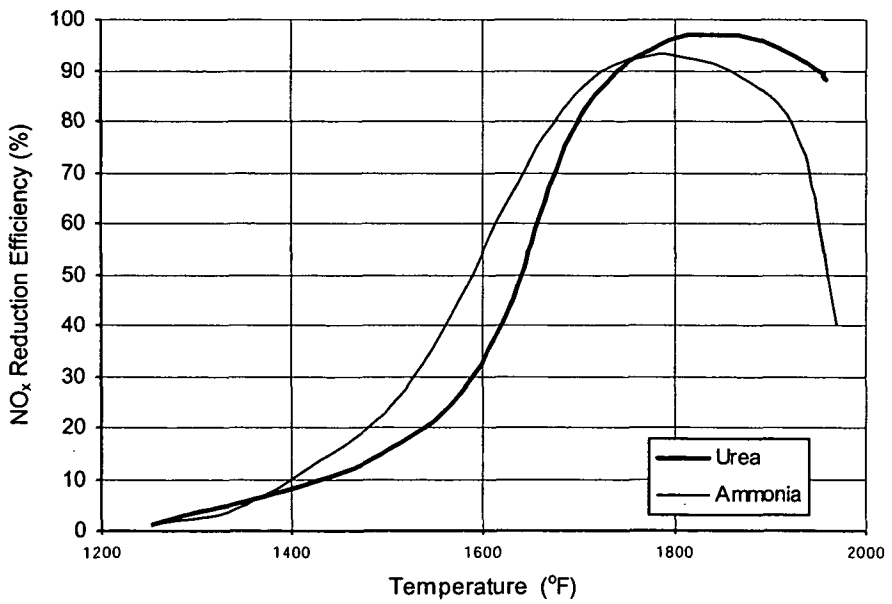


Figure 1.3: Effect of Temperature on NO_x Reduction

2,000EF) [1]. Its optimum temperature range can be reduced by injecting hydrogen gas with the ammonia. For urea, the optimum temperature range is between 900EC to 1150EC (1,650EF to 2,100EF) [1]. Additives to the urea reagent may be used to widen the effective temperature window. These additives are generally proprietary (see Section 1.2.4 Other Considerations).

The reagent is injected into the boiler in regions where the combustion gas temperature is within the specified range. Since reaction temperatures are high, the injection takes place in the boiler itself. In general, the reagent is injected within the boiler superheater and reheater radiant and convective regions, where the appropriate temperature range is typically available [1]. Proper placement of the injection ports results in higher NO_x reduction efficiency.

Flue gas temperature within the boiler depends on the boiler design and operating conditions. These are generally set to meet steam generation requirements and are not always ideal for the SNCR process. Flue gas temperatures in the upper furnace through the convective pass may vary by ±150EC (300EF) from one boiler to the next [1]. In addition, fluctuations in the boiler load profile affect the temperature within the boiler. At lower load profiles, the temperature within the boiler is lower. Variations in the flue gas temperature make the design and operation of an SNCR system more difficult.

Residence Time

Residence time is the amount of time the reactants are present within the chemical reactor, the upper area of the furnace and convective passes. Before the reactants exit the boiler, all the steps in the SNCR process must be complete, including:

- Mixing of the injected urea with the flue gas;
- Evaporation of water;
- Decomposition of urea to NH_3;
- Decomposition of NH_3 to NH_2 and free radicals; and
- NO_x reduction reaction chemistry.

Increasing the residence time available for mass transfer and chemical reactions generally increases the NO_x removal. In addition, as the temperature window for the reaction is lowered, greater residence time is required to achieve the same NO_x reduction level. Residence time can vary from 0.001 to 10 seconds [1]. However, the gain in performance for residence times greater than 0.5 seconds is generally minimal [10]. Figure 1.4 shows the effect of residence time and temperature on NO_x reduction.

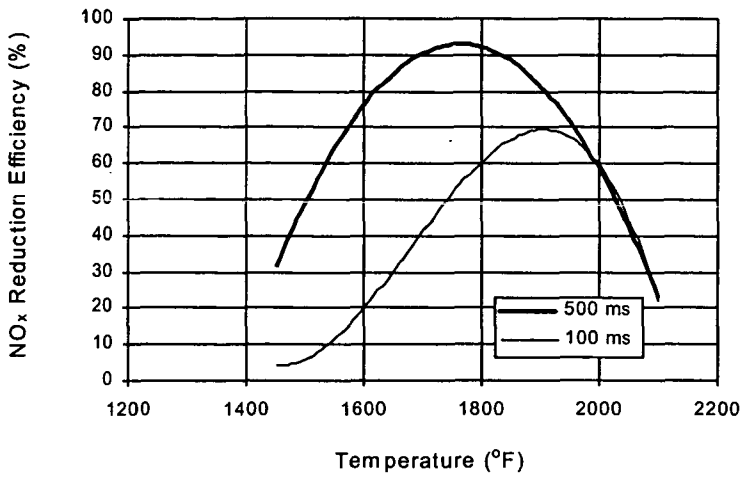


Figure 1.4: Effect of Residence Time on NO_x Reduction

The amount of residence time depends on the dimensions of the boiler gas path and the volumetric flow rate of the flue gas along the boiler gas path. These design parameters are optimized for boiler operations, not the SNCR process. The boiler residence time is set to meet steam generation requirements while maintaining a flue gas velocity to prevent erosion of the watertubes in the boiler. Because of these boiler design requirements, the residence time in the boiler is not always ideal for the SNCR process.

Degree of Mixing

For the reduction reaction to take place, the reagent must be dispersed and mixed throughout the flue gas. The dispersion must occur rapidly due to the volatility of ammonia. The mixing requirements are generally boiler specific and dependant on the air flow profiles through the boiler [1]. Areas of stagnation or high flow must be accommodated. Dispersion and mixing of the reagent and flue gas is more difficult on large boilers due to their size and flow patterns.

Mixing is performed by the injection system. The injectors atomize the reagent and control the spray angle, velocity, and direction of the injected reagent. These systems are boiler and reagent specific. Numeric modeling of the flue gas and reagent flow optimizes the design of the injection system. (See Section 1.2.5 Other Considerations).

To assist in dispersion of aqueous urea, the reagent is atomized into droplets by specially designed nozzles which optimize the droplet size and distribution. Evaporation time and trajectory are a function of the diameter of the droplet. Larger droplets have more momentum and penetrate farther into the flue gas stream. However, they require a longer time to volatilize, increasing the required residence time. [1]

Inadequate mixing results in insufficient NO_x reduction. Mixing patterns can be improved by several methods:

- Increase the energy imparted to the droplets;
- Increase the number of injectors;
- Increase the number of injection zones; and
- Modify the atomizer nozzle design to improve the solution droplet size, distribution, spray angle, and direction.

Uncontrolled NO_x

The concentration of the reactants also affects the reaction rate of the NO_x reduction process. The reaction kinetics decrease as the concentration of reactants decreases. This is due to thermodynamic considerations which limit the reduction process at low NO_x concentrations [1]. For lower NO_x inlet concentrations, the optimum temperature for the reaction is lower, hence, the percent NO_x reduction is lower. Figure 1.5 shows the NO_x reduction efficiency as a function of temperature for several uncontrolled NO_x levels.

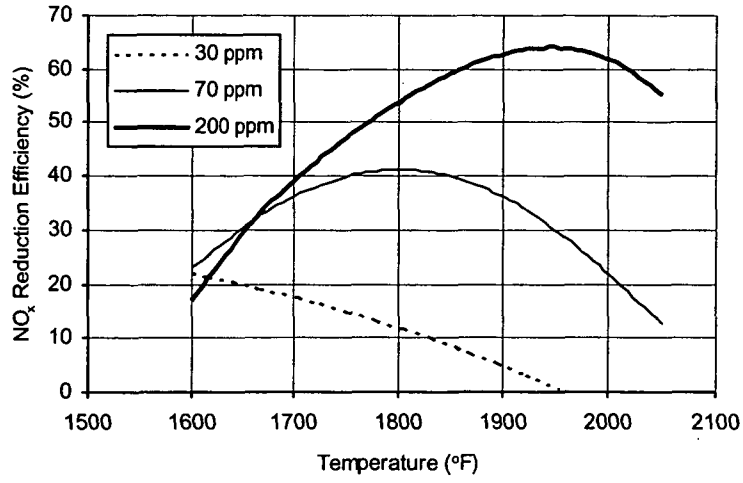


Figure 1.5: Effect of Uncontrolled NO_X Level on NO_X Reduction Efficiency

Normalized Stoichiometric Ratio

The Normalized Stoichiometric Ratio (NSR) defines the amount of reagent needed to achieve the targeted NO_x reduction. Based on reaction equations (1.1) and (1.2), theoretically two moles of NO_x can be removed with one mole of urea or two moles of ammonia. In practice, more than the theoretical amount of reagent needs to be injected into the boiler flue gas to obtain a specific level of NO_x reduction. This is due to the complexity of the actual chemical reactions involving NO_x and injected reagent and mixing limitations between reagent and flue gas (rate kinetics). Typical NSR values are between 0.5 to 3 moles of ammonia per mole of NO_x [10]. Because capital and operating costs depend on the quantity of reagent consumed, determining the appropriate NSR is critical. The factors that influence the value of NSR include:

- Percent NO_x reduction;
- Uncontrolled NO_x concentration in the flue gases;
- Temperature and residence time available for the NO_x reduction reactions;
- Extent of mixing achievable in the boiler;
- Allowable ammonia slip; and
- Rates of competing chemical reactions.

Section 4-1.1.3 Design Parameters provides further discussion of these influences and a method for estimating the NSR.

Figure 1.6 shows the NO_x reduction as a function of the NSR. Note that as the NSR increases, the NO_x reduction increases. However, as the NSR increases, the increment of NO_x reduction decreases exponentially. Rate kinetics limit the possible NO_x reduction to much less than the theoretical value. Increasing the quantity of reagent does not significantly increase the NO_x reduction for NSR values over 2.0.

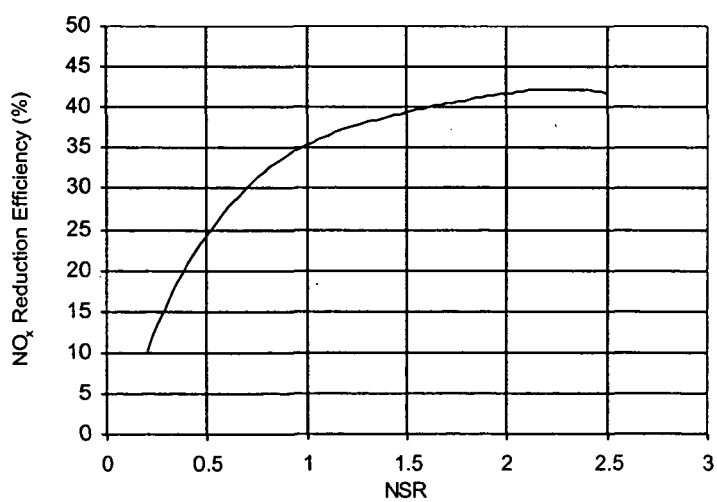


Figure 1.6: Effect of NSR on NO_x Reduction

Ammonia Slip

Typical NSR values require significantly more reagent to be injected in practice than required by the theoretical stoichiometric ratio. In addition, the amount of NO_x removed is generally much less than the amount of uncontrolled NO_x. This leaves a large portion of the injected reagent unreacted. Most of the excess reagent used in the process is destroyed through other chemical reactions. However, a small portion remains in the flue gas as ammonia slip. Figure 1.7 shows an example of the NO_x reduction efficiency which can be achieved for an uncontrolled NO_x level of 120 ppm and various ammonia slip levels.

Ammonia in the flue gas stream has several negative impacts. As shown in Table 1.1, ammonia has a detectable odor at levels of 5 ppm or greater and poses a health concern at levels of 25 ppm or greater. It can cause a stack plume visibility problem by the formation of ammonium chlorides which occur when burning fuels containing chlorine compounds. Furthermore, ammonium bisulfate and ammonium sulfate form when burning sulfur-cotaining fuels. Ammonia-sulfur salts can plug, foul, and corrode downstream equipment such as air heater, ducts, and fans. Lastly, the ability to sell the fly ash as a secondary product is affected by its ammonia concentration. Ammonia slip impacts are discussed further in Section 4-1.1.2.4 Other Considerations.

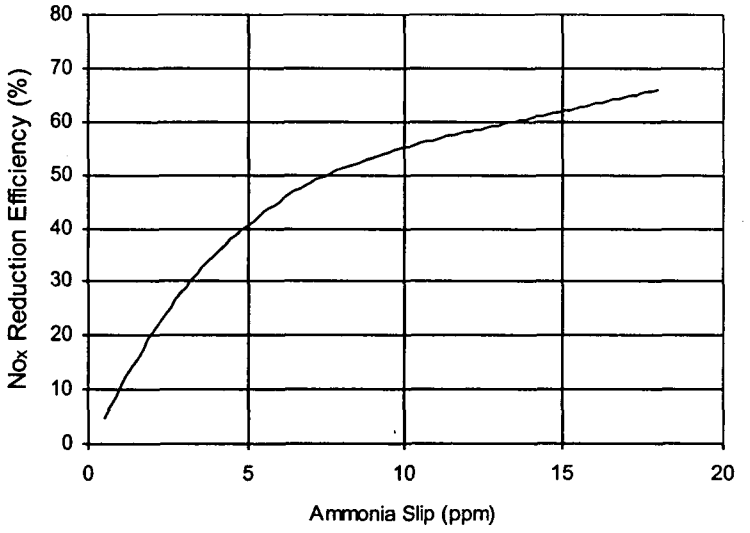


Figure 1.7: NO_x Reduction for Various Ammonia Slip Levels

Limits on acceptable ammonia slip, imposed by either regulatory limits or by design requirements, place constraints on SNCR performance. Injection of urea at higher NSR values can improve NO_x reduction, but may also increase ammonia slip. In addition, variation in the temperature profile of the boiler during operations can increase ammonia slip. In general, current SNCR

systems control ammonia slip between 5 to 10 parts per million (ppm) [12]. Instrumentation for monitoring ammonia slip is currently being developed but not commercially available [11]. One method to quantify ammonia slip is to determine the ammonia concentration in collected fly ash.

1.2.4 SNCR System

There are two basic designs for the application of SNCR. The first is an ammonia based system known as Thermal DeNO$_x$ ® that was developed and patented by Exxon Research and Engineering Company in 1975. The second system was developed and patented by The Electric Power Research Institute (EPRI) in 1980. It is a urea-based process known by the trade name NO$_x$ OUT®. The technology was licensed to Fuel Tech, which holds several additional patents that claim improvements and enhancements to the basic process. Fuel Tech has several sub-licensees authorized to supply and install SNCR technology in several industrial sectors.

An SNCR system has four basic steps to accomplish:

- Receiving and storage of the reagent;
- Metering, dilution, and mixing of the reagent;
- Injection of diluted reagent at appropriate locations in the boiler; and
- Mixing of the reagent with flue gas.

These steps are common to both urea and amonia SNCR. However, the design and equipment specifications for the two systems are different, since ammonia is injected as a vapor while urea is injected as an aqueous solution. Urea is typically used in large boiler applications of SNCR because it is safer to store and has better dispersion properties. A discussion of the SNCR equipment is given below. Figure 1.8 presents a simplified system flow schematic and Table 1.2 lists the equipment requirements for **urea-based SNCR.**

Urea-based systems typically employ a modular design to allow for boiler-specific design requirements while minimizing capital costs. Modular shops assembly of pumps, valves, internal piping, instruments, and controls reduces field insallation time and relatied costs while providing flexibility for future expansion [1]. The components are assembled into functional units and mounted on stainless steel skid modules. These modules can then be transported o the site and installed directly. The skid modules shown in Figure 1.8 will be discusses futher in the next sections.

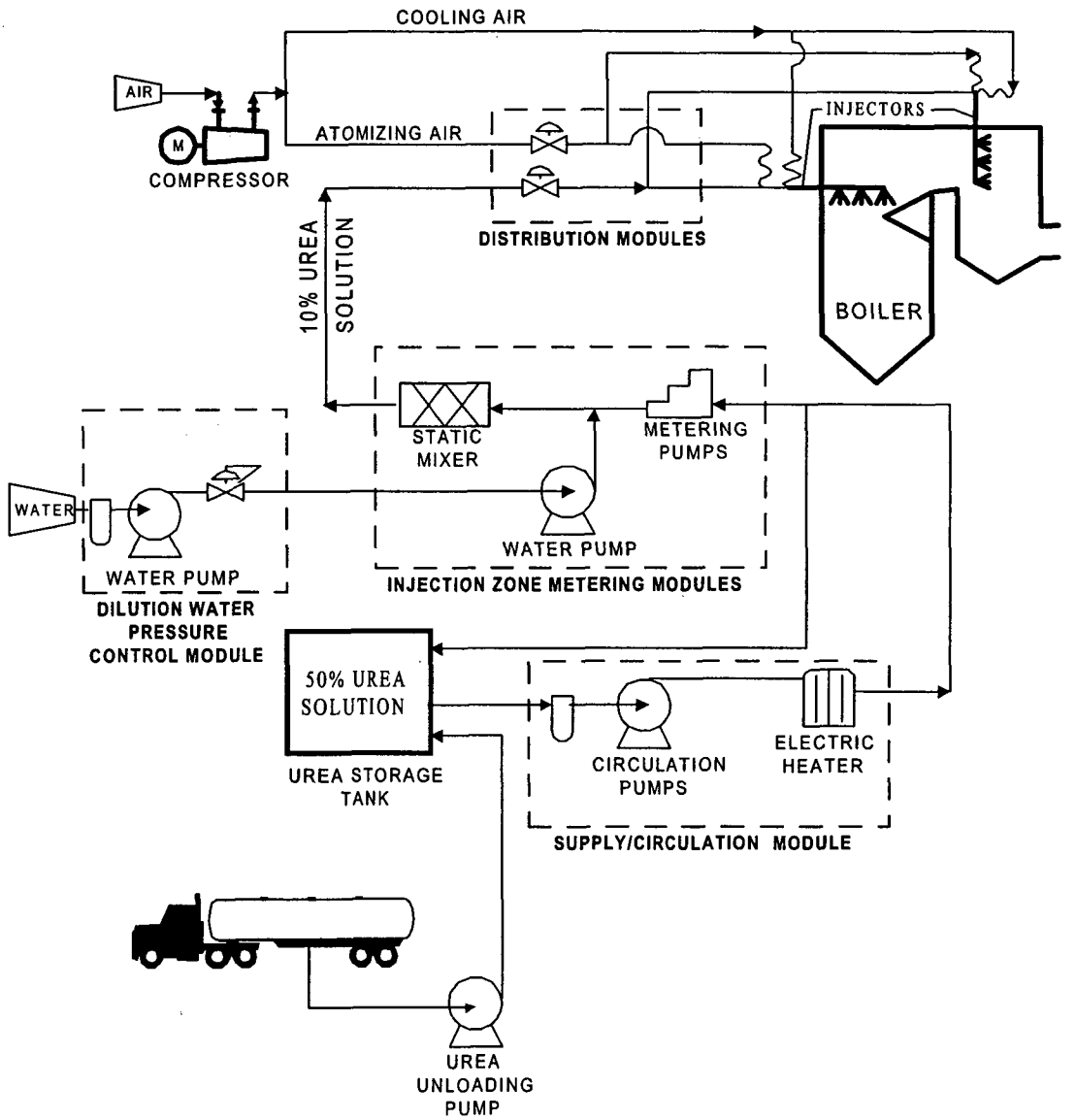


Figure 1.8: Urea SNCR Process Flow Diagram

Table 1.2: Urea-Based SNCR System Equipment

Item	Description/Size
Urea unloading skid	Centrifugal pumps with hoses to connect to rail tank car or truck
Urea storage tanks	Vertical, insulated fiberglass reinforced plastic (1 or more tanks) (vinyl ester resin) tank, atmospheric pressure design, and equipped with a vent, caged ladder, manway, and heating pads
Circulation module	Skid-mounted circulation module consisting of: - Circulation pumps, - Electric heaters, - Insulated/heat traced piping - Isolation valves for pumps and heaters, - Instrumentation for flow, pressure, temperature, and a control panel
Injection zone metering (IAM) modules (1 to 5 modules)	Skid mounted metering modules consisting of: - Metering pumps, hydraulic diaphragm type equipped with a variable speed motor drive, - Water booster pumps, turbine type, - Insulated/heat traced piping - Isolation and control valves for pumps, - Instrumentation for flow, pressure, temperature, and a control panel
Air compressor Distribution modules (1 to 5 modules)	Rotary type Urea solution distribution module consisting of: - Valved connections for urea and atomizing air, - Isolation valve and a pressure control valve for the air/urea supply to each injector, - Pressure indicator for air/urea supply to each injector, - Flow indicator for urea supply to each injector
IInjectors (4 to 12 per distribution	Wall-type: Dual-fluid type wall injector, with modules) furnace wall panels, and hoses for air and urea supplies Lance-type: Dual-fluid type lance injector, with furnace wallpanels, and hoses for air and urea supplies
Piping	Between urea unloading skid and urea tank; urea tank and circulation module; and circulation module and IZM modules(s). Insulate/heat traced piping, stainless steel
Piping	Between IZM module(s) and distribution modules. Insulated/heat traced tubing, stainless steel
Tubing	Between distribution modules and injectors. Insulated/heat traced tubing, stainless steel
Dilution water piping	Insulated/heat traced piping, carbon steel, with isolation and pressure reducing valves
Miscellaneous piping	Piping/tubing and valves for flushing water, atomizing air, and control air
Piping supports	Structural support steel, including a pipe bridge, for supporting all piping
Economizer outlet emission monitors	Monitor NO and O_2 in the flue gas and provide a feedback signal for urea injection control
Instrumentation and controls	Instrumentation and stand-alone, microprocessor-based controls for the SNCR system with feedback from the plant controls for the unit load, NO emissions, etc.
Enclosures	Pre-engineered, heated and ventilated enclosure for the circulation and metering skids
Foundations	Foundations and containment walls for the tank and equipment skids, enclosure, and piping support steel, as required
Platforms/stairways	Platform/stairway modifications and additions for access to injectors
Asbestos removal	Asbestos removal and reinsulation, for a retrofit installation

from 10,000 to 20,000 gallons per tank to maintain sufficient volume for 1 to 3 weeks of SNCR operations. A closed top, flat bottom, vertical tank is used for urea storage. These tanks are usually constructed of fiber- reinforced polyester and have a corrosion barrier coating on the inside made of premium-grade vinyl ester resin. The tanks are equipped with level and temperature indicators, a manway, vent, and access ladder, and other appurtenances. The applicability of heat tracing, insulation, and seismic design criteria are determined based on site-specific conditions. The tank should be mounted on a concrete pad and surrounded by a spill containment structure such as a dike.

Circulation Module

The circulation module maintains continuous circulation of the stored urea and supplies high-flow, high-pressure urea to the injection system. The circulation module pumps the urea from the storage tank to the components on the module. The urea solution is filtered to avoid clogging of the injectors and heated to prevent the solution from freezing. The urea is then returned to the tank or sent to the injection system. The module also provides a local/remote control and monitoring station for the storage tank and circulation system. This module contains multistage stainless steel centrifugal pumps, in-line duplex strainers, electric heaters, and instrumentation and controls for reagent pressure, flow, temperature, and quantity. [1]

Dilution, metering, and mixing of the reagent

Dilution Water Pressure Control Module

The dilution water pressure (DWP) control module provides filtered plant water at the proper pressure for reagent dilution. The plant water is filtered to less than 50 milligrams per liter (mg/l) of suspended solids and low dissolved solids. The DWP typically consists of two full-flow multistage stainless steel centrifugal pumps, an in-line duplex strainer, pressure control valves, and the required pressure/flow instrumentation. Through the use of backpressure controllers and multistage pumps, this system maintains a constant supply of dilution water, at the design pressure, in response to the changing SNCR process demand signals. [1]

Injection Zone Metering Module

The injection zone metering (IZM) module meters and controls the reagent concentration and flow to each zone of injection in the boiler. The aqueous urea generally requires dilution before injection to achieve the correct NSR between the reagent and flue gas NO_x. The reagent is diluted using filtered plant water from the DWP module. Each IZM module includes a chemical metering pump, a water pump, an in-line static mixer, a local control panel, zone isolation valves, and magnetic flow

meters and control valves for chemicals and water. The module design generally incorporates independent chemical flow and zone pressure valves which respond to signals from the control systems, the master control module and local programmable logic controller (PLC). Through the control system, the module adjusts solution flow rates and activates or deactivates injection zones in response to changes in outlet NO_x concentration, boiler load, or fuel quality. Urea-based SNCR systems typically employ one to five IZM modules, depending on the boiler size, configuration, the uncontrolled NO_x concentration, and desired NO_x removal efficiency. Several IZM modules can be combined onto one skid-mounted system. [1]

Injection of diluted reagent at appropriate locations in the boiler

Reagent Distribution Module

The mixed and diluted urea solution is transported from the IZM to the distribution modules, which are typically located adjacent the boiler. The distribution modules control the flow of the solution to each injector. Each of the distribution modules consists of flow meters, balancing valves, and regulators connected to an automatic control system. The control system accurately controls and displays the reagent and atomizing air or steam flow to each injector. The modules also include Manual ball valves, gauges, and stainless steel tubing to adequately control the urea injection process. There is one distribution module for each IZM module providing reagent to multiple injectors. [1]

Injection Locations

The urea solution flows from a given distribution module to a set of injectors. For large boiler applications, multiple injectors are located within several different zones of the boiler and can be operated independently or in groups (sub-zones) via the IZM. Controlling the amount and location of reagent injection gives the system flexibility to respond to variation in the boiler operating conditions and to maintain ammonia slip levels.

The number and location of the zones is determined by the temperature and flow patterns of the boiler. The locations are optimized using numeric modeling of flow and chemical reactions (see Section 1.2.5 Other Considerations). Typical designs employ between one and five injection zones with 4 to 12 injectors per zone [1]. Injectors are located in open areas of the boiler such as the region between the superheater and reheater sections. Figure 1.2 illustrates this configuration. For SNCR retrofit of existing boilers, optimal locations for injectors may be occupied by boiler equipment such as the watertubes. Removal or relocation of this equipment increases the installation costs. Installation in sub-optimal boiler areas decreases the NO_x reduction efficiency which can be achieved by the system while maintaining the required ammonia slip level.

Mixing of the reagent with flue gas and reduction of NO_x

Injectors

The injectors assist in dispersion and mixing of the reagent with the flue gas. There are two types of injectors, wall and lance.

> Wall injectors are attached to the inner wall of the boiler at specified locations. There is generally one nozzle for each injector location. Wall injectors are used in smaller boilers and urea based systems, where short-range injection is sufficient to mix the reagent with the flue gas. They have a longer operating life than lance injectors since they are not directly exposed to hot flue gas.
>
> Lance injectors consist of a small pipe which protrudes from the boiler wall into the flue gas pathway. Nozzles are located along the pipe directly in the flue gas pathway. Lance injectors are used for ammonia gas systems and in large boilers where mixing of the flue gas and reagent is more difficult. In some designs, the lance extends across the entire width of the boiler pass. Lance injectors can be single or multi-nozzle designs. Multi-nozzle lances are more complicated design, therefore, they are more expensive than single-nozzle lance or wall injectors [1].

SNCR systems may employ one or both types of injectors.

Injectors are subject to high-temperatures and to flue gas impingement which cause erosion, corrosion, and structural integrity degradation. Therefore, injectors are generally constructed of stainless steel and designed to be replaceable. In addition, injectors are often cooled with air, steam, or water. Lance injectors and some wall injectors are also designed to be retractable when not in use. This minimizes their exposure to the hot flue gas when the SNCR system is not being operated because of seasonal operations, boiler startup or shutdown, or other operational reasons.

The reagent is injected under pressure and atomized by specially designed nozzle tips to create droplets of the optimum size and distribution. The spray angle and velocity of the injection control the trajectory of the reagent. Urea systems often inject a carrier fluid, typically air or steam, along with the urea through a dual-fluid atomizer nozzle. The reagent can be injected with a low energy or high energy system. A low-energy system uses little or no pressurized air while a high-energy system uses large amounts of compressed air or steam to inject and vigorously mix the solution with the flue gas. Lance injectors in large boilers typically use high-energy systems. High energy systems are more expensive to build and operate since they require a larger compressor, a more robust injection system, and consume more electric power.

The reagent injection systems used for ammonia-based systems are generally more complicated and expensive than those used in urea-based systems [1]. These systems inject gas phase ammonia rather than an aqueous solution. For this reason, ammonia-based systems often use high-energy lance systems with multiple injectors. The lances are placed in a grid formation across the width and height of the boiler passes.

1.2.5 Other Considerations

Retrofit

The difficulty of SNCR retrofit on exiting large coal-fired boilers is considered to be minimal. The primary concern is adequate wall space within the boiler for installation of injectors. The injectors are installed in the upper regions of the boiler, the boiler radiant cavity and convective cavity. Movement and/or removal of existing watertubes and asbestos from the boiler housing may be required. In addition, adequate space adjacent to the boiler must be available for the distribution system equipment and for performing maintenance. This may require modification or relocation of other boiler equipment such as ductwork. Retrofit installation of the SNCR system generally calls for additional expenditures in the range of 10% to 30% of the SNCR system cost. [10]

Ammonium Sulfate Deposition

Sulfur trioxide (SO_3) forms during the combustion of fuels which contain sulfur. It reacts with ammonia in the flue gas downstream of the boiler (ammonia slip) to form ammonium bisulfate and ammonium sulfate. The amount formed depends on the sulfur content of the fuel and the amount of ammonia slip. Ammonia-sulfur salts can plug, foul, and corrode downstream equipment such as the air heater, ducts, and fans. Depending on the rate of ammonia-sulfur salt deposition on downstream equipment more frequent acid washing of this equipment may be warranted. Increased acid washing generates additional wastewater which must be disposed or treated by the plant. Ammonia slip limits are generally imposed as part of the design requirements to avoid impacts on downstream equipment.

Ammonia sulfates also deposit on the fly ash that is collected by particulate removal equipment. The ammonia sulfates are stable until introduced into an aqueous environment with a elevated pH levels. Under these conditions, ammonia gas can release into the atmosphere. This results in an odor problem or, in extreme instances, a health and safety concern. Plants that burn alkali coal or mix the fly ash with alkali materials can have fly ash with high pH. In general, fly ash is either disposed of as waste or sold as a byproduct for use in processes such as concrete admixture. Ammonia content in the fly ash greater than 5 ppm can result in off-gassing which would impact the salabilty of the ash as a byproduct and the storage and disposal of the ash by landfill. [10]

Computational Fluid Dynamics and Chemical Kinetic Modeling

Each boiler unit has a unique temperature and flow gradient with areas of high flow and stagnation. In addition, temperature and flow profiles vary according to the load capacity under which the boiler is operating. A mathematical model is developed to describe this stratification and variation of important species such as NO_x and SO_3 in the flow stream. To develop the model, the flue gas temperature and velocity within the boiler are measured at many locations. These measurements are used in a computational fluid dynamics (CFD) model for the convective passes of the boiler. The model predicts the temperature and gas flow within the boiler for various operating conditions and injection scenarios.

The residence times and temperatures predicted by the CFD model are input into a chemical kinetic model (CKM). CKM defines the chemical reactions associated with the SNCR process in the boiler. Analysis of the fuel and flue gas constituents are required to develop this model. The model predicts the reactions and rates of reactions within the boiler in order to estimates the NO_x reduction along the flue gas pathway.

Modeling optimizes the SNCR design for the boiler of concern to obtain the maximum NO_x reduction within acceptable ammonia slip limits. It determines design parameters such as the NSR, injector locations, and optimum droplet size and distribution. In general, SNCR vendors obtain the required measurements and develop the models. The cost of model development is generally included in the purchased equipment cost for SNCR. [1]

Additives/Enhancers

Additives to the reagents are called enhancers and can be used to lower the temperature range at which the NO_x reduction reaction occurs. During low-load operation, the location of the optimum temperature region shifts upstream within the boiler. This shift requires the injection point of the reagent to be moved upstream. The use of an enhancer reduces the need for additional injection locations which are required to compensate for variable load operation. Fewer injection locations decrease capital costs and the need for modifications to the boiler. In addition, the larger temperature range available with enhancers increases the available residence time for the reduction reaction, further reducing NO_x emissions.

Additional reagent is injected with the enhancer to maintain the same NO_x reduction efficiency since some of the reagent reacts with the enhancer as opposed to the NO_x. This can increase the reagent usage by up to 10%. In addition, enhancers can result in increased levels of CO and N_2O in the stack effluent. Enhancers require additional storage, distribution, and control system equipment. The increase in annual cost from the usage of additional reagent and enhancer can be significant. Enhancer formulations are generally proprietary. [10]

Energy Consumption

An SNCR process reduces the thermal efficiency of a boiler. The reduction reaction uses thermal energy from the boiler, which decreases the energy available for power or heat generation. As a result, additional energy is required for the boiler to maintain the same steam output. Pretreatment and injection equipment, pumps, compressors, and control systems, also require electricity. This increased usage of fuel and electricity increases the annual costs required to operate the boiler [1]. Section 1.4.2 Total Annual Costs presents a method for estimating the additional fuel and electricity usage.

1.3 Design Parameters

SNCR system design is a proprietary technology. Extensive details of the theory and correlations that can be used to estimate design parameters such as the required NSR are not published in the technical literature [5]. Furthermore, the design is highly site-specific. In light of these complexities, SNCR system design is generally undertaken by providing all of the plant- and boiler-specific data to the SNCR system supplier, who specifies the required NSR and other design parameters based on prior experience and computational fluid dynamics and chemical kinetic modeling. [1]

The procedure given below in Section 1.3.1 Design Parameters for Study-Level Estimates is a step-by-step approach to estimate design parameters based on a procedure developed in the draft EPA report Selective Noncatalytic Reduction for NO_x Control on Coal-fired Boilers [1]. This procedure assumes SNCR system size and cost are based on three main parameters: the boiler size or heat input, the required level of NO_x reduction, and the reagent consumption. Data requirements for obtaining vendor cost estimates based on design specifications or performance specifications are outlined in Section 1.3.2.

1.3.1 Design Parameters for Study Level Estimates

Boiler Heat Input

The primary cost estimation parameter in the methodology presented in Reference [1] is the maximum potential heat released by the boiler or heat input rate, Q_B, expressed as million British thermal units per hour (MMBtu/hr). It is obtained from the high heating value, *HHV*, of the

fuel in Btu per pound (Btu/lb) multiplied by the maximum fuel consumption rate in pounds per hour (lb/hr),

$$Q_B = HHV \ \dot{m}_{fuel} \tag{1.3}$$

where Table 1.3 gives the HHV for various coals and a conversion of 10^{-6} MMBtu/Btu is reequired.

Table 1.3: High Heating Values for Various Coals

Type of Coal	Energy Content (Btu/ lb)
Lignite	5000-7500
Subbituminous	8000-10,000
Bituminous	11,000 - 15,000
Anthracite	14,000

If the boiler produces electricity, then its heat input can be estimated using the boiler net plant heat rate, *NPHR* in Btu per kilowatt-hour (Btu/kWh):

$$Q_B = B_{MW} \ NPHR \tag{1.4}$$

where B_{MW} is the boiler megawatt (MW) rating at full load capacity. Note that if *NPHR* is not known (e.g., a cogeneration unit), a value of 9,500 Btu/kWh can be used as a reasonable estimate conversions of 1,000 kW/MW and 10^{-6} MMBtu/Btu are required. Using this value, the heat input rate, Q_B, is:

$$Q_B\left(\frac{MMBtu}{hr}\right) = 9.5\left(\frac{MMBtu}{MWh}\right) B_{MW}(MW) \tag{1.5}$$

System Capacity Factor

The total system capacity factor, CF_{total}, is a measure of the average annual use of the boiler in conjunction with the SNCR system. CF_{total} is given by:

$$CF_{total} = CF_{plant} \times CF_{SNCR} \tag{1.6}$$

For industrial boilers, the capacity factor of the boiler, CF_{plant}, is the ratio of actual quantity of fuel burned annually to the potential maximum quantity of fuel burned annually in pounds (lbs). CF_{plant} is given by:

$$CF_{plant} = \frac{actual\ m_{fuel}}{maximum\ m_{fuel}} \quad (1.7)$$

SNCR can be operated year-round or only during the specified ozone season. The capacity facor for the SNCR system, CF_{SNCR}, is the ratio of the actual number of SNCR operating days, t_{SNCR}, to the total number of days per year, 365 days:

$$CF_{SNCR} = \frac{t_{SNCR}}{365\ days} \quad (1.8)$$

Uncontrolled NO_x and Stack NO_x

Uncontrolled NO_x, represented as NO_{xin}, is the NO_x concentration in the flue gas after any reductions provided by combustion controls, but prior to the SNCR system. The uncontrolled NO_x concentration, obtained from analyzing the boiler flue gas stream, is generally given in pounds per MMBtu (lb/MMBtu) of NO_2 [2].

The stack NO_x, represented as NO_{xout} is the required NO_x emission limit at the stack outlet. It is generally set by the plant or regulatory limits and also given in pounds per MMBtu (lb/ MMBtu) of NO_2.

NO_x Removal Efficiency

The NO_x removal efficiency, represented as η_{NOx}, is determined from the uncontrolled NO_x level of the boiler at maximum heat input rate, $CF_{plant} = 1.0$, and the required stack emission limit using the following equation:

$$\eta_{NO_x} = \frac{NO_{x_{in}} - NO_{x_{out}}}{NO_{x_{in}}} \quad (1.9)$$

Normalized Stoichiometric Ratio

The Normalized Stoichiometric Ratio, *NSR*, indicates the actual amount of reagent needed to achieve the targeted NO_x reduction. The actual quantity of reagent is greater than the theoretical

quantity due to reaction kinetics (See Section 1.2.3 Performance Parameters). The *NSR* is defined as:

$$NSR = \frac{moles\ of\ equivalent\ NH_3\ injected}{moles\ of\ uncontrolled\ NO_x} \tag{1.10}$$

For estimating purposes, the moles of NO_x are equivalent to the moles of NO_2. Note that the moles of equivalent NH_3 in Equation (1.10) are the moles of NH_2 that will be released from the reagent.

The actual stoichiometric ratio, *ASR*, is defined as:

$$ASR = \frac{moles\ of\ reagent\ injected}{moles\ of\ uncontrolled\ NO_x} \tag{1.11}$$

ASR can also be calculated from the *NSR* using the following equation:

$$ASR = \frac{NSR}{SR_T} \tag{1.12}$$

where SR_T is the ratio of equivalent moles of NH_3 per mole of reagent injected. From the chemical formula for ammonia (NH_3) and urea ($CO(NH_2)_2$) given in the reaction equations (1.1) and (1.2), SR_T is 1 for ammonia and 2 for urea.

Reagent utilization is the ratio of moles of reagent reacted to the moles injected. This indicates how much reagent is being reacted versus how much reagent is passing through as ammonia slip. Utilization of reagent can be calculated from the *NSR* and the NO_x reduction efficiency as follows:

$$Utilization = \frac{\eta_{NO_x}}{NSR} \tag{1.13}$$

The derivation for this equation is presented in Reference [1].

Methods for estimating *NSR* are considered proprietary. A simplified *NSR* estimation procedure was developed by The Cadmus Group, Bechtel Power, Inc. and SAIC in the EPA draft report, Selective Noncatalytic Reduction for NO_x Control on Coal-fired Boilers [1]. This procedure was developed using linear regression and NSR data from References [2] and [3]. The values of *NSR* derived using this approach should not be used for equipment design or guarantee purposes.

The NSR estimation equation is valid from 0 to 50 percent NO_x reduction [1]. The equation used to estimate the NSR for urea reagent is where Noiu is given in lb/MMBtu.

$$NSR = \frac{\left[2NO_{x_{in}} + 0.7\right]\eta_{NO_x}}{NO_{x_{in}}} \tag{1.14}$$

Figure 1.8 provides a graphical representation of this *NSR* estimation method. Generally, the value of the NSR ranges between 0.5 and 2.0 in industrial and utility boilers with utilization ranging between 25 and 50 percent.

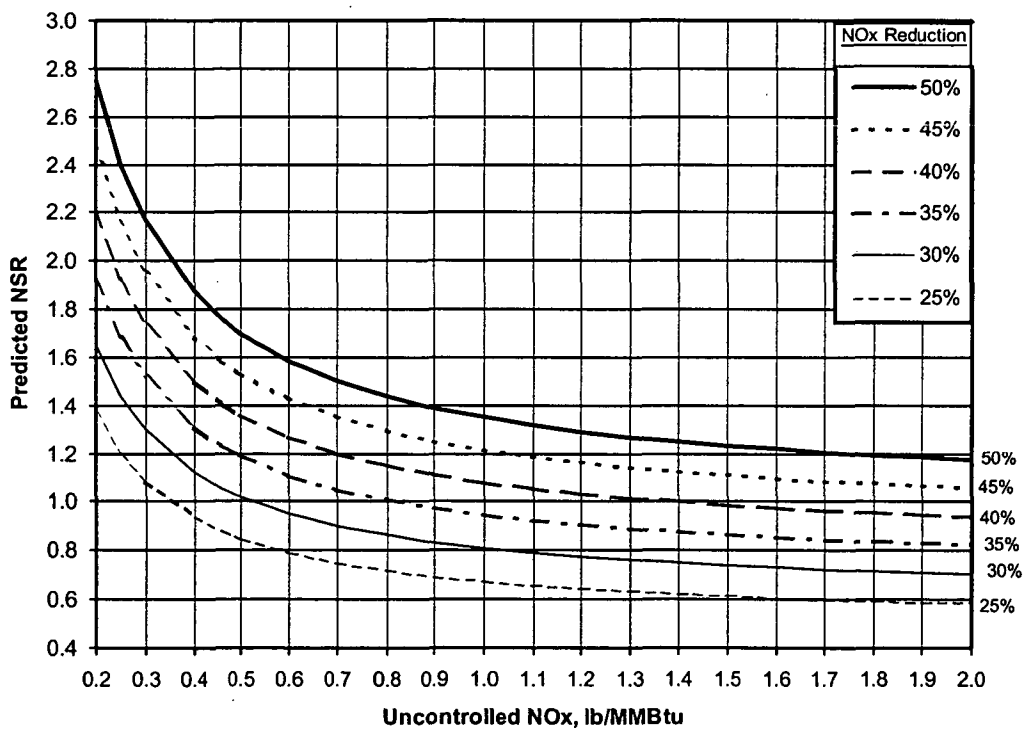


Figure 1.8: Approximate NSR Estimation for Urea

In a design developed by a system supplier, the NSR would be adjusted to account for several parameters that are not included in the NSR estimation equation. The following parameters are used by the system supplier to more accurately predict the NSR for a given boiler:

Reaction temperature range available within the boiler superheater (radiative and convective section) and primary reheater (convective section or cavity) region. If the required temperature window occurs in the radiant section of the boiler, the NSR could decrease. However, if the temperature window occurs in the convective section, the NSR may increase.

Residence time available in the desirable temperature range. The required NSR decreases as the available residence time increases.

Degree of mixing between the injected chemical and the flue gases. The NSR decreases as the degree of mixing increases.

Ammonia slip vs required NO_x reduction. Tighter constraints on ammonia slip would dictate lower NSRs, thereby limiting the achievable NO_x reduction.

Estimating Reagent Consumption and Tank Size

Once the *NSR* is estimated, the rate of reagent consumption or mass flow rate of the reagent, $\dot{m}_{reagent}$ expressed as pounds per hour (lb/hr), can be calculated using:

$$\dot{m}_{reagent} = \frac{NO_{x_{in}} \; Q_B \; \eta_{NO_x} \; NSR \; M_{reagent}}{M_{NO_x} \; SR_T} \qquad (1.15)$$

where the $M_{reagent}$ is the molecular weight of the reagent (60.06 grams per mole for urea, 17.07 grams per mole for ammonia) and M_{NOx} is the molecular weight of NO_2 (46.01 grams per mole). The molecular weight of NO_2 is used because the NO_x emissions, NO_{xin}, are given in lb/MMBtu of NO_2. As stated previously, SR_T is the ratio of equivalent moles of NH_3 per mole of reagent (1 for ammonia and 2 for urea).

For urea or ammonia, the mass flow rate of the aqueous reagent solution is given by:

$$\dot{m}_{sol} = \frac{\dot{m}_{reagent}}{C_{sol}} \qquad (1.16)$$

where C_{sol} is the concentration of the aqueous reagent solution by weight.

The solution volume flow rate q_{sol}, generally expressed as gallons per hour (gph), is given by:

$$q_{sol} = \frac{\dot{m}_{sol}}{\rho_{sol}} \tag{1.17}$$

where ρ_{sol} is the density of the aqueous reagent solution inlb/ft^3 A conversion of 7.4805 gal/ft^3 is required to correct the units. given in Table 1.1, 71.0 lb/ft^3 for 50% urea and 56 lb/ft^3 for 29% ammonia.

The total volume stored in the tank, or tanks, is based on the volume that the SNCR system requires for operating a specified number of days. The volume stored on site for the number of operating days, $t_{storage}$, is :

$$V_{tank} = q_{sol}\, t_{storage} \tag{1.18}$$

where V_{tank} is in gallons, q_{sol} is in gph and a conversion of 24 hr/day is required.

Note that the tank volume is typically based on the maximum annual heat input rate, so the capacity factor is not included in Equation 1.18. A common on site storage requirement is for 14 days of SNCR operation.

1.3.2 Design Parameters for Detailed/Performance Specifications

Cost Estimates Based on Detailed Specifications

This subsection is included to describe the information that must be assembled and furnished to a supplier to prepare design specifications, particularly that component information with the greatest influence on system cost. SNCR capital and operating costs can be estimated if the major cost items are identified and the system is defined in adequate detail, [1]. The following data are provided to the system supplier for SNCR system design:

- The boiler capacity in terms of heat input rate (MMBtu/hr);
- Boiler capacity profile – percent of time the boiler operates at a given heat input rate;
- Type of combustion unit – dry/wet bottom boiler, wall-fired, tangentially-fired, cyclone fired, other (e.g., stoker-fired); year built, and manufacturer;

- Boiler Dimensions – sectional side view; sectional front view; plan section through furnace (width × depth); furnace height (floor to furnace exit); firing zone height; sections through radiant and connective heat transfer cavities (zones); other unique features, e.g., division wall/panels, etc., in the furnace or backpass;

- Locations of boiler furnace overfire air ports;

- Locations and sizes of boiler observation ports, temperature probe ports, sootblower openings, and other locations for potential new ports;

- Air preheater design and operational data, including sootblower data;

- Fuel Data – proximate and ultimate analyses and high heating value (HHV) for primary and secondary fuels;

- Fuel firing rates at full and partial loads (e.g., 100 percent, 70 percent, and 30 percent);

- Test Data or Combustion Calculations - flue gas flow rate at design or actual conditions; excess air rate at full and partial loads; flue gas composition including O_2, NO_x, CO, SO_2, and HCl;.

- Flue gas temperature profile from furnace exit to economizer (i.e., where temperature drops to about 1,400 °F) at various loads;

- Flue gas residence time – available flue gas residence time in the upper furnace and convective pass within the temperature window for urea at various loads;

- Existing or planned uncontrolled NO_x and CO emission data in ppm or lb/MMBtu without the proposed SNCR system, including any change in emissions related to other installed or planned technologies (e.g., low-NO_x burner, gas recirculation). This should be specified for boiler operations at full load and selected partial loads;

- Minimum expected NO_x reduction or permit requirement for stack NO_x emission level (ppm or lb/MMBtu). This should be specified for boiler operations at full load and selected partial loads; and

- Allowable byproduct emission rates for regulated emissions such as ammonia and nitrous oxide.

The boiler supplier/manufacturer can furnish most of this information for existing or planned new units. For fuel data, the designer needs typical or design values, as well as the expected range. To define the temperature and flue gas velocity profiles in existing boilers, it is preferable to obtain actual measurements.

To obtain a representative cost quotation from an SNCR system supplier, the request should contain sufficient details to minimize design assumptions by the supplier. The request for quotation (RFQ) should include the technical specifications, as well as commercial terms and conditions.

Two important parts of the specification are work included in the scope of the supplier and work not included (i.e., work performed by the owner/operator). The more precise and detailed the specification of the work, the more accurate the overall system design and cost. For a turnkey scope (design, supply, and erect all equipment, and demonstrate commercial operation while meeting all performance criteria), the excluded work is minimal.

Cost Estimates Based on Performance Specifications

Preparation of detailed specifications involves significant time and effort (for both owner and supplier) and is not critical for budgetary cost estimates. To simplify the process, a performance specification approach may be used in the RFQ. In this approach, the basic required plant and fuel data are provided along with the required SNCR system performance requirements, excluding equipment-related details (e.g., materials of construction, equipment redundancy, and level of instrumentation and controls) [1].

The performance specification should include a description of the system and components in enough detail to understand the type and quality of system proposed by the supplier. A cost breakdown of major components and subsystems also should be obtained from the supplier to enable independent assessment, deletion, or addition, and to compare other bids on an equitable basis. The SNCR performance specification typically should request the following items regarding NO_x emission control performance, chemical consumption, and other consumption rates at full and partial loads:

- Guaranteed and expected NO_x emission rates in units of lb/MMBtu and lb/hr with averaging period as defined in the air quality permit of the facility;
- Guaranteed and expected NH_3 slip, ppm (other conditions such as dry basis, percent O_2, per the air permit);
- Guaranteed and expected N_2O emissions (guarantees needed if it is a permit condition);

- Other emission limits as specified (or anticipated) in the permit.
- Normalized stoichiometric ratio (NSR) proposed to achieve the required NO_x reduction;
- Guaranteed and expected reagent consumption rate;
- Guaranteed and expected dilution air, steam, or water consumption rate;
- Atomizing and cooling air (or steam) pressure and consumption rate; and
- Guaranteed and expected electrical power consumption.

1.4 Cost Analysis

The cost estimating methodology presented here provides a tool to estimate study-level SNCR capital and annual costs. Actual selection of the most cost-effective option should be based on a detailed engineering study and cost quotations from system suppliers. The costs are presented in 1998 dollars.

The cost estimating equations presented in this section are based on equations developed in the draft EPA report Selective Noncatalytic Reduction for NO_x Control on Coal-fired Boilers [1]. These equations follow the costing methodology used by the Electric Power Research Institute (EPRI) [7]. In the EPRI method, both the purchased equipment cost (PEC) and direct installation cost are estimated together. This methodology is different from the EPA Air Pollution Control Cost Manual methodology, which estimates equipment costs and installation costs separately. Due to the limited availability of equipment cost data and installation cost data, the equations for SNCR capital costs were not reformulated.

The capital and annual cost equations were developed for coal-fired wall and tangential utility and industrial boilers with heat input rates ranging from 250 MMBtu/hr to 6000 MMBtu/hr. The SNCR system design used for the cost estimate is an **urea-based system**. An ammonia-based system would have different storage, distribution and injection equipment costs. The cost equations are sufficient for NO_x reduction efficiencies up to 50%. Allowed ammonia slip for the SNCR system ranges from 2 to10 ppm. [1]

The cost equations are applicable to retrofit of SNCR on existing boilers. The cost estimating procedure, however, is suitable for retrofit or new boiler applications of SNCR on all types of coal-fired electric utilities and large industrial boilers. The increased cost due to retrofit is approximately 10% to 30% of the cost of SNCR applied to a new boiler [10].

1.4.1 Total Capital Investment

Total Capital Investment (TCI) includes direct and indirect costs associated with purchasing and installing SNCR equipment. Costs include the equipment cost (EC) for the SNCR system itself, the cost of auxiliary equipment, direct and indirect installation costs, additional costs due to installation such as asbestos removal, costs for buildings and site preparation, offsite facilities, land, and working capital. In general, SNCR does not require buildings, site preparation, offsite facilities, land, and working capital. A more detailed discussion of capital costs can be found in Section 1, Chapter 2 of this Manual.

Direct Capital Costs

The direct capital cost equation developed in Reference [1] is based on data for one representative boiler. This data is used to develop an equation for the boiler size range of interest by applying a scaling methodology from the EPRI Technical Assistance Guide (TAG) [3]. The equation assumes that performance parameters vary in direct proportion to the boiler size. The costs are adjusted to 1998 dollars using the Chemical Engineering Cost Index.

The direct capital cost estimate includes PEC such as SNCR system equipment, instrumentation, sales tax and freight. This includes costs associated with field measurements, numerical modeling and system design. It also includes direct installation costs such as auxiliary equipment (e.g., ductwork, compressor), foundations and supports, handling and erection, electrical, piping, insulation and painting. In addition costs such as asbestos removal are included. Table 1.2 presents a list of the equipment and installation requirements for a urea-based SNCR system.

The direct capital cost equation is a function of both boiler heat input rate *QB* in MMBtu/hr, and the NO_x removal efficiency -η_{NOx}. The boiler heat input rate indicates the size of the boiler and the flue gas flow rate, which directly influence the size of the SNCR system. The NO_x removal efficiency effects both the required urea flow rate and associated urea-related equipment. Increases in the NO_x removal efficiency will require higher urea flow rates and increase the size and/or number of the urea related equipment such as storage tanks. The equation for direct capital cost (DC) of urea-based SNCR in dollars is:

$$DC(\$) = \frac{\$950}{MMBtu}\ Q_B\left(\frac{MMBtu}{hr}\right)\left(\frac{2375\frac{MMBtu}{hr}}{Q_B\left(\frac{MMBtu}{hr}\right)}\right)^{0.577}\left(0.66+0.85\eta_{NO_x}\right) \qquad (1.19)$$

where 950 $/(MMBtu/hr) is the direct capital cost for a 2375 MMBtu/hr boiler at 40% NO_z removal efficiency [1].

Indirect Capital Costs

Indirect installation costs include costs such as; construction and contractor fees, startup and testing, inventory capital, and any process and project contingency costs. As explained in the Manual Section 1 Introduction, average values of indirect installation factors are applied to the direct capital cost estimate to obtain values for indirect installation costs. These costs are estimated as a percentage of the TCI. The equations for indirect installation costs are presented in Table 1.4.

Table 1.4: Capital Cost Factors for an SNCR Application

Description	**Data or Formula**
Boiler Size (MMBtu/hr)	Q_B
NO_x Removal Efficiency	n_{NOx}
Cost Year	1998
Total Direct Capital Costs ($)	$DC(\$) = \frac{\$950}{\frac{MMBtu}{hr}} \; Q_B \left(\frac{2375 \frac{MMBtu}{hr}}{Q_B} \right)^{0.577} \left(0.66 + 0.85 \, \eta_{NO_x} \right)$
Indirect Installation Costs General Facilities ($) Engineering and Home Office Fees ($) Process Contingency ($)	 0.05 A 0.10 x A 0.05 x A
Total Indirect Installation Costs ($)	B = A x (0.05 + 0.1 + 0.05)
Project Contingency ($)	C= 0.15 *x (A + B)*
Total Plant Cost ($)	D= *A + B + C*
Allowance for Funds During Construction ($)	E = 0 (Assumed for SNCR)
Royalty Allowance ($)	F = 0 (Assumed for SNCR)
Preproduction Cost ($)	G = 0.02 *x (D + E)*
Inventory Capital* ($)	$H = Vol_{reagent}$ *(gal) x* $Cost_{reagent}$ *($/gal)*
Initial Catalyst and Chemicals ($)	I = 0 (Assumed for SNCR)
Total Capital Investment (TCI) ($)	$TCI = D + E + F + G + H + I$

* Cost for urea stored at site, i.e., the first fill of the reagent tanks

1.4.2 Total Annual Costs

Total annual costs (TAC) consist of direct costs, indirect costs, and recovery credits. Direct annual costs are those associated or proportional to the quantity of waste gas prodent of the operation of the control system and would be incurred even if it were shut down. No byproduct recovery credits are included because there are no salvageable byproducts generated from the SNCR [1]. Each of these costs is discussed in the sections below. A more detailed discussion of capital costs can be found in Section 1, Chapter 2 of this Manual.

Design parameters are estimated using the maximum annual heat input rate of the boiler to ensure adequate sizing of the SNCR system. Annual costs are calculated using the average heat input rate of the boiler and SNCR system using CF_{total}. This ensures that annual costs are based on the actual operating conditions rather than the design case.

Direct Annual Costs

Direct annual costs (DAC) include variable and semivariable costs. Variable direct annual costs account for purchase of reagent, utilities (electrical power and water), and any additional coal and ash disposal resulting from the operation of the SNCR. Semivariable direct annual costs include operating and supervisory labor and maintenance (labor and materials).

$$DAC = \left(\begin{matrix}Annual\\Maintenance\\Cost\end{matrix}\right) + \left(\begin{matrix}Annual\\Reagent\\Cost\end{matrix}\right) + \left(\begin{matrix}Annual\\Electricity\\Cost\end{matrix}\right) + \left(\begin{matrix}Annual\\Water\\Cost\end{matrix}\right) + \left(\begin{matrix}Annual\\Coal\\Cost\end{matrix}\right) + \left(\begin{matrix}Annual\\Ash\\Cost\end{matrix}\right) \qquad (1.20)$$

Operating and Supervisory Labor

In general, no additional personnel is required to operate or maintain the SNCR equipment for large industrial facilities. Therefore, the cost of operating or supervisory labor is assumed be negligible.

Maintenance

The annual maintenance labor and material cost, including nozzle tip replacement for the injectors, is assumed to be 1.5% of the Total Capital Investment, *TCI*, in dollars. The equation is for annual maintenance cost in $/hr, *AMC*, given by:

$$Annual\ maintenance\ cost = 0.015\ TCI \qquad (1.21)$$

Reagent Consumption

The annual cost of reagent purchase in $/yr is estimated using the aqueous reagent volume flow rate in gph q_{sol}, the total operating time t_{op} in hours.

$$Annual\ reagent\ cost = q_{sol}\ Cost_{reag}\ t_{op} \tag{1.22}$$

were $Cost_{reag}$ is the cost of reagent in dollars per gallon ($/gal) and t_{op} in hours can be calculated using the total capacity fuel or CF_{total}:

$$t_{op} = CF_{total}\ 8760 \frac{hr}{yr} \tag{1.22a}$$

Utilities

The electrical power consumption in kilowatts (kw) estimated for SNCR operations is derived in Appendix B of the draft EPA report, Selective Noncatalytic Reduction for NO_x Control on Coal-fired Boilers [1]. It is based on linear regression of electrical power consumption data correlated to the uncontrolled NO_x concentration in lb/MMBtu, NO_{xin}, the *NSR* and the boiler heat input in MMBtu/hr, Q_B.

$$P = \frac{0.47\ NO_{x\,in}\ NSR\ Q_B}{9.5} \tag{1.23}$$

Using the estimated power consumption, *Power*, the annual cost of electricity is estimated from the following equation:

$$Annual\ electricity\ cost = P\ Cost_{elect}\ t_{op} \tag{1.24}$$

where $Cost_{elect}$ is the cost of electricity in dollars per kW ($/kw) and t_{op} is defined in equation 1.22a.

Water Consumption

The volumetric flowrate of water for diluting the urea is calculated from the aqueous urea mass flow rate in lb/hrand the concentration of the aqueous urea during storage, $C_{urea\ sol\ stored}$ and the average percent concentration of the injected urea, $C_{urea\ sol\ inj}$. The flowrate, q water, in gallons per hour (gph)is:

$$q_{water} = \frac{\dot{m}_{sol}}{\rho_{water}} \left(\frac{C_{urea\ sol_{stored}}}{C_{urea\ sol\ inj}} - 1 \right) \qquad (1.25)$$

where ρ_{water} is the density of water, 8.345 lb/gal. For urea dilution from a 50% solution to a 10% solution, Equation (1.25) becomes:

$$q_{water} = \frac{4\ \dot{m}_{sol}}{\rho_{water}} \qquad (1.26)$$

Using this estimate for the volume flow rate of water (gph) the annual cost of water consumption in $/yr is given by:

$$Annual\ water\ cost = q_{water}\ Cost_{water}\ t_{op} \qquad (1.27)$$

where $Cost_{water}$ is the cost of water in dollars per gallon ($/gal) and t_{op} is given by equation 1.22a.

Coal and Ash

The additional coal required as a result of the heat used to evaporate the water in the injected solution (water in the stored urea solution and the dilution water) is estimated using the following equation:

$$\Delta Coal \left(\frac{Btu}{hr}\right) = Hv \left(\frac{Btu}{lb}\right) \times \dot{m}_{reagent} \left(\frac{lb}{hr}\right) \times \left(\frac{1}{C_{urea\ sol\ inj}} - 1 \right) \qquad (1.28)$$

where $C_{urea\ sol\ inj}$ is the percent concentration of the injected aqueous urea agent and $\dot{m}$ reagent is the mass flowrate in lb/hr. H_v, the approximate heat of vaporization of water at 310°F, is 900 Btu/lb, which is a representative temperature for flue gas exiting the air heater.

Although the water from the urea solution is evaporated in the furnace at higher temperatures (due to urea injection in the furnace zones at over 1,500°F), the temperature at the air heater exit is used because it is the thermodynamic end point of the combustion process. The quantity of fuel burned in the boiler depends on the boiler efficiency, which, in turn, depends on the air heater exit temperature and the moisture in the air heater exit gas. The boiler is fired to maintain the required steam flow (e.g., for the steam turbine). Because the water from the urea solution evaporates in the boiler, the boiler efficiency decreases. Consequently, more fuel needs to be burned to maintain the required steam flow.

With urea as the reagent, injected as a 10% solution and $H_v = 900$ lb/Btu, Equation (1.28) in MMBtu per hour becomes:

$$\Delta Coal\left(\frac{MMBtu}{hr}\right) = \frac{900\left(\frac{Btu}{lb}\right)\dot{m}_{reagent}\left(\frac{lb}{hr}\right)9}{10^6\left(\frac{Btu}{MMBtu}\right)} \tag{1.29}$$

As a result of burning additional coal, additional ash is generated. This ash must be disposed or sold as byproduct. This cost methodology assumes that the ash is disposed. The estimated additional ash to be disposed in lb/hr is given by:

$$\Delta Ash = \frac{\Delta Coal \;\; ash\ product}{HHV} \tag{1.30}$$

where *ash product* is the fraction of ash produced as a byproduct of burning a given type of coal to and *HHV* is the high heating value of the coal in Btu/lb as given in Table 1.3.

The cost of the additional coal in $/yr required to maintain the same boiler steam output is:

$$Annual\ \Delta Coal\ cost = \Delta Coal\ Cost_{coal}\ t_{op} \tag{1.31}$$

where $Cost_{coal}$ is the cost of coal in dollars per MMBtu ($/MMBtu) Coal is in lb/hr, and t_{op} is given in equation 1.22a.

The cost of additional ash disposal due to the addition fuel usage is given by:

$$Annual\ \Delta Ash\ cost \quad = \quad \Delta Ash\ Cost_{ash}\ \ t_{op} \tag{1.32}$$

where $Cost_{ash}$ is the cost of ash disposal in dollars per ton ($/ton). Ash is in lb/hr and t is given equation 1.22a. A conversion of 2,000 lb per ton is required to convert $/ton to ¢/lb.

Indirect Annual Costs

In general, indirect annual costs (fixed costs) include the capital recovery cost, property taxes, insurance, administrative charges, and overhead. Capital recovery cost is based on the anticipated equipment lifetime and the annual interest rate employed. An economic lifetime of 20 years is assumed for the SNCR system. (The remaining life of the boiler may also be a determining factor)

In many cases property taxes do not apply to capital improvements such as air pollution control equipment, therefore, for this analysis, taxes are assumed to be zero [9]. The cost of overhead for an SNCR system is also considered to be zero. An SCR system is not viewed as risk-increasing hardware (e.g., a high energy device such as a turbine). Consequently, insurance on an SNCR system is on the order of a few pennies per thousand dollars annually [9]. The administrative charges, covering sales, research and development, accounting, and other home office expenses, incurred in operation of an SNCR system are not significant for the cost estimation procedure presented here. Finally, there are two categories of overhead, payroll and plant. Payroll overhead includes expenses related to labor employed in operation and maintenance of hardware; whereas plant overhead accounts for items such as plant protection, control laboratories, and parking areas. Because this procedure assumes that no additional labor is needed to operate an SNCR system, payroll overhead is zero and plant overhead is considered negligible.

Using these assumptions, indirect annual costs in $/yr, *IDAC*, can be expressed as:

$$IDAC = CRF\ TCI \tag{1.33}$$

where *TCI* is the total capital investment in dollars and CRF is the capital recovery factor.

Capital recovery factor was defined in Section 1 as:

$$CRF = \frac{i(1+i)^n}{(1+i)^n - 1} \tag{1.34}$$

where i is the assumed interest rate and n is the economic life of the SNCR system.

Total Annual Cost

The total annual cost, *TAC*, for owning and operating an SNCR system is the sum of direct and indirect annual costs as given in the following equation:

$$TAC = DC + IDAC \tag{1.35}$$

The tons of NO_x removed annually are:

$$NO_x\ Removed = NO_{x_{in}}\ \eta_{NO_x}\ Q_B\ t_{op} \tag{1.36}$$

where NO_{xin} is given as lb/MMBtu, Q_B as MMBtu/hr and top in hours and a conersion of 2,000 lb/ton is required.

The cost in dollars per ton of NO_x removed per year, cost efficiency is:

$$Cost\ Efficiency = \frac{TAC}{NO_x\ Removed} \tag{1.37}$$

1.5 Example Problem

An example problem, which calculates both the design parameters and capital and annual costs, is presented below. The design basis is a retrofit SNCR system being applied to a 1000 MMBtu/hr, wall-fired, industrial boiler firing subbituminous coal. The following assumptions are made to perform the calculations:

Fuel High Heating Value	10,000 Btu/lb
Maximum Fuel Consumption Rate	1.0×10^5 lb/hr
Average Annual Fuel Consumption	4.38×10^8 lb
Number of SNCR operating days	155 days
Uncontrolled NO_x Concentration	0.46 lb/MMBtu
Required Controlled NO_x Concentration	0.30 lb/MMBtu
Percent Fuel Ash Weight	7.5%
Stored Urea Concentration	50% urea solution
Injected Urea Concentration	10% urea solution
Number of Days of Storage for Urea	14 days

In addition to these assumptions, the estimated economic factors required to perform the calculations are taken from References [2] and [3]. These estimates are:

Cost Year	December, 1998
Equipment Life	20 years
Annual Interest Rate	7%
Coal Cost, Eastern Bituminous	1.60 $/MMBtu
Ash Disposal Cost	11.28 $/ton
50% Urea Solution Cost	0.85 $/gal
Water Consumption Cost	0.0004 $/gal

1.5.1 Design Parameter Example

The boiler annual heat input rate, Q_B, is calculated from the High Heating Value for subbituminous coal given in Table 1.3 and the maximum fuel consumption rate, $\dot{m}_{fuel}$:

$$Q_B = \frac{10{,}000\ \frac{Btu}{lb} \times 100{,}000\ \frac{lb}{hr}}{10^6\ \frac{Btu}{MMBtu}} = 1{,}000\ \frac{MMBtu}{lb}$$

The plant capacity factor is calculated from the maximum and annual average fuel consumption:

$$CF_{plant} = \frac{4.38 \times 10^8\ lb}{1 \times 10^5 \left(\frac{lb}{hr}\right) \times 8760\ \frac{hr}{yr}} = 0.5 = 50\%$$

The SNCR system capacity factor is calculated from the months of SNCR operation, 5 months:

$$CF_{SNCR} = \frac{155\ days}{365\ days} = 0.42 = 42\%$$

The total capacity factor including both plant and SNCR capacity factors is given by:

$$CF_{toal} = 0.5 \times 0.42 = 0.21 = 21\%$$

The NO_x removal efficiency, η_{NOx}, is calculated from the inlet NO_x concentration and the required stack NO_x concentration:

$$\eta_{NO_x} = \frac{0.46\ \frac{lb}{MMBtu} - 0.30\ \frac{lb}{MMBtu}}{0.46\ \frac{lb}{MMBtu}} = 0.35 = 35\%$$

The equation derived in Reference [1] is used to estimate the NSR for the SNCR system. The estimate is given by:

$$NSR = \frac{\left[\left(2 \times 0.46\ \frac{lb}{MMBtu}\right) + 0.7\right] \times 0.35}{0.46\ \frac{lb}{MMBtu}} = 1.23$$

The reagent utilization can then be calculated based on the required NO_x removal efficiency and NSR value using:

$$Utilization = \frac{0.35}{1.23} = 0.28 = 28\%$$

The value of the the NSR indicates that 1.23 moles of NH_3 are required per mole of uncontrolled NO_x to reduce the NO_x level by 35%. This translates to a reagent utilization of 0.28, the ratio of moles of reagent reacted to the moles injected. This indicates that 28% of the injected reagent is being utilized for NO_x removal. The remainder of the reagent is being destroyed or passing through as ammonia slip.

The mass flow rate of the reagent is calculated using the molecular weight of the reagent, 60.06 g/mole and NO_2, 46.01g/mole and the SR_T for urea, 2. For an NSR of 1.23, the reagent mass flow rate is given by:

$$\dot{m}_{reagent} = \frac{0.46\frac{lb}{MMBtu} \times 1000\frac{MMBtu}{hr} \times 0.35 \times 1.23 \times 60.06\frac{g}{mole}}{2 \times 46.01\frac{g}{mole}} = 130\frac{lb}{hr}$$

The flow rate of the diluted solution, where the concentration of the aqueous solution is 50% urea, is given by:

$$\dot{m}_{sol} = \frac{130\frac{lb}{hr}}{0.50} = 260\frac{lb}{hr}$$

The solution volume flow rate can then be calculated where ρ is the density of the aqueous reagent solution, 71.0 lb/ft^3 for 50% aqueous urea solution at 60°F.

$$q_{sol} = \frac{260\frac{lb}{hr} \times 7.481\frac{gal}{ft^3}}{71.0\frac{lb}{ft^3}} = 27\,gph$$

The total volume stored in the tank, or tanks, is based on the volume that the SNCR system requires for 14 days of operation. The on site storage requirement is given by:

$$Volume_{tank} = 27\ gph \times 14\ days \times \frac{24hr}{day} = 9072\ gal$$

The on site storage requirement for urea is 9,072gallons per 14 days. This shows that for a large boiler, 1000 MMBtu/hr with 35% NO_x removal efficiency the volume of urea required to operate an SNCR system for155 days during summer months is approximately 10,500 gallons.

An estimates for power consumption is given by:

$$Power = \frac{0.47 \times 0.46 \frac{lb}{MMBtu} \times 1000 \frac{MMBtu}{hr}}{9.5} = 23\ kW$$

Water consumption, assuming a 50% urea solution stored and a 10% urea solution injected, is:

$$q_{water} = \frac{260 \frac{lb}{hr}}{8.345 \frac{lb}{gal}} \times \left(\frac{0.5}{0.10} - 1 \right) = 125\ gph$$

The estimated additional coal and ash required to maintain the same net heat output are given by:

$$\Delta Coal = \frac{260 \frac{lb}{hr} \times 900 \frac{Btu}{lb}}{10^6 \frac{Btu}{MMBtu}} \times \left(\frac{1}{0.10} - 1 \right) = 2 \frac{MMBtu}{hr}$$

$$\Delta Ash = \frac{2 \frac{MMBtu}{hr} \times 0.077 \times 10^6 \frac{Btu}{MMBtu}}{10{,}000 \frac{Btu}{lb}} = 15 \frac{lb}{hr}$$

1.5.2 Cost Estimation Example

Once the SNCR system is sized, the capital and annual costs for the SNCR system can be estimated. The direct capital costs are estimated using:

$$DC = \frac{\$950}{MMBtu} \times 1000 \left(\frac{MMBtu}{hr} \right) \times \left(\frac{2375 \frac{MMBtu}{hr}}{1000 \frac{MMBtu}{hr}} \right)^{0.577} \times [0.66 + (0.85 \times 0.35)] = \$1{,}498{,}152$$

Indirect capital costs are estimated from the direct capital costs using the following equations:

Indirect Installation Costs	
General Facilities ($)	$1,498,152 x 0.05 = $74908
Engineering and Home Office Fees ($)	$1,498,152 x 0.10 = $149,815
Process Contingency ($)	$1,498,152 x 0.05 = $74908
Total Indirect Installation Costs ($)	$74908 + $149,815 + $74908 = $299,630
Project Contingency ($)	($1,498,152 + $299,630) x 0.15 = $269,667
Total Plant Cost ($)	$149,815 + $299,630 + $269,667 = $719,112
Preproduction Cost ($)	$719,112 x 0.02 = $143,82
Inventory Capital* ($)	9072 gal x 0.85 $ / gal = $7,711
Total Capital Investment (TCI) ($)	$719,112 + $143,82 + $$7,711 = $741,205

Annual costs are based on the economic factors listed above. In addition, the SNCR system is assumed to operate for 5 months of the year with a boiler loading of 65%, resulting in a total capacity factor of 27%. The annual variable costs are given by:

$$Annual\ maintenance\ cost = 0.015 \times \$741{,}205 = \frac{\$11{,}118}{yr}$$

$$Annual\ reagent\ cost = 27 gph \times 0.85 \frac{\$}{gal} \times \left[0.21 \times 8760 \frac{hr}{yr}\right] = \frac{\$42{,}218}{yr}$$

$$Annual\ electricity\ cost = 23 kW \times 0.05 \frac{\$}{kW} \left[0.21 \times 8760 \frac{hr}{yr}\right] = \frac{\$2{,}115}{yr}$$

$$Annual\ water\ cost = 125 gph \times 0.0004 \frac{\$}{gal} \times \left[0.21 \times 8760 \frac{hr}{yr}\right] = \frac{\$92}{yr}$$

$$Annual\ \Delta Coal\ cost = 2 \frac{MMBtu}{hr} \times 1.60 \frac{\$}{MMBtu} \times \left[0.21 \times 8760 \frac{hr}{yr}\right] = \frac{\$5{,}887}{yr}$$

$$Annual\ \Delta Ash\ cost = \frac{15 \frac{lb}{hr} \times 11.28 \frac{\$}{ton} \times \left[0.21 \times 8760 \frac{hr}{yr}\right]}{2{,}000 \frac{lb}{ton}} = \frac{\$156}{yr}$$

The total variable direct annual cost, the sum of the cost of the reagent, electricity, water, coal and ash, is given by:

$$Total\ Variable\ Cost = (42{,}218 + 2{,}115 + 92 + 5{,}887 + 156)\frac{\$}{yr} = \frac{\$50{,}468}{yr}$$

and the total direct annual cost is given by the sum of the maintenance and variable cost:

$$Total\ Direct Annual\ Cost\ (DAC) = \frac{\$31{,}748}{yr} + \frac{\$50{,}468}{yr} = \frac{\$82{,}216}{yr}$$

Total indirect annual costs assumes that the property tax factor, F_{tax}, and the overhead factor, F_{ovhd}, are both zero. The capital recovery factor, CRF, is defined by:

$$CRF = \frac{0.07\ (1+0.07)^{20}}{(1+0.07)^{20} - 1} = 0.09439$$

and the indirect annual costs (IDAC) are calculated by:

$$IDAC = 0.09439 \times \$741{,}205 = \frac{\$69{,}962}{yr}$$

The total annual cost is the sum of the direct annual and indirect annual costs given by:

$$TAC = \frac{\$82{,}216}{yr} + \frac{\$69{,}962}{yr} = \frac{\$152{,}178}{yr}$$

The annual cost in terms of NO_x removed can be calculated using the total annual cost and the tons of NO_x removed annually by:

$$NO_x\ Removed = \frac{0.46\frac{lb}{MMbtu} \times 0.35 \times 1000\frac{MMBtu}{hr}\left[0.21 \times 8670\frac{hr}{yr}\right]}{2000\frac{lb}{ton}} = 147\ tons$$

and

$$NO_x\ Removed = \frac{0.46\frac{lb}{MMbtu} \times 0.35 \times 1000\frac{MMBtu}{hr}\left[0.21 \times 8670\frac{hr}{yr}\right]}{2000\frac{lb}{ton}} = 147\ tons$$

References

[1] Selective Noncatalytic Reduction for NO_x Control on Coal-fired Boilers, Draft Report, prepared for the U.S. Environmental Protection Agency by The Cadmus Group, Inc., Bechtel Power Corporation, and Science Applications International Corporation. May 1998.

[2] Investigation of Performance and Cost of NO_x Controls as Applied to Group 2 Boilers. Revised Draft Report, prepared for the U.S. Environmental Protection Agency by The Cadmus Group, Inc., Bechtel Power Corporation, and Science Applications International Corporation. August 1996.

[3] Cost Estimates for Selected Applications of NO_x Control Technologies on Stationary Combustion Boilers, Draft Report and Responses to Comments on the Draft Report. prepared for the U.S. Environmental Protection Agency by The Cadmus Group, Inc., and Bechtel Power Corporation. June 1997.

[4] Analyzing Electric Power Generation Under the CAAA. EPA Office of Air and Radiation. March 1998.

[5] Committee of the Institute of Clean Air Companies, Inc. White Paper, Selective Non-Catalytic Reduction (SNCR) for Controlling NO_x Emissions. SNCR. October 1997.

[6] Tanner Industries, Inc. Storage & Handling of Ammonium Hydroxide. Philadelphia, PA. October 1995.

[7] Electric Power Research Institute (EPRI). Technical Assessment Guide, Vol. 1. Revision 7, 1993.

[8] Northeast States for Coordinated Air Use Management (NESCAUM). Status Report on NO_x Control Technologies and Cost Effectiveness for Utility Boilers. June 1998.

[9] United States Patent, Method for the Reduction of the Concentration of NO In Combustion Effluents Using Ammonia. Inventor: Richard K. Lyon, Patent Number 3,900,554. August 19, 1975.

[10] Rini, M.J., J.A. Nicholson, and M.B. Cohen. Evaluating the SNCR Process for Tangentially-Fired Boilers. Presented at the 1993 Joint Symposium on Stationary Combustion NO_x Control, Bal Harbor, Florida. May 24 - 27, 1993.

[11] Institute of Clean Air Companies, (ICAC). Proceedings from ICAC Forum: Cutting NO_x Emissions. March, 2000.

[12] State and Territorial Air Pollution Program Administrators (STAPPA) and Association

of Local Air Pollution Conrol Officials (ALAPCO). Controlling Nitrous Oxides. July 1994.

[13] Personal communication with Mike Knenlein of Fuel Tech to Paula Hemmer (Pechan-Avanti) on April 2, 2000.

[14] Majors, Russ K., Dr. Russell Hill, Robert McMurry, and Scott Thomas, A Study of the Impact of Ammonia Injection on Marketable Fly Ash Including Quality Control Procedures.

EPA/452/B-02-001

Section 4

NO_x Controls

EPA/452/B-02-001

Section 4.2

NO_x Post- Combustion

Introduction

Nitrogen oxides (NO_x) are gaseous pollutants that are primarily formed through combustion process. While flue gas is within the combustion unit, about 95% of the NO_x exists in the form of nitric oxide (NO). The balance is nitrogen dioxide (NO_2), which is unstable at high temperatures. Once the flue gas is emitted into the atmosphere, most of the NO_x is ultimately converted to NO_2. NO_x in the atmosphere reacts in the presence of sunlight to form ozone (O_3), one of the criteria pollutants for which health-based National Ambient Air Quality Standards have been established. Since ozone formation requires sunlight and high temperatures, ozone formation is greatest in summer months.

NO_x is generated in one of three forms; fuel NO_x, thermal NO_x, and prompt NO_x. Fuel NO_x is produced by oxidation of nitrogen in the fuel source. Combustion of fuels with high nitrogen content such as coal and residual oils produces greater amounts of NO_x than those with low nitrogen content such as distillate oil and natural gas. Thermal NO_x is formed by the fixation of molecular nitrogen and oxygen at temperatures greater than 3600°F (2000°C). Prompt NO_x forms from the oxidation of hydrocarbon radicals near the combustion flame and produces an insignificant amount of NO_x.

Selective Noncatalytic Reduction (SNCR) and Selective Catalytic Reduction (SCR) are post-combustion control technologies based on the chemical reduction of nitrogen oxides (NO_x) into molecular nitrogen (N_2) and water vapor (H_2O). The primary difference between the two technologies is that SCR utilizes a catalyst to increase the NO_x removal efficiency, which allows the process to occur at lower temperatures. The technologies can be used separately or in combination with other NO_x combustion control technologies such as low NO_x burners (LNB) and natural gas reburn (NGR). SNCR and SCR can be designed to provide NO_x reductions year-round or only during summer months, when ozone concerns are greatest.

This section presents design specifications and a costing methodology for SNCR and SCR applications for large industrial boilers (greater than 250 MMBtu/hr). Its primary purpose is to provide study-level cost estimates. These estimates can be used to compare the approximate costs of SNCR, SCR, and alternative NO_x control technologies. The costing methodologies are based on cost estimations for SNCR and SCR applications to utility boilers, which were developed by the Office of Research and Development (ORD), U.S. Environmental Protection Agency (EPA), Research Triangle Park, NC.

As a prelude to the cost methodology, the section describes the process chemistry, performance parameters, and system components of SNCR and SCR. In addition, impacts to the boiler performance and facility operations resulting from the installation of SNCR and SCR are presented. The section also estimates important underlying design parameters including the normalized stoichiometric ratio, catalyst volume, and reagent consumption. Lastly, it presents

assumptions and equations for estimating capital costs, annual operation and maintenance costs, and annualized costs. SNCR and SCR are discussed in separate chapters, however, the chapter on SCR builds on the concepts discussed in the SNCR chapter.

Information on key aspects of the design of SNCR and SCR systems is considered proprietary by vendors, including methods for estimating certain design parameters and costs. This information is not available to develop cost methodologies for SNCR and SCR. In order to obtain estimates of the proprietary design parameters and costs, it is necessary to develop mathematical correlations from available data using regression and curve fitting techniques. These expressions are derived in EPA reports prepared by The Cadmus Group, Bechtel Power, Inc. and SAIC in SNCR and SCR References [1, 2 and 3] from documented comprehensive SNCR and SCR performance data and costs based on quotations provided by suppliers and facilities.

EPA/452/B-02-001

Chapter 2

Selective Catalytic Reduction

Daniel C. Mussatti
Innovative Strategies and Economics Group
Air Quality Strategies and Standards Division
Office of Air Quality Planning and Standards
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

Dr. Ravi Srivastava
Office of Research and Development
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

Paula M. Hemmer and
Randy Strait
E.H. Pechan & Associates, Inc.
Durham, NC 27707

October 2000

Contents

2.1 Introduction 2-3

2.2 Process Description 2-4
2.2.1 Sampling Systems 2-5
2.2.2 SCR Performance Parameters 2-9
2.2.3 SCR System Configurations 2-18
2.2.4 High Dust SCR System Equipment 2-23
2.2.5 SCR System Auxiliary Equipment 2-26
2.2.6 Other Considerations 2-28

2.3 Design Parameters 2-30

2.4 Cost Analysis 2-40
2.4.1 Total Capital Investment 2-41
2.4.2 Total Annual Costs 2-44

2.5 Example Problem 2-50
2.5.1 Design Parameter Example 2-51
2.5.2 Cost Estimation Example 2-54

References 2-59

2.1 Introduction

Selective Catalytic Reduction (SCR) has been applied to stationary source, fossil fuel-fired, combustion units for emission control since the early 1970s and is currently being used in Japan, Europe, and the United States. It has been applied to large (≥250 million British thermal units per hour (MMbtu/hr)) utility and industrial boilers, process heaters, and combined cycle gas turbines. There has been limited application of SCR to other combustion devices and processes such as simple cycle gas turbines, stationary reciprocating internal combustion engines, nitric acid plants, and steel mill annealing furnaces [4]. In the United States, SCR has mainly been applied to coal- and natural gas-fired electrical utility boilers ranging in size from 250 to 8,000 MMbtu/hr (25 to 800 megawatts (MW)). SCR can be applied as a stand-alone NO_x control or with other technologies such as combustion controls. SCR systems have experienced relatively few operational or maintenance problems [1].

SCR is typically implemented on stationary source combustion units requiring a higher level of NO_x reduction than achievable by SNCR or combustion controls. Theoretically, SCR systems can be designed for NO_x removal efficiencies up to 100 percent (%). Commercial coal-, oil- and natural gas-fired SCR systems are often designed to meet control targets of over 90%. However, maintaining this efficiency is not always practical from a cost standpoint. In practice, SCR systems operate at efficiencies in the range of 70% to 90%. [4]

SCR capital costs vary by the type of unit controlled and the fuel source. Capital costs for retrofit of coal-fired electric utility boilers range from 50[1] to 70[2] dollars per kilowatt ($/kW) for wall- and tangential-fired boilers and 50[1] to 80[1] $/kW for cyclone and wet bottom boilers. New utility boiler capital costs are generally less than 40[1] $/kW. Retrofit of oil- and gas- fired utility boilers range from 25[1] to 30[1] $/kW since they have lower NO_x and particulate emissions. Typical operation and maintenance costs are less than 0.1[1] cents per kilowatt-hour. Capital costs for installing SCR on new industrial boilers range from 4000[1] to 6000[1] $/MMBtu/hr for small oil and gas-fired units to over 10,000[1] $/MMBtu/hr for large coal-fired units. Gas turbines have capital costs of 30[1] to 100[1] $/kW depending on the size of the turbine. The cost of retrofitting a gas-fired reciprocating internal combustion engine is approximately 125[1] $/horsepower, which is less than retrofit of combustion control technology for this type of unit. [4]

Although the cost data used in this report are based on retrofit applications of SCR systems on existing utility coal-fired boilers, this report's costing procedure accounts for the major differences between a new and retrofit installation. Therefore, the costing procedure is suitable for retrofit or new boiler applications of SCR on all types of large (≥250

[1] Cost are in 1997 dollars

MMBtu/hr) coal-fired industrial boilers. The costing methodology incorporates certain approximations, consequently, it should be used to develop study-level (± 30%) cost estimates of SCR applications.

2.2 Process Description

Like SNCR, the SCR process is based on the chemical reduction of the NO_x molecule. The primary difference between SNCR and SCR is that SCR employs a metal-based catalyst with activated sites to increase the rate of the reduction reaction. A nitrogen based reducing agent (reagent), such as ammonia or urea, is injected into the post combustion flue gas. The reagent reacts selectively with the flue gas NO_x within a specific temperature range and in the presence of the catalyst and oxygen to reduce the NO_x into molecular nitrogen (N_2) and water vapor (H_2O).

The use of a catalyst results in two primary advantages of the SCR process over SNCR. The main advantage is the higher NO_x reduction efficiency. In addition, SCR reactions occur within a lower and broader temperature range. However, the decrease in reaction temperature and increase in efficiency is accompanied by a significant increase in capital and operating costs. The cost increase is mainly due to the large volumes of catalyst required for the reduction reaction.

Figure 2.1 shows a simplified process flow schematic for SCR. Reagent is injected into the flue gas downstream of the combustion unit and economizer through an injection grid mounted in the ductwork. The reagent is generally diluted with compressed air or steam to aid in injection. The reagent mixes with the flue gas and both components enter a reactor chamber containing the catalyst. As the hot flue gas and reagent diffuse through the catalyst and contact activated catalyst sites, NO_x in the flue gas chemically reduces to nitrogen and water. The heat of the flue gas provides energy for the reaction. The nitrogen, water vapor, and any other flue gas constituents then flow out of the SCR reactor. More detail on the SCR process and equipment is provided in the following sections.

There are several different locations downstream of the combustion unit where SCR systems can be installed. Flue gas temperature and constituents vary with the location of the SCR reactor chamber. SCR reactors located upstream of the particulate control device and the air heater have higher temperatures and higher levels of particulate matter. An SCR reactor located downstream of the air heater, particulate control devices and flue gas desulfurization system is essentially dust- and sulfur-free but its temperature is generally below the acceptable range. In this case, reheating of the flue gas may be required which significantly increases the SCR operational costs. Section 2.2.3 discusses the various SCR system configurations.

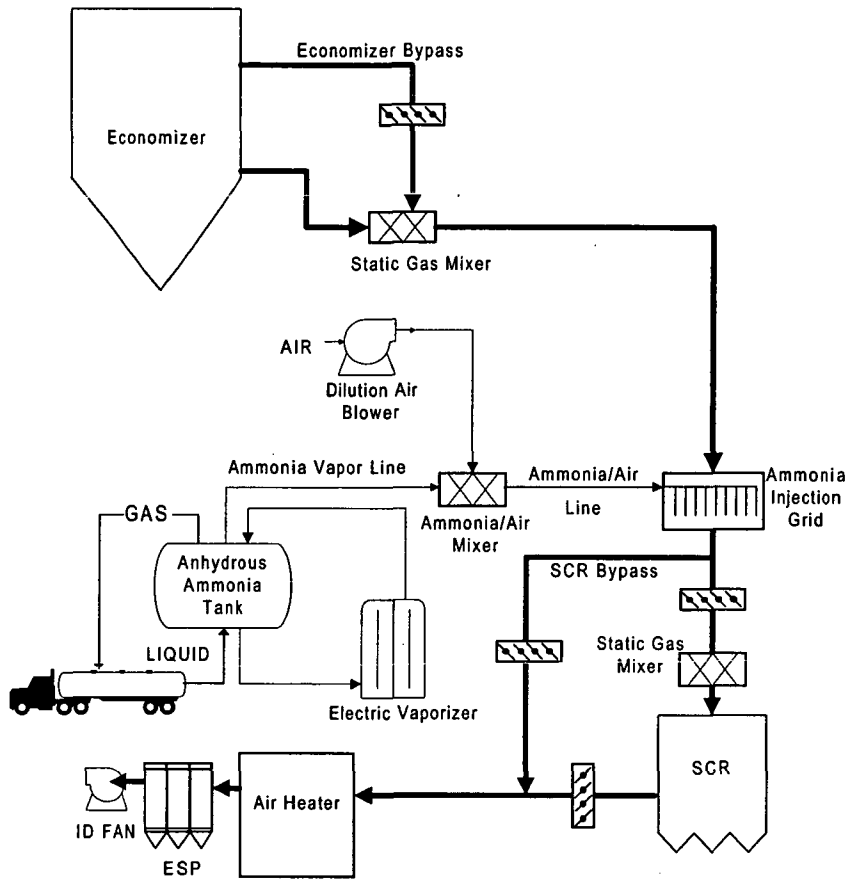


Figure 2.1: SCR Process Flow Diagram [6, 22]

2.2.1 Reduction Chemistry, Reagents and Catalyst

The reducing agent employed by the majority of SCR systems is gas-phase ammonia (NH_3) since it penetrates the catalyst pores more readily than aqueous urea. The ammonia, either in anhydrous or aqueous form, is vaporized before injection by a vaporizer. Within the appropriate temperature range, the gas-phase ammonia then decomposes into free radicals including NH_3 and NH_2. After a series of reactions, the ammonia radicals come

into contact with the NO_x and reduce it to N_2 and H_2O. The global representation of these reactions is given below. Note that the NO_x is represented as nitrogen oxide (NO) since it is the predominant form of NO_x within the boiler. The ammonia reaction equation is represented by:

$$2NO + 2NH_3 + \frac{1}{2}O_2 \xrightarrow{catalyst} 2N_2 + 3H_2O \qquad (2.1)$$

The equation indicates that 1 mole of NH_3 is required to remove one mole of NO_x. The catalyst lowers the required activation energy for the reduction reaction and increases the reaction rate. In the catalytic reaction, activated sites on the catalyst rapidly adsorb ammonia and gas phase nitric oxide to form an activated complex. The catalytic reaction, represented by Equation 2.1, occurs resulting in nitrogen and water, which are then desorbed to the flue gas. The site at which the reaction occurs is then reactivated via oxidation.

The high temperature of the flue gas converts the ammonia to free radicals and provides the activation energy for the reaction. The reaction also requires excess oxygen, typically 2 to 4 percent, to achieve completion. NO_x reduction with ammonia is exothermic, resulting in the release of heat. However, because the NO_x concentration in the flue gas at the inlet of the SCR is typically 0.02 to 0.01 percent by volume, the amount of heat released is correspondingly small. Thermodynamic equilibrium is not a limiting factor in NO_x reduction if the flue gas is within the required temperature range. [1]

Reagent

The SCR system can utilize either aqueous or anhydrous ammonia for the reduction reaction. Anhydrous ammonia is nearly 100% pure ammonia. It is a gas at normal atmospheric temperature, therefore, it must be transported and stored under pressure. Anhydrous ammonia often requires special permits for transportation and storage.

SCR applications using aqueous ammonia generally transport and store it at a concentration of 29.4% ammonia in water, though some recent applications utilize a 19% solution [1]. The use of aqueous ammonia reduces transport and storage problems. In addition, certain locations may not require permits for concentrations below 28%. Aqueous ammonia however, requires more storage capacity than anhydrous ammonia. While the 29.4% solution has substantial vapor pressure at normal air temperatures, to provide sufficient ammonia vapor to the SCR system generally requires a vaporizer.

The cost of the reagent impacts the annual costs of an SCR system. Depending on the local permitting requirements and on owner preference, either anhydrous or aqueous ammonia may be used in the SCR process. This presentation is valid for anhydrous or aqueous ammonia, however, the design parameters and cost analysis assume aqueous ammonia as a reagent since it is more commonly employed. Table 2.1 gives the properties of

anhydrous ammonia and the properties of a 29.4% aqueous ammonia solution.

Table 2.1: Ammonia Reagent Properties

Property	Anhydrous Ammonia [15,16]	Aqueous Ammonia
Liquid or gas at normal air temperature	Liquid	Liquid
Concentration of reagent normally supplied	99.5 % (by weight)	29.4 % (by weight of NH_3)
Molecular weight of reagent	17.03	17.03 (as NH_3)
Ratio of ammonia to Solution	99.5% (by weight of NH_3)	29.4 % (by weight of NH_3)
Density of liquid at 60 °F	5.1 lb/gal	7.5 lb/gal
Vapor pressure @ 80 °F	153 psia	14.6 [15, p. 3]
Flammability limits in air	16-25 % NH_3 (by volume)	16-25 % NH_3 (by volume)
Short-term exposure limit	35 ppm	35 ppm
Odor	Pungent odor @ 5 ppm or more	Pungent odor @ 5 ppm or more
Acceptable materials for storage	Steel tank, rated for at least 250 psig pressure (no copper or copper-based alloys, etc.)	Steel tank, rated for at least 25 psig pressure (no copper or copper-based alloys, etc.)

Catalyst

SCR catalysts are composed of active metals or ceramics with a highly porous structure. Within the pores of the catalyst are activated sites. These sites have an acid group on the end of the compound structure where the reduction reaction occurs. As stated previously, after the reduction reaction occurs, the site reactivates via rehydration or oxidation. Over time, however, the catalyst activity decreases, requiring replacement of the catalyst. Catalyst designs and formulations are generally proprietary. Both the catalyst material and configuration determine the properties of the catalyst.

Originally, SCR catalysts were precious metals such as platinum (Pt). In the late 1970s, Japanese researchers used base metals consisting of vanadium (V), titanium (Ti), and tungsten (W), which significantly reduced catalyst cost. In the 1980's, metal oxides

such as titanium oxide (TiO_2), zirconium oxide (ZrO_2), vandium pentoxide (V_2O_5), and silicon oxide (SiO_2) were employed to broaden the reaction temperature range. Zeolites, crystalline alumina silicates, were also introduced for high temperature (675°F to 1000°F) applications, however, zeolites tended to be cost prohibitive.

Improvements to the catalyst formulations decrease unwanted side reactions such as sulfur oxide conversions (SO_2 to SO_3) and increase the resistance to flue gas poisons. Newer catalyst designs also increase catalyst activity, surface area per unit volume, and the temperature range for the reduction reaction. As a consequence, there is a corresponding decrease in the required catalyst volumes and increase in the catalyst operating life. For coal-fired boiler applications, SCR catalyst vendors typically guarantee the catalyst for an operating life ranging between 10,000 hours to 30,000 hours [11]. Applications using oil and natural gas have a longer operating life, over 32,000 hours [19]. In addition, operating experience indicates that actual catalyst deactivation rates are lower than the design specifications [1].

Catalyst formulations include single component, multi-component, or active phase with a support structure. Most catalyst formulations contain additional compounds or supports to give thermal and structural stability or to increase surface area [4]. Catalysts configurations are generally ceramic honeycomb and pleated metal plate (monolith) designs in a fixed-bed reactor, which provide high surface area to volume ratio. Pellet catalyst in fluidized beds are also available. Pellets have greater surface area than honeycombs or pleated plates but are more susceptible to plugging. This limits the use of pellets to clean burning fuels such as natural gas.

Catalyst elements placed in a frame form a catalyst module. The modules stack together in multiple layers to create a reactor bed of the total required catalyst volume. A typical module is 3.3 ft to 6.6 ft in area (1 m x 2 m) and 3.3 ft (1 m) in height. A crane hoists the large catalyst modules into the reactor from either the interior or exterior of the reactor, depending on the reactor design.

Catalyst replacement is infrequent, generally less than one layer per year for fixed bed designs. Most SCR manufacturers offer a disposal service. The catalyst is either reactivated for reuse or its components are recycled for other uses [4]. If the catalyst cannot be recycled or reused, the facility operator must dispose of the spent catalyst in an approved landfill. In the United States, most catalyst formulations are not considered hazardous waste [4].

Catalysts greatly accelerate the NO_x reduction reaction rate, but some catalysts have more favorable properties for a given application. Performance requirements that drive the choice of catalyst include reaction temperature range, flue gas flow rate, fuel source, catalyst activity and selectivity, and catalyst operating life. In addition, the design must consider the cost of the catalyst, including disposal costs since catalyst costs can account for 20% or

more of the capital costs for a SCR system [1].

2.2.2 SCR Performance Parameters

The rate of the reduction reaction determines the amount of NO_x removed from the flue gas. The major design and operational factors that affect the NO_x removal performance of SCR are
similar to those presented in Chapter 1 SNCR. The factors discussed previously for SNCR include the following:

- Reaction temperature range;
- Residence time available in the optimum temperature range;
- Degree of mixing between the injected reagent and the combustion gases;
- Molar ratio of injected reagent to uncontrolled NO_x;
- Uncontrolled NO_x concentration level; and
- Ammonia slip.

The majority of the discussion regarding SNCR design and operational factors is valid for the SCR process excepting small variations due to the use of a catalyst and the reaction chamber being separate from the combustion unit. Additional design and operational factors to consider, which are specific to the SCR process, include:

- Catalyst activity
- Catalyst selectivity
- Pressure drop across the catalyst
- Catalyst pitch
- Catalyst deactivation
- Catalyst management

The major differences between SNCR and SCR are discussed below.

Temperature

The NO_x reduction reaction is effective only within a given temperature range. The use of a catalyst in the SCR process lowers the temperature range required to maximize the NO_x reduction reaction. At temperatures below the specified range, the reaction kinetics decrease and ammonia passes through the boiler (ammonia slip). At temperatures above the specified range, nitrous oxide (N_2O) forms and catalyst sintering and deactivation occurs.

In an SCR system, the optimum temperature depends on both the type of catalyst utilized in the process and the flue gas composition. For the majority of commercial cata-

lysts (metal oxides), the optimum temperatures for the SCR process range from 480 °F to 800 °F (250 °C to 427 °C) [11]. Figure 2.2 is a graph of the NO_x removal efficiency as a function of temperature for a typical metal oxide-type catalyst [11]. The figure shows the rate of the NO_x removal increases with temperature up to a maximum between 700°F to 750°F (370°C to 400°C). As the temperature increases above 750°F, the reaction rate and resulting NO_x removal efficiency begin to decrease.

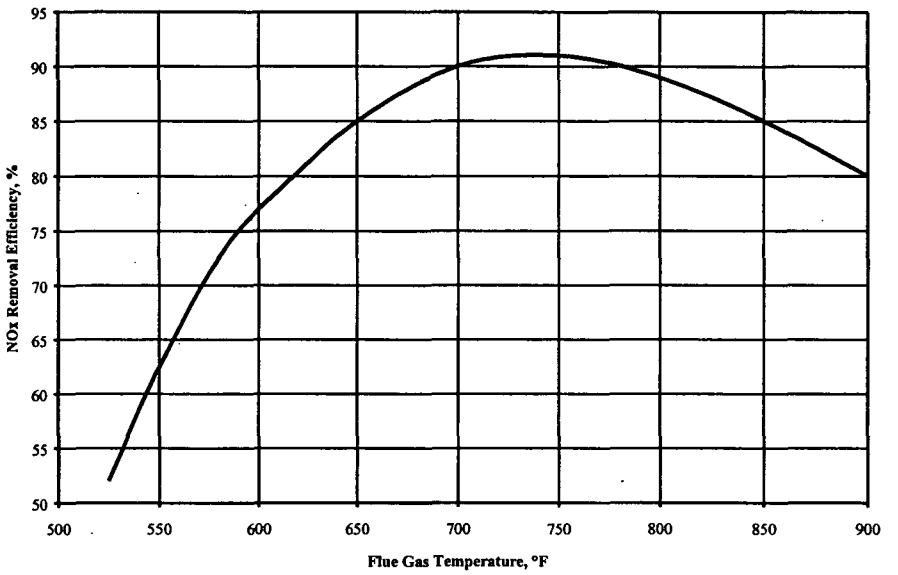


Figure 2.2: NO_x Removal versus Temperature [11]

As flue gas temperature approaches the optimum, the reaction rate increases and less catalyst volume achieves the same NO_x removal efficiency. Figure 2.3 shows the change in the required catalyst volume versus temperature [10]. There is approximately a 40% decrease in the required catalyst volume as flue gas temperature increases from 600 °F (320 °C) to the optimum range, 700 to 750 °F (370 °C to 400 °C). This decrease in catalyst volume also results in a significant decrease in capital cost for the SCR system.

The relationships between flue gas temperature, catalyst volume, and NO_x removal are complicated functions of the catalyst formulation and configuration. The physical and chemical properties of each catalyst are optimized for a different operating conditions. For a given catalyst formulation, the required catalyst volume and/or temperature range can even change from one manufacturer of the catalyst to another. The selection of catalyst, therefore, is critical to the operation and performance of the SCR system.

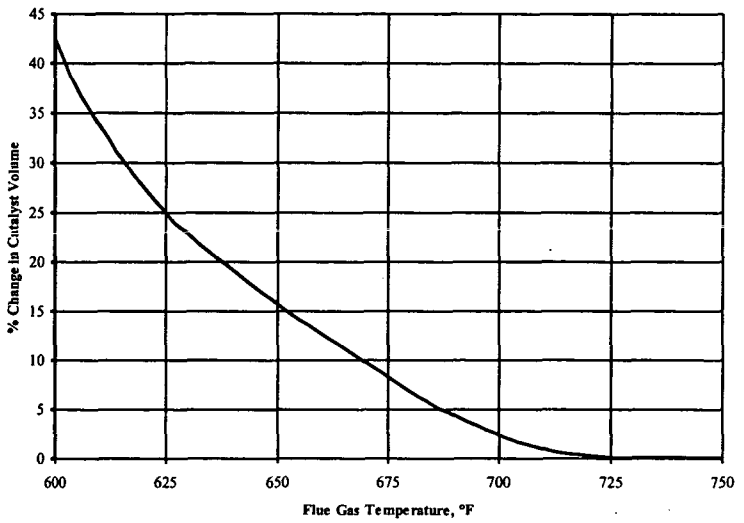


Figure 2.3: Change in Catalyst Volume vs. Temperature [10]

Since the optimum temperature window of the SCR process is lower than that of SNCR, the reagent injection into a reactor chamber occurs downstream of the combustion unit, rather than inside the combustion unit. As discussed previously, there are several options for the location of the SCR reactor. The flue gas temperature at each of these locations is different. Most designs install the reactor downstream of the economizer and prior to the air preheater, where the flue gas is at the appropriate temperature for metal oxide-based catalysts. Reheating of the flue gas may be required for reactors located downstream of the air pre-heater. Reheating significantly increases the SCR operational costs.

Boiler operation at reduced loads decreases the gas flow rate. At reduced gas flow rates, the economizer outlet gas temperature decreases because boiler heat transfer surfaces absorb more heat from the flue gas. Typical SCR systems tolerate temperature fluctuations of ± 200°F (± 93°C) [1]. At low boiler loads, however, the temperature can decrease below the optimum range. For example, a coal-fired utility boiler has an economizer exit flue gas temperature of 690°F (366°C) at 100% load, but only 570°F (300°C) at 50% load [1]. For low-load operations, an economizer bypass can be used to raise the flue gas temperature. An economizer bypass diverts part of the hot flue gas from within the economizer through a bypass duct and mixes it with the relatively cooler flue gas exiting the economizer. An economizer feedwater bypass also raises the flue gas temperature. The use of an economizer bypass results in less energy transfer to the feedwater for steam generation, consequently, there is a small reduction in boiler efficiency. Lower boiler efficiencies require more fuel to be burned to meet the required boiler steam output.

Residence Time and Space Velocity

Residence time is the time the reactants are within the reactor. Higher residence times generally result in higher NO_x removal rates. Temperature also affects the required residence time. The required residence time decreases as the temperature approaches the optimum temperature for the reduction reaction. Residence time is often expressed as space velocity, the inverse of residence time. The space velocity of a reactor is experimentally determined from the measured flue gas flow rate divided by the superficial volume of the catalytic reactor. The NO_x removal efficiency increases with decreasing space velocity, i.e. increasing catalyst volume, for a given flue gas flow rate.

The optimal residence time for an SCR system is a function of the number of active catalyst sites available for the reduction reaction and the gas flow rates within those active sites (interstitial flow rate). The "area velocity" is a parameter used by SCR vendors which relates the number of sites and the interstitial flow rate to residence time. The area velocity is defined as the space velocity divided by the catalyst pore surface area (specific surface area). For coal-fired boilers, typical specific surface areas range from 90 to 3,800 square feet per cubic feet (ft^2/ft^3) (300 to 1,200 square meters per cubic meters (m^2/m^3)) [1]. Increasing the catalyst specific surface area increases the NO_x removal for a given flue gas flow rate. This can be accomplished by either increasing the catalyst volume, which increases the reactor size, or increasing the pore space of the catalyst, which generally increases the catalyst cost.

Degree of Mixing

The reagent must be dispersed and mixed throughout the flue gas to ensure sufficient contact between the reactants. Mixing is performed by an injection system which injects pressurized gas-phase ammonia into the flue gas. The injection system controls the spray angle, velocity and direction of the injected reagent. Some systems inject the ammonia with a carrier fluid such as steam or air to increase penetration into the flue gas. Injection systems are application specific. Numeric modeling of the flue gas and reagent flow optimizes the design of the injection system. (See 2.2.6).

Mixing of the flue gas and ammonia occurs before entering the SCR reactor. If mixing is not adequate, the NO_x reduction is inefficient. SCR designs must incorporate adequate duct length between the ammonia injection and the reactor inlet to allow for mixing. Mixing patterns can be improved by:

- Installation of static mixers upstream of the reactor;
- Increase the energy imparted to the injected fluids;
- Increase the number of injectors and/or injection zones; and
- Modify the nozzle design to improve the reagent distribution, spray angle, and direction.

Actual Stoichiometric Ratio

The actual stoichiometric ratio (the moles of reagent injected per mole of uncontrolled NO_x) defines the quantity of reagent needed to achieve the targeted NO_x reduction. According to Equation 2.1, the theoretical stoichiometric ratio for the reduction reaction with ammonia is equal to 1. This assumption of a 1 to 1 linear relationship between the quantity of reagent and the NO_x removed is good up to about 85% NO_x reduction [11]. After 85%, the removal efficiency begins to level off and more than the theoretical amount of ammonia is required for additional NO_x removal. This is due to the portion of NO_x that is in the form of NO_2 rather than NO and reaction rate limitations. SCR systems typically employ a stoichiometric ratio 1.05 moles of ammonia per mole of NO_x [1]. Because capital and operating costs depend on the quantity of reagent consumed, the actual stoichiometric ratio is an important design parameter that is determined by the SCR designer.

Uncontrolled NOx Concentration

The concentration of the reactants also affects the reaction rate of the NO_x reduction process. In general, higher uncontrolled NO_x inlet concentrations result in higher NO_x removal efficiencies [1] due to reaction kinetics. However, NO_x levels higher than approximately 150 parts per million (ppm), generally do not result in increased performance. Low NO_x inlet levels result in decreased NO_x removal efficiencies because the reaction rates are slower, particularly in the last layer of catalyst [1]. In general, though, SCR performs better than SNCR on sources with low uncontrolled NO_x levels such as natural gas-fired boilers.

For a given NO_x removal efficiency, higher NO_x levels at the SCR inlet require more catalyst volume. For example, to achieve 90% NO_x removal, requires10% more catalyst at inlet NO_x level of 1.7 lb/MMBtu versus inlet levels of 0.8 lb/MMBtu [9]. SCR is generally more cost effective for sources which emit less NO_x, since the required catalyst volume is minimal.

Ammonia Slip

Ammonia slip refers to the excess reagent passing through the reactor. Ammonia in the flue gas causes a number of problems which were discussed in Chapter 1 SNCR, including health effects, visibility of the stack effluent, salability of the fly ash, and the formation of ammonium sulfates. Limits on acceptable ammonia slip, imposed by either regulatory limits or by design requirements, place constraints on SCR performance.

Ammonia slip does not remain constant as the SCR system operates but increases as the catalyst activity decreases. Properly designed SCR systems, which operate close to the theoretical stoichiometry and supply adequate catalyst volume, maintain low ammonia slip

levels, approximately 2 to 5 ppm. Reliable instrumentation for monitoring ammonia slip is currently being developed but not commercially available [11]. One method to quantify ammonia slip is to determine the ammonia concentration in collected fly ash [13].

Catalyst Activity

Catalyst activity is a measure of how much the catalyst accelerates the NO_x reduction reaction rate. A higher catalyst activity results in a faster reaction rate and more NO_x removal. Catalyst activity is a function of many variables including catalyst composition and structure, diffusion rates, mass transfer rates, gas temperature, and gas composition [14]. As the catalyst activity decreases, the NO_x reduction reaction rate also decreases. This results in lower NO_x removal and higher ammonia slip levels.

The following equation describes the deactivation of the catalyst activity, K, with time, t [8]:

$$K = K_o \, e^{\left(\frac{t}{\tau}\right)} \tag{2.2}$$

where K_o is the original catalyst activity and τ is the catalyst operating life time constant. Figure 2.4 shows a typical catalyst deactivation curve based on Equation 2.2. As the catalyst activity decreases, the NO_x removal efficiency is usually kept constant by injecting more ammonia, thereby increasing the ammonia slip. When the ammonia slip reaches the maximum design or permitted level, new catalyst must be installed.

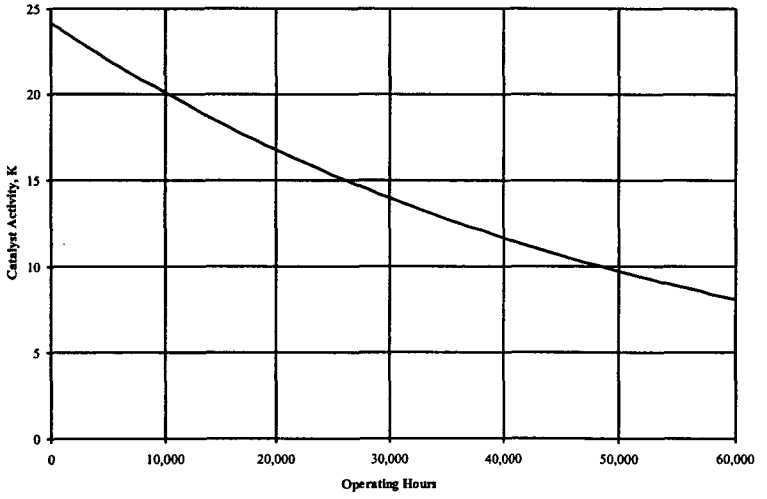


Figure 2.4: Typical Catalyst Deactivation per Equation 2.2 with $K_o = 24.12$; $\tau = 50{,}000$

Catalyst Reaction Selectivity

SCR favors the NO_x reduction reaction over competing reactions given the reactants are at the appropriate temperature and oxygen is present. However, competing reactions still occur and the catalyst accelerates these reactions as well. Each catalyst has different chemical reaction selectivity properties. In general, catalysts promote the formation of two undesirable compounds, sulfur trioxide (SO_3) and nitrous oxide (N_2O). SO_3 is formed by the oxidation of SO_2 to SO_3. Sulfur oxides (SO_x) are regulated under the 1990 Clean Air Act. SO_3 reacts with ammonia in the flue gas to form ammonia sulfates. Ammonium sulfur salts deposit on the catalyst and on downstream equipment such as the air preheaters. N_2O is both an ozone depletor and a greenhouse gas but is currently not regulated.

Pressure Loss

The flue gas pressure decreases as the flue gas flows across the catalyst. The decrease in pressure is a function of the length of the catalyst and the catalyst configuration. Deposition of fly ash and other particulates on the catalyst over time increases this pressure drop across the catalyst. The flue gas pressure can be increased by installing new draft fans or by upgrading existing fans. To minimize the pressure loss across the catalyst, the SCR reactor ductwork can be expanded and flow rectifiers and turning vanes can be installed. Pressure loss is of greater concern in turbine applications which rely on air flow, rather than heat transfer, to generate power.

Catalyst Pitch

Catalyst pitch is a term used in association with honeycomb and metal plate catalyst and affects the flue gas velocity in interstitial spaces [1]. As shown in Figure 2.5 pitch, represented as *p*, is the width of the catalyst cell plus the cell wall thickness, *a*. For a given

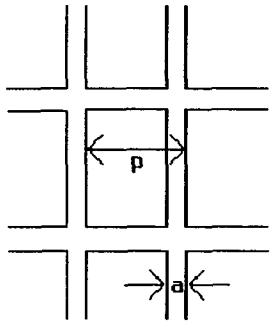


Figure 2.5: Pitch for a Honeycomb Catalyst Configuration

flow rate, wider pitch will result in lower interstitial gas velocities. Appropriate catalyst pitch is important to assure that ash will not deposit and bridge over catalyst cells and pores. Plugging of the catalyst reduces the effective surface area by decreasing the number of active sites available for the NO_x reduction reaction.

Catalyst Deactivation

Catalysts lose their activity over time for various reasons. The primary mechanisms for catalyst deactivation and surface area loss are discussed below.

Poisoning - Certain fuel constituents which are released during combustion act as catalyst poisons. Catalyst poisons include calcium oxide and magnesium oxide, potassium, sodium, arsenic, chlorine, fluorine, and lead. These constituents deactivate the catalyst by diffusing into active pore sites and occupying them irreversibly. Catalyst poisoning represents the main cause of catalyst deactivation.

Thermal Sintering - High flue gas temperatures within the SCR reactor cause sintering, a permanent loss of catalyst activity due to a change in the pore structure of the catalyst. Thermal sintering can occur at temperatures as low as 450°F (232°C). The amount of thermal sintering depends on the composition and structure of the catalyst. Newer catalyst materials are less susceptible to thermal sintering which increases their operating life.

Blinding/Plugging/Fouling - Ammonia-sulfur salts, fly ash, and other particulate matter in the flue gas cause blinding, plugging or fouling of the catalyst. The particulate matter deposits on the surface and in the active pore sites of the catalyst. This results in a decrease of the number of sites available for NO_x reduction and an increase in flue gas pressure loss across the catalyst.
Erosion - Impingement of particulate matter and high interstitial gas velocities erode

the catalyst material. Catalysts with hardened leading edges or increased structural strength are less susceptible to erosion. Increasing catalyst strength through hardening, however, reduces the number of active pore sites.

Aging - Catalyst aging is a change in the physical and chemical properties of the catalyst pores that occurs over time.

There are a number of measures which can be taken to decrease the rate of deactivation and deterioration of the catalyst. These measures are discussed in the following section.

Catalyst Formulation - Each catalyst formulation has different physical and chemical properties. Catalyst formulations with the following properties will have decreased deactivation.

- Increased activity per unit volume
- Greater thermal resistance
- Chemical and physical resistance to poisons
- Wider thermal operating range
- Greater structural strength and hardened leading edges
- Lower interstitial velocities (i.e., wider catalyst pitch)

To obtain the optimum catalyst formulation and SCR design for an application, the catalyst supplier and SCR vendor should be informed of the fuel constituents, such as sulfur, chlorine, fluorine, alkali, and trace metals. These fuel and ash constituents can be determined by chemical analyses. The associated analytical data can then be used to modify the catalyst composition, to determine catalyst volume, and to design the SCR reactor components.

Soot Blowers - Deposits on the surface of the catalyst can be dislodged by soot blowers which are generally installed between each catalyst layer and operated on a periodic basis, such as once a week.

Turning Vanes and Rectifier Grids - Particulate matter can be removed from the flue gas by gas-flow turning vanes and flow rectifier grids near the front of the catalyst layer. Particles impact the surface of the vanes or grid and fall out of the flue gas stream. In addition to removing particles, turning vanes and flow rectifier grids decrease the linear velocity of the flue gas and align its vector with the flow path of the catalyst.

Catalyst Management Plan

Catalyst deactivation is an inherent part of the SCR process. As the catalyst activity decreases with time, the NO_x reduction reaction rate decreases and ammonia slip increases. When the ammonia slip level reaches the design limit, the catalyst must be replaced or new catalyst must be added. The catalyst life is the time the catalyst activity for a given catalyst volume (layer volume) maintains ammonia slip below the design limit. Currently, vendor-guaranteed life for a catalyst layer in coal-fired applications is approximately three years [10]. The actual catalyst layer lifetimes being experienced in such applications are in the 5

to 7 year range, depending on the condition of untreated flue gas [1]. Gas- and oil-fired applications experience even longer catalyst layer lifetimes.

A catalyst management plan, as shown in Figure 2.6, schedules periodic replacement of catalyst to maintain ammonia slip limits. Most catalyst management plans call for the SCR reactor design to provide two or more layers filled with catalyst and one or more empty or spare catalyst layers. When the initial catalyst layers deactivate to the point ammonia slip reaches the maximum design value, the facility adds catalyst to the empty layer. Catalyst addition is managed so that the total catalyst activity of all the layers (the two or three older catalyst layers plus new catalyst layers) is sufficient to meet the ammonia slip requirement for a relatively long period of time. As the catalyst continues to deactivate, ammonia slip begins to rise. When ammonia slip again reaches the maximum design value, one of the older catalyst layers is removed and replaced with new catalyst. Figure 2.6 shows a catalyst management plan for an application in which the maximum design value of ammonia slip is 2 ppm [7].

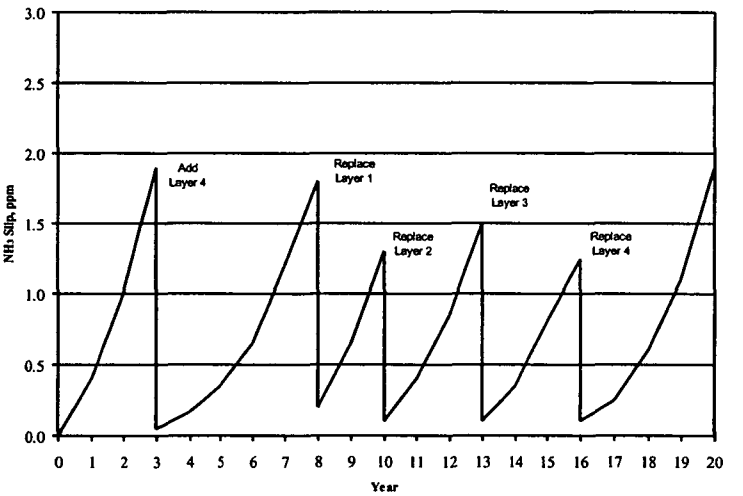


Figure 2.6: Typical Catalyst Management Plan, [5]

Catalyst cost is a significant portion of the annual cost of operating an SCR system. For SCR designs which utilize a catalyst management plan, only a fraction of the total catalyst inventory, rather than the entire volume, is replaced at any one time. This distributes the catalyst replacement costs more evenly over the lifetime of the system.

2.2.3 SCR System Configurations

Electric utility and large industrial boiler applications implement several different SCR system configurations, including high-dust, low-dust, and tail-end arrangements. SCR configurations for gas turbine applications are dependant on the type of engine cycle, such as combined-cycle or simple cycle. The various configurations for boilers and gas-fired turbines are discussed below. In addition, there are two different SCR reactor designs; full SCR and in-duct SCR which are also discussed.

High-Dust SCR

Figure 2.7 shows a high-dust SCR system for coal-fired boiler applications. The SCR reactor location is downstream of the economizer and upstream of the air heater and particulate control devices. The flue gas temperature in this location is usually within the optimum temperature window for NO_x reduction reactions using metal oxide catalysts. In this configuration, however, the flue gas contains particulates when it enters the SCR reactor.

Coal-fired boilers generally use a vertical SCR reactor, where the flue gas flows downward through the catalyst. The reactor generally contains multiple layers of catalyst. The volume of catalyst required varies with each installation, as discussed previously. Sootblowers are installed to remove particulates from the catalyst surfaces. For design which utilize a honeycomb catalyst, the catalyst pitch is typically about 7 to 9 mm (compared with 3 or 4 mm for gas-fired boilers) to allow easy passage of ash particles without deposition

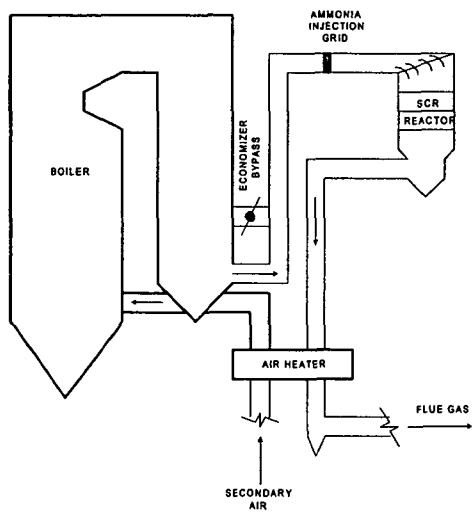


Figure 2.7: High Dust SCR Arrangement, [4]

and for ease of cleaning with sootblowers. To obtain uniform gas flow and remove particulates, high-dust SCR designs usually include turning vanes and a flow-rectifying grid in the ductwork prior to the reactor.

A hopper at the bottom of the SCR reactor collects ash and particulates separated from the flue gas stream. The hopper outlet connects to the plant fly ash handling system for periodically removal of the accumulated ash. Flue gas exits the reactor via an opening at the top of the hopper and is directed to the air heater inlet. Some designs eliminate the need for hoppers by keeping flue gas velocities high enough in these areas so fly ash remains entrained in the flue gas.

Natural gas-fired and distillate oil-fired boilers generate flue gas that is relatively free of dust and SO_2 (for low-sulfur oil). Consequently, SCR systems for these boilers place the reactor upstream of the air heater, the high-dust SCR configuration.

Low-Dust SCR

Coal-fired units with an electrostatic precipitator (ESP) located upstream of the air heater (hot-side ESP), typically use a low-dust SCR configuration. Figure 2.8 shows a low dust configuration, which locates the SCR reactor downstream of the ESP. In this location, the flue gas is relatively dust free. The ash removed by the ESP typically contains arsenic, alkali metals, and other constituents that are detrimental to catalyst performance and life.

A low-dust SCR system increases catalyst life by reducing concentrations of particulates and catalyst poisons in the SCR reactor. In addition, low-dust SCR configurations

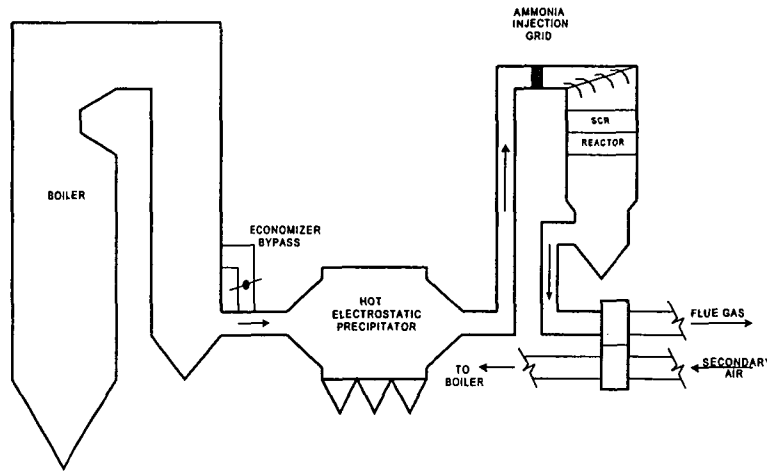


Figure 2.8: Low-Dust SCR Arrangement, [4]

do not need ash hoppers. For designs employing honeycomb catalyst, the catalyst pitch can be reduced to approximately 4 to 7 mm, resulting in lower catalyst volume. Longer catalyst life, lower catalyst volume and the elimination of the ash hopper mean lower costs for low-dust SCR compared to high-dust configurations. The only disadvantage of low dust SCR is the temperature drop of the flue gas as it flows through the ESP. Flue gas temperatures generally do not decrease to the point where reheating is required. However, an increase in the size of the existing economizer bypass duct may be required to maintain the flue gas temperature within the optimum range.

Tail-end SCR

Early installations of coal-fired boilers in Europe and Japan employ tail-end SCR configurations. This configurations places the SCR reactor downstream of all air pollution control equipment installed on a unit Figure 2.9 depicts a tail-end system for a plant with a particulate control device and a wet flue gas desulfurization (FGD) system. The air pollution control equipment removes most flue gas constituents detrimental to SCR catalyst before it enters the SCR reactor. However, because the flue gas temperature at the tail-end is below the range required for the ammonia/NO_x reaction, the flue gas needs to be reheated. Tail end SCR systems use oil- or natural gas-fired duct burners or steam coil gas heaters for reheating. Some of the energy used to reheat the gas is recovered in a recuperating gas-to-gas heater.

A tail-end system typically costs the most of the three SCR systems because of the additional equipment and operational costs required for flue gas reheating and heat recovery. Operating experience and the availability of improved catalysts for the high-dust SCR system makes the tail-end SCR system the least attractive of the three options. New low temperatures catalysts are currently being developed and may make tail end systems a more cost effective option in the future [25].

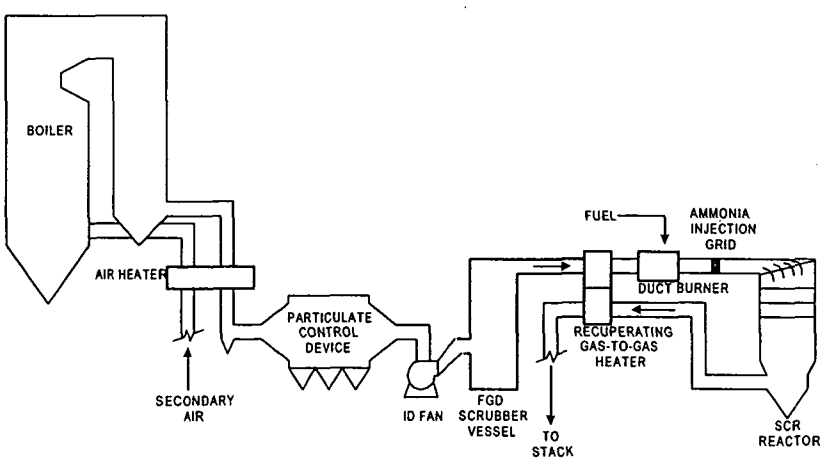


Figure 2.9: Tail-end SCR Arrangement, [4]

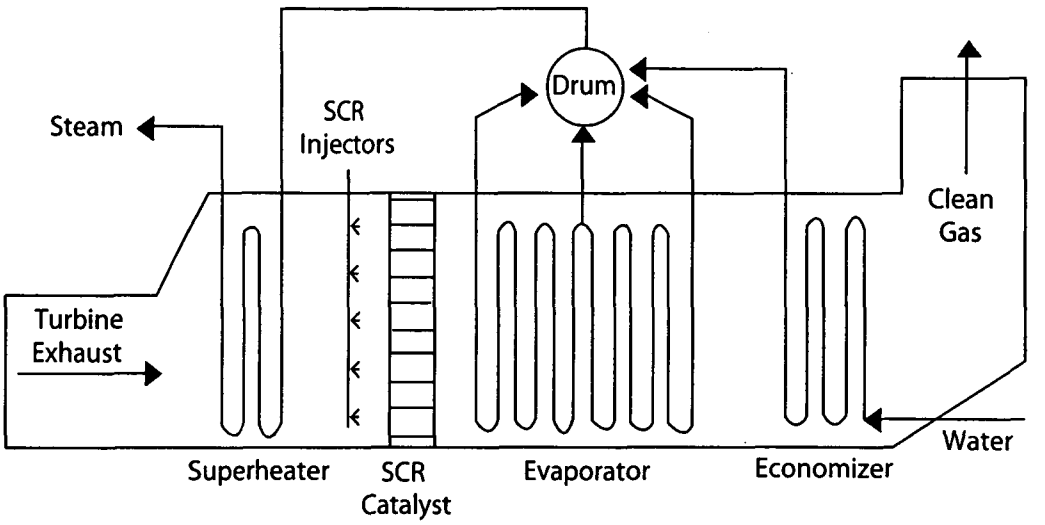


Figure 2.10: SCR Arrangement for a Combined Cycle Gas Turbine, [4]

Gas Turbines

Natural gas-fired turbine applications frequently use SCR technology for post-combustion NO_x control. There are two basic gas turbine configurations; combined cycle (co-generation cycle) and simple cycle. The majority of SCR systems are installed as combined cycle applications. A typical combined-cycle SCR design places the reactor chamber after the superheater within a cavity of the heat recovery steam generator system (HRSG) as shown in Figure 2.10. The flue gas temperature in this area is within the operating range for base metal-type catalysts . Some combined-cycle gas turbine designs locate the SCR reactor downstream of the HRSG and prior to the economizer at temperatures ranging from 350°F to 400°F. Simple cycle applications of SCR place the reactor chamber directly at the turbine exhaust where the flue gas temperature is in the range of 850°F to 1000°F (450°C to 540°C). This requires the use of a high temperature catalyst such as zeolite. [4]

SCR Reactor Designs

The reactor design affects capital and operating costs of the SCR system and the catalyst management plan. There are two different types of SCR reactors; full SCR and in-duct SCR. Full SCR designs house the catalyst in a separate reactor chamber. The boiler flue gas must be ducted from the economizer outlet, to the SCR reactor, then to the air heater inlet. A separate reactor allows a large volume of catalyst to be installed in layers, which increases NO_x reduction and catalyst lifetime. It also increases the duct length avai able for the mixing of reactants prior to entering the reactor chamber. A separate reactor,

however, requires a large amount of space adjacent the boiler to install the reactor and ductwork. The additional ductwork often necessitates upgrades to the draft fan system.

In-duct (in-line) SCR systems house the reactor within the plant's existing ductwork rather than in a separate reactor chamber. The ductwork is generally enlarged to provide sufficient room for the catalyst. In-duct systems save on costs for the ductwork, reactor chamber, and ID fan. In-duct designs limit catalyst volume and mixing length, therefore, they are commonly used in conjunction with other NO_x control technologies [19]. Catalyst erosion is generally higher for in-duct systems. Installation and maintenance of in-duct systems typically require more boiler outages. Natural gas-fired boilers, which have low catalyst volumes, frequently employ in-duct systems. Coal-fired boilers may apply in-duct SCR reactors where space limitations restrict the installation of a full reactor [19].

2.2.4 High Dust SCR System Equipment

The majority of SCR designs utilize Thermal DeNO$_x$®, which is a an ammonia-based NO_x reduction system developed and patented by Exxon Research and Engineering Company in 1975. The SCR system has five basic steps to accomplish. These steps are:

- Receiving and storage of the ammonia;
- Vaporizing the ammonia and mixing with air;
- Injecting the ammonia/air mixture at appropriate locations;
- Mixing the ammonia/air with flue gas; and
- Diffusing the reactants into the catalyst and reducing the NO_x.

While the basic steps in an SCR system are similar for all configurations, the system design and equipment specifications are somewhat different. A discussion of the SCR system design and equipment is given below for an aqueous ammonia-based, high dust configuration, full reactor SCR for a large (≥2,500 MMBtu/hr) coal-burning industrial boiler. The SCR process steps, related auxiliary equipment, and the potential impacts of SCR operation on existing plant equipment are also discussed. A simplified system flow schematic is presented in Figure 2.1 and Figure 2.7 and a list of equipment is presented in Table 2.2.

Ammonia Storage and Vaporization

Aqueous ammonia is typically available as a 19% to 29.4% solution in water. Generally, a 29.4 percent solution is used in this application. Anhydrous ammonia is nearly 100% pure ammonia and stored as a liquid under pressure. Table 2.1 presents the properties of aqueous and anhydrous ammonia.

Table 2.2: Major Equipment List for an SCR Application

Item	Description/Size
SCR reactors (1 to 2)	Vertical flow type, 805,000 acfm capacity, 44 ft.× 44 ft.× 31 ft. high (excluding outlet duct and hoppers), equipped with 9,604 ft^3 of ceramic honeycomb catalyst, insulated casing, sootblowers, hoppers, and hoisting mechanism for catalyst replacement
Anhydrous ammonia tank (1 or more)	Horizontal tank, 250 psig design pressure, storage tanks 15,000 gal, 34-ton storage capacity
Air compressor (2)	Centrifugal type, rated at 3,200 acfm and 30 hp motor
Vaporizers (2)	Electrical type, rated at 80 kW
Mixing chamber	Carbon steel vessel for mixing or air and ammonia
Ammonia injection grid	Stainless steel construction, piping, valves and nozzles
Ammonia supply piping diameter, with valves and fittings	Piping for ammonia unloading and supply, carbon steel pipe: 1.0"
Sootblowing steam	Steam supply piping for the reactor soot-piping blowers, 2" diameter pipe with an on-off control valve and drain and vent valved connections
Air ductwork	Ductwork between air blowers, mixing chamber, and ammonia injection grid, carbon steel, 14 in. dia., with two isolation butterfly dampers and expansion joints
Flue gas ductwork	Ductwork modifications to install the SCR modifications reactors, consisting of insulated duct, static mixers, turning vanes, and expansion joints
Economizer bypass	Ductwork addition to increase flue gas temperature during low loads consisting of insulated duct, flow control dampers, static mixers, turning vanes, expansion joints, and an opening in the boiler casing
Ash handling	Extension of the existing fly ash handling modifications system: modifications consisting of twelve slide gate valves, twelve material handling valves, one segregating valve, and ash conveyor piping
Induced draft fans	Centrifugal type, 650,000 acfm at 34" wg and 4,000 hp motor
Controls and instrumentation	Stand-alone, microprocessor-based controls for the SCR system with feedback from the plant controls for the unit load, NO_x emissions, etc., including NO_x analyzers, air and ammonia flow monitoring devices, ammonia sensing and alarming devices at the tank area, and other miscellaneous instrumentation
Electrical supply	Electrical wiring, raceway, and conduit to connect the new equipment and controls to the existing plant supply systems
Electrical equipment	System service transformer OA/FA/-60 Hz, 1,000/1,250 kVA (65 °C)
Foundations	Foundations for the equipment and ductwork/piping, as required
Structural steel	Steel for access to and support of the SCR reactors and other equipment, ductwork, and piping

Facilities receive anhydrous or aqueous ammonia via a tank-truck or rail car and pump it into one or more storage tanks. Ammonia is typically stored as a liquid in horizontal cylindrical tanks. The aqueous ammonia tank is an enclosed tank rated for only slightly elevated pressure, while the anhydrous ammonia tank is a pressure vessel rated for at least 250 pounds per square inch gauge (psig). The anhydrous ammonia tank can be filled only to about 85% of its total volume to allow for a vapor space above the liquid level. The tanks are equipped with level and temperature indicators, a manway, vent, and access ladder, and other appurtenances. The applicability of heat tracing, insulation, and seismic design criteria are determined based on site-specific conditions. The tank should be mounted on a concrete pad and surrounded by a spill containment structure such as a dike.

SCR applications on large boilers generally require one to five tanks with volumes ranging from 10,000 to 20,000 gallons per tank to maintain sufficient volume for 1 to 3 weeks of SCR operations. The ammonia storage tank may be sized for 3 to 30 days of storage.

The high end of the range would be used in conservative design practice. Alternatively, if ammonia distributors are located nearby and considered reliable, the plant owner might opt for a smaller tank, sized for fewer days of ammonia storage.

Aqueous ammonia is vaporized by pumping it to a vessel where it mixes with hot air. The air from the dilution air fan is heated in an electric heater or other heat exchanger (e.g., steam). In most aqueous ammonia applications, the ammonia-air mixture leaves the vaporizer vessel at about 300 °F. The vaporization energy required for aqueous ammonia is much greater than that required for anhydrous ammonia because the water in the aqueous ammonia solution also must be vaporized.

If anhydrous ammonia is used, it is fed to the electrical vaporizer by gravity, and the vaporized gas is returned to the storage tank vapor space. Vapor is drawn from the vapor space and piped to the ammonia/air mixer. Alternatively, liquid anhydrous ammonia may be pumped to a vaporizer and piped to the ammonia/air mixer.

Ammonia (aqueous or anhydrous), diluted with air at a ratio of about 20:1 (air:NH_3), is transported to the ammonia injection grid. The high proportion of air helps ensure good mixing of air and ammonia and keeps the mixture below the flammable limit.

Ammonia Injection

For either aqueous or anhydrous ammonia, the ammonia-air mixture is directed through a flow-balancing skid to the ammonia injection grid (AIG) where it is injected under pressure. The flow-balancing skid consists of flow meters and manual valves to adjust the flow to each part of the AIG.

The AIG consists of a network of pipes or lances connected in parallel, and perforated with several holes or nozzles. The lances are placed in a grid formation across the width and height of the ductwork. The lances and holes are sized to distribute the ammonia uniformly into the flue gas. The spray angle and velocity of the injection control the trajectory of the ammonia. Injectors are subject to high-temperatures and to flue gas impingement which cause erosion, corrosion, and structural integrity degradation. Therefore, injectors are generally constructed of stainless steel and designed to be replaceable. Multiple injection zones may be utilized to increase the distribution of ammonia.

The ammonia can be injected with a low energy or high energy system. A low-energy system uses little or no pressurized air while a high-energy system uses large amounts of compressed air or steam to inject and vigorously mix the solution with the flue gas. AIG systems in large boilers typically use high-energy systems. High energy systems are more expensive to build and operate since they require a larger compressor, a more robust injection system, and consume more electric power.

Uniform distribution and mixing with flue gas is critical to maintain desired low levels of ammonia slip. Cold gas flow modeling and numerical flow modeling are generally performed for the AIG and SCR system to ensure uniform mixing and dispersion before the gases enter the SCR reactor. If duct length is inadequate to ensure thorough mixing, or results from the model study indicate poor gas mixing characteristics, devices such as turning vanes or static gas mixers may be added.

An essential part of an AIG system is the controller used to regulate ammonia injection. Boiler load, inlet NO_x, and inlet gas temperatures set the feed-forward signal to establish the base ammonia injection rate. A feedback signal measuring the SCR outlet NO_x concentration is used to trim the base ammonia injection rate.

Catalytic Reduction of NOx

The catalytic reduction of NO_x in the SCR reactor occurs when the NO_x and ammonia in the flue gas contact the catalyst layers. The catalyst itself is the key component of the SCR system. The catalyst composition, type (honeycomb or plate), and physical properties affect performance, reliability, catalyst quantity required, and cost. However, because the SCR system supplier and catalyst supplier must guarantee catalyst life and performance, most catalyst characteristics are selected by the SCR system supplier.

2.2.5 SCR System Auxiliary Equipment

SCR Inlet and Outlet Ductwork

In retrofit installations, new ductwork is required to integrate the SCR system with the existing equipment. In high-dust SCR systems, the reactor is located between the econo-

mizer outlet, the air heater inlet. In the low-dust SCR, the SCR reactor is located between the outlet duct of the particulate control device and the the air heater inlet duct. In the tail-end SCR, the ductwork tie ins are downstream of the FGD system and also require the integration of the flue gas reheating equipment.

SCR Bypass Duct

Low load boiler operations can decrease the temperature at the SCR reactor inlet below the SCR operating range. In addition, startup and shutdown of the boiler causes drastic temperature fluctuations. For these operating conditions, an SCR bypass may be required to route the flue gas around the reactor chamber. The bypass prevents catalyst poisoning and fouling during periods when the SCR is not operating. This system must include zero-leakage dampers to protect the catalyst from flue gas leakage and deactivation during SCR shut down. A bypass system may also be considered for seasonal operation of the SCR system.

Sootblower

In coal-fired boilers, sootblowers are usually installed in the SCR reactor to remove particulates that may mask or block active catalyst surfaces and gas passages. Sootblowing helps maintain acceptable flue gas pressure drop in the SCR reactor by keeping the catalyst gas passages free of particulate. Sootblowers also keep the air heater gas passages open and thereby reduce system pressure drop. This is especially true for SCR retrofits where the air heater plate spacing is generally narrow, making it more susceptible to fouling or clogging by ammonia-sulfur salts.

Retractable rake-type sootblowers which use steam or air for blowing are utilized in SCR designs. The sootblowers are typically located above each catalyst layer. Sootblowing is usually performed on one catalyst layer or part of one catalyst layer at a time. Sootblowing of all the catalyst layers takes between 30 minutes to 2 hours, but it is usually done infrequently. In European SCR installations, sootblowing is done approximately once or twice a week [18].

Economizer Bypass Duct

Although the SCR reaction occurs within a temperature window of 600°F to 750°F (320°C to 400°C), the catalyst for a given application is designed for a somewhat narrower range, the economizer outlet temperature at normal boiler operating load. Maintaining the flue gas temperature within the required window is essential for optimizing the NO_x reduction reaction. When the economizer outlet flue gas temperature decreases because the plant is operating at reduced loads, the temperature can be raised using an economizer bypass.

The economizer bypass duct generally has a modulating damper to regulate the amount of hot bypass gas flow to be mixed with the cooler economizer outlet flue gas. The lower the boiler load, the more this damper opens, thus admitting more hot gas. The economizer outlet duct also needs a modulating damper to provide enough backpressure to allow the required volume of gas to flow through the bypass. The main design considerations for an economizer bypass involve maintaining the optimum gas temperature and ensuring uniform mixing of the two gas streams prior to entering the SCR reactor.

Upgraded or New Induced Draft (ID) Fan

The new ductwork and the SCR reactor's catalyst layers decrease the flue gas pressure. In order to maintain the same flow rate through the duct work, additional energy is required. The existing induced draft (ID) fan may be unable to provide the required increase in static pressure. In such cases, an upgraded or new ID fan is installed. The existing fan and motor foundation also may need modification. Replacement involves installation of a new fan or booster fan. In all cases, SCR systems require additional electric power for the ID fan equivalent to approximately 0.3 percent of the plant's electric output. Refer to Equation (2.49) for a more precise estimate of the additional electric power needs.

2.2.6 Other Considerations

Formation of SOx

Sulfur trioxide (SO_3) forms during the combustion of fuels which contain sulfur. It reacts with ammonia in the flue gas downstream of the reactor (ammonia slip) to form ammonium bisulfate and ammonium sulfate. The amount formed depends on the sulfur content of the fuel and the amount of ammonia slip. Ammonium bisulfate condenses as the flue gas stream temperatures lowers. It then deposits on the SCR catalyst and downstream equipment such as the air heater, ducts, and fans. Ammonia slip limits are generally imposed as part of the SCR design requirements to avoid impacts on downstream equipment.

There are several methods for limiting the impact of ammonia-sulfur salt deposition. Soot blowers can be installed between catalyst layers to remove surface deposits by blowing air or steam across the catalyst. Increased acid washing of the air preheater and other equipment may be required to remove deposits. However, more frequent acid washing generates additional wastewater which must be disposed or treated by the plant. The sulfur content of the flue gas can be decreased by coal desulfurization processes or fuel switching. Lastly, the flue gas temperature may be raised to prevent condensation.

Elevated SO_3 concentrations raise the acid dew point of the flue gas. This phenomenon potentially leads to more corrosion on the air heater's cold-end surfaces if the flue gas temperature is below the acid dew point. To protect against this possibility, the cold-end baskets of the air heater can be replaced with enamel-coated baskets.

Ammonia sulfates also deposit on the fly ash. Ammonia content in the fly ash greater than 5 ppm can result in off-gassing which would impact the salabilty of the ash as a byproduct and the storage and disposal of the ash by landfill. [10] (See Chapter 1 SNCR)

Formation of Arsenic Oxide

Arsenic oxides (As_2O_3), formed during combustion of fuel containing arsenic, cause catalyst deactivation by occupying active pore sites. Coal burning boilers are particularly susceptible to arsenic poisoning. Limestone ($CaCO_3$) can be injected into the flue gas to generate the solid $Ca_3(AsO_4)_2$, which does not deposit on the catalyst and can be removed from the flue gas with a precipitator.

Retrofit Versus New Design

Retrofit of SCR on an existing boiler has higher capital costs than SCR installed on a new boiler system. The magnitude of the cost differential is a function of the difficulty of the retrofit. A large part of the capital costs are not impacted by retrofit including ammonia storage, vaporization, and injection equipment costs. The increase in cost is primarily due to modifications to existing ductwork, the cost of structural steel and reactor construction, auxiliary equipment costs, such as additional fans, and engineering costs. In addition, significant demolition and relocation of equipment may be required to provide space for the reactor. These costs can account for over 30% of the capital costs associated with SCR [9]. Retrofit costs for cyclone or wet bottom wall-fired boilers are somewhat higher than retrofit costs for dry bottom wall- or tangentially-fired boilers [4]. Differential retrofit cost for SCR in Germany is approximately 200 $ per MMBtu/hr (20 $/kW) [4].

Combustion Unit Design and Configuration

Boiler size is one of the primary factors that determines the SCR system capital costs. In addition, boiler configuration influences SCR costs. Boiler configurations that split the flue gas flow for two or more air preheaters and/or particulate removal systems require more than one SCR reactor. Additional reactors substantially increase capital costs. Boiler operations that have varying operating load, frequent startup/shutdowns, or seasonal operations require an SCR bypass. Additional ductwork, dampers, and control systems increase the SCR system capital costs. The SCR system may require modifications to draft fans and/or installation of additional fans. This increases both capital and operating costs of the SCR system. In addition, boiler and duct modifications may be required for implosion protection to accommodate increased draft requirements. [9]

Fuel Source

Industrial boilers use coal, distillate oil, residual oil, and natural gas. The fuel type and grade affects the SCR design and, therefore, the capital costs of the SCR system. Fuels

with high heating value have higher gas flow rates, which in turn results in an increase of the required SCR reactor size and catalyst volume. Coal-fueled applications are more costly than oil- and natural gas-fired boilers due to their higher flue gas flow rates. [9]

The quantity of nitrogen, fly ash and pollutants in the flue gas stream varies according to the type and grade of the fuel. This affects the volume of catalyst required as well as the catalyst design, composition and rate of deactivation. Coal flue gas contains a greater amount of fly ash, sulfur dioxide, sulfur trioxide, arsenic and other trace pollutants than oil and natural gas. Natural gas is the cleanest fuel and contains the least amount of nitrogen, therefore, burning natural gas results in the least amount of NO_x and pollutants in the flue gas stream. The amount of nitrogen and pollutants in oil-based fuels sources varies with the grade and type of oil, either refined or residual.

Modeling of the SCR System

Computational fluid dynamics (CFD) and chemical kinetic modeling are performed as part of the design process for SCR (See Chapter 1 SNCR). In addition to CFD and chemical kinetic modeling, three dimensional, physical flow modeling, also referred to as cold flow modeling, is generally required. Cold flow modeling ensures that the flow through the SCR reactor provides adequate residence time, achieves uniform mixing of flue gas and ammonia, minimizes linear velocities to prevent catalyst erosion, and minimizes pressure drop across the catalyst layers. It involves constructing a model of the ammonia injection system, mixing area and reactor chamber. In a high-dust configuration, this involves modeling from the economizer outlet to the inlet of the air heater. Typical model scales range from a 1:10 ratio to a 1:12 ratio for large electric utility boilers.

2.3 Design Parameters

SCR system design is a proprietary technology. Extensive details of the theory and correlations that can be used to estimate design parameters such as the required catalyst volume are not published in the technical literature [4]. Furthermore, the design is highly site-specific. In light of these complexities, SCR system design is generally undertaken by providing all of the plant- and boiler-specific data to the SCR system supplier, who specifies the required catalyst volume and other design parameters based on prior experience and computational fluid dynamics and chemical kinetic modeling. [1]

This section presents a step-by-step approach to estimate design parameters based on a procedure developed in the draft EPA report Selective Catalytic Reduction for NO_x Control on Coal-fired Boilers [1]. This procedure assumes SCR system size and cost are based on three main parameters: the boiler size or heat input, the required level of NO_x reduction, and the catalyst volume. The approach to SCR sizing described in this section is based on the catalyst volumes for a base case and several sensitivity cases developed to

support the cost estimating procedures in Reference [1]. Although this approach is based on SCR data for utility boilers, it provides sufficient accuracy and detail to develop the capital and annual cost estimates for SCR as applied to industrial boilers.

Boiler Heat Input

The primary cost estimation parameter in the methodology presented in Reference [1] is the maximum potential heat released by the boiler or heat input rate, Q_B, expressed as million British thermal units per hour (MMBtu/hr). It is obtained from the high heating value, *HHV*, of the fuel in Btu per pound (Btu/lb) multiplied by the maximum fuel consumption rate in pounds per hour (lb/hr), $\dot{m}_{fuel}$:

$$Q_B = HV \ \dot{m}_{fuel} \tag{2.3}$$

where Table 2.3 gives the HHV for various coals. A conversion of 10^{-6} MMBtu/Btu is required for HHV values.

Table 2.3: High Heating Values for Various Coals

Type of Coal	Energy Content (Btu/ lb)
Lignite	5000-7500
Subbituminous	8000-10,000
Bituminous	11,000 - 15,000
Anthracite	14,000

If the boiler produces electricity, then its heat input can be estimated using the boiler net plant heat rate, in Btu per kilowatt-hr (Btu/kWh) *NPHR*:

$$Q_B = B_{MW} \ NPHR \tag{2.4}$$

where B_{MW} is the boiler megawatt (MW) rating at full load capacity. Note that if *NPHR* is not known (e.g., a cogeneration unit), a value of 9,500 Btu/kWh can be used as a reasonable

estimate. Using this value, the heat input rate, Q_B, is:

$$Q_B\left(\frac{MMBtu}{hr}\right) = 9.5\ \left(\frac{MMBtu}{MWh}\right) B_{MW}\,(MW) \tag{2.5}$$

System Capacity Factor

The total system capacity factor, CF_{total}, is a measure of the average annual use of the boiler in conjunction with the SCR system. CF_{total} is given by:

$$CF_{total} = CF_{plant}\ CF_{SNCR} \tag{2.6}$$

For industrial boilers, the capacity factor of the boiler, CF_{plant}, is the ratio of actual quantity of fuel burned annually to the potential maximum quantity of fuel burned annually in pounds. CF_{plant} is given by:

$$CF_{plant} = \frac{actual\ m_{fuel}}{maximum\ m_{fuel}} \tag{2.7}$$

SCR can be operated year-round or only during the specified ozone season. The capacity facor for the SCR system, CF_{SCR}, is the ratio of the actual number of SCR operating days, t_{SCR}, to the total number of days per year:

$$CF_{SCR} = \frac{t_{SCR}}{365\ days} \tag{2.8}$$

Uncontrolled NOx and Stack NOx

Uncontrolled NO_x, represented as NO_{xin}, is the NO_x concentration in the flue gas after any reductions provided by combustion controls, but prior to the SCR system. The uncontrolled NO_x concentration, obtained from analyzing the boiler flue gas stream, is generally given in pounds per MMBtu (lb/MMBtu) of NO_2 [1].

The stack NO_x, represented as NO_{xout} is the required NO_x emission limit at the stack outlet. It is generally set by the plant or regulatory limits and also given in pounds per MMBtu (lb/MMBtu) of NO_2 [1].

NOx Removal Efficiency

The NO_x removal efficiency, represented as η_{NOx}, is determined from the uncontrolled NO_x level of the boiler at maximum heat input rate, CF_{plant} = 1.0, and the required stack emission limit.

The equation for the NOx removal efficiency is given by:

$$\eta_{NO_x} = \frac{NO_{x_{in}} - NO_{x_{out}}}{NO_{x_{in}}} \qquad (2.9)$$

The required NO_x removal efficiency is one of the most influential parameters on the overall SCR system cost [9].

Actual Stoichiometric Ratios

The Actual Stoichiometric Ratio (*ASR*) indicates the actual amount of reagent needed to achieve the targeted NO_x reduction. Typical *ASR* values are higher than theoretical values due to the complexity of the reactions involving the catalyst and limited mixing. Higher ASR values generally result in increased NO_x reduction. The *ASR* is an important parameter in SCR system design because it establishes the reagent use of the SCR system. The *ASR* is defined as:

$$ASR = \frac{moles\ of\ equivalent\ NH_3\ injected}{mole\ of\ uncontrolled\ NO_x} \qquad (2.10)$$

For estimating purposes, the moles of NO_x are equivalent to the moles of NO_2. Note that the moles of equivalent NH_3 in Equation 2.10 the moles of NH_3 that will be released from the reagent. When using ammonia as the reagent, the equivalent moles of NH_3 injected is equal to the moles of ammonia injected.

In a design developed by a system supplier, the *ASR* would be adjusted to account for temperature, residence time, degree of mixing, catalyst activity, and allowable ammonia slip for a specific boiler. No equation for estimating *ASR* was available for SCR. The value for *ASR* in a typical SCR system is approximately:

$$ASR = 1.05 \qquad (2.11)$$

This value incorporates design margins for ammonia slip and the small amount of NO_2 in the boiler flue gas which requires two moles of NH_3 per mole of NO_2 instead of one mole of NH_3 per mole of NO as shown in Equation 2.1.

Flue Gas Flow Rate

The full-load flue gas flow rate, including the typical design margin of 5 to 15 percent, is used to size the SCR reactors and associated catalyst inventory. This flow rate should be obtained from test data or a combustion calculation.

If flow rate data are not available, an approximation of the flue gas flow rate to each of the SCR reactors can be calculated. The estimation of the volumetric flow rate in actual cubic ft. per minute (acfm) should be based on the expected operating gas temperature at the SCR inlet. The equation for the volumetric flow rate is:

$$q_{fluegas} = \frac{q_{fuel}\ Q_B\ (460 + T)}{(460 + 700°\text{F})\ n_{SCR}} \tag{2.12}$$

where n_{SCR} is the number of SCR reactors chambers and q_{fuel} is the fuel volumetric flow rate for the boiler in units of ft^3/min - MMBtu/hr and Q_b is given in MMBtu/lb.

The fuel flow rate q_{fuel}, is based on combustion calculations using typical fuels, typical boiler parameters (e.g., 20 % excess air), and typical SCR flue gas pressure (-10 in. w.g) [20]. Estimated values for q_{fuel} are given in Table 2.4 for various types of coal.

Table 2.4: Estimated Values of Volumetric Flow Rate for Various Coals

Coal Type	Estimate value of q_{fuel} (ft^3/min-MMBtu/hr)
Eastern Bituminous	484
Powder River Basin	516
Lignite	547

Note that, in general, the number of reactors, n_{SCR}, is site specific. One SCR reactor per boiler unit is typically required in high-dust system designs. However, two SCR reactors may be needed to treat flue gas from a boiler equipped with two air preheaters. The system designs developed for the base and sensitivity cases of this report use one reactor. Study-level costs of a two-reactor system are expected to be similar to the cost of a corresponding one-reactor system because the catalyst, ammonia, economizer bypass, and ID fan costs are essentially identical.

Space Velocity and Area Velocity

The space velocity is defined as the inverse of the residence time given by the equation:

$$V_{space} = \frac{1}{ResidenceTime} \tag{2.13}$$

Space velocity is calculated from the experimentally measured flue gas volumetric flow rate at the reactor inlet, represented as $q_{flue\,gas}$, and the reactor volume, represented as $Vol_{reactor}$, given by the equation:

$$V_{space} = \frac{q_{flue\,gas}}{Vol_{reactor}} \tag{2.14}$$

SCR system designers and vendors use the concept of area velocity, V_{area}, to account for the reaction being limited to active catalyst sites. The area velocity is calculated from the specific surface area of the catalyst per catalyst volume, $A_{specific}$, in the following equation:

$$V_{area} = \frac{V_{space}}{A_{specific}} \tag{2.15}$$

$A_{specific}$ is given in units of length^2/length^3 and must be provided by the catalyst manufacturer.

Theoretical NOx Removal

Equation 2.9 defines the NOx removal efficiency. However, in SCR NOx removal efficiency changes with catalyst activation.
The following theoretical equation allows for estimation of the NO_x removal efficiency, η_{NOx}, based on the catalyst activity constant, $K_{catalyst}$, at a specified time, t [1]. The theoretical NO_x removal efficiency is:

$$\eta_{NO_x} SR \quad (1 - e^{a}) \tag{2.16}$$

where

$$a = \left(\frac{-K_{catalyst} \quad A_{specific}}{V_{space}} \right) \tag{2.16a}$$

Both $K_{catalyst}$ and $A_{specific}$ are generally provided by the catalyst manufacturer.

According to this equation, the NO_x removal efficiency increases with increasing NH_3/NO_x ratio and decreasing space velocity (i.e., increasing catalyst volume for a given gas flow rate). In addition, the equation shows that as the activity of the catalyst decreases over time, the NO_x removal also decreases.

The theoretical ammonia slip in parts per million by volume can be calculated from the value of η_{NOx}, using the equation [14]:

$$Slip = \left(ASR - \eta_{NO_x} \right) \tag{2.17}$$

Catalyst Volume

The theoretical catalyst volume required for the SCR system is based on the factors discussed in Section 2.2 Process Description. Equation 2.17 can be rearranged to determine the theoretical catalyst volume [8]. Substituting the definition of space velocity into Equation 2.17, the volume of the catalyst is given by:

$$Vol_{catalyst} = \frac{-\left(q_{fluegas} \times ln\left[1 - \left(\frac{\eta_{NO_x}}{ASR} \right) \right] \right)}{K_{catalyst} \times A_{specific}} \tag{2.18}$$

An empirical equation was developed in Reference [1] as a function of several sensitivity variables. The sensitivity variables were determined from catalyst volume estimates obtained from catalyst suppliers for base and sensitivity cases. Adjustment factors for these variables were then developed using regression techniques.

The empirical equation for catalyst volume is given below:

$$Vol_{catalyst} = 2.81 \times Q_B \times \eta_{adj} \times Slip_{adj} \times NO_{x_{adj}} \times S_{adj} \times \frac{T_{adj}}{N_{SCR}} \tag{2.19}$$

and the adjustment factors include:

- NO_x efficiency adjustment factor:

$$\eta_{adj} = 0.2869 + (1.058 \times \eta) \tag{2.20}$$

- NO_x adjustment factor for inlet NO_x:

$$NO_{x_{adj}} = 0.8524 + \left(0.3208 \times NO_{x_{in}}\right) \tag{2.21}$$

- Ammonia slip adjustment factor for ammonia slips between 2 and 5 ppm:

$$Slip_{adj} = 1.2835 - \left(0.0567 \times Slip\right) \tag{2.22}$$

- Sulfur in coal adjustment factor:

$$S_{adj} = 0.9636 + \left(0.0455 \times S\right) \tag{2.23}$$

where S is the sulfur content of the fuel by weight fraction.

- The temperature adjustment factor for gas temperatures other than 700°F:

$$T_{adj} = 15.16 - \left(0.03937 \times T\right) + \left(2.74 \times 10^{-5} \times T^2\right) \tag{2.24}$$

where T is the temperature of the flue gas at the reactor inlet in degrees Fahrenheit (°F).

SCR Reactor Dimensions

The cross sectional area of the SCR reactor is sized for the flow rate of the flue gas in acfm and the superficial velocity. A typical value for the superficial velocity is 16 feet per second (960 ft/min). Using this value for velocity the equation for the catalyst cross-sectional area is given by:

$$A_{catalyst} = \frac{q_{fluegas}}{16\left(\frac{ft}{sec}\right) \times \left(\frac{60\ sec}{min}\right)} \tag{2.25}$$

The SCR reactor cross-sectional area is approximately 15 percent greater than the catalyst cross-sectional area to account for the module geometry and hardware:

$$A_{SCR} = 1.15 \times A_{catalyst} \tag{2.26}$$

The actual dimensions of the SCR depend on the module arrangement in the catalyst layer. The typical cross-sectional dimensions of a module are 3.3 feet wide by 6.6 feet long. Therefore, the SCR plan dimensions are approximately multiples of these dimensions. Depending on the number of modules in width and in length, the SCR reactor may be square or rectangular.

For the purposes of this report, the SCR reactor can be treated as a square. The screening costs are valid for rectangular SCR reactors as long as the aspect ratio (length divided by width) is not too large. Industry standard aspect ratios are between 1.0 and 1.5. For a square reactor, the length, l, and width, w, are estimated by:

$$l = w = (A_{SCR})^{1/2} \tag{2.27}$$

An initial value for the number of catalyst layers is estimated first. This estimate is then checked by calculating the catalyst height for each layer. The initial estimate for the number of catalyst layers can be determined from the total catalyst volume, the cross sectional area of the catalyst, and estimating the height of the catalyst element. A nominal height for the catalyst, h'_{layer}, is 3.1 feet. A first estimate for the number of catalyst layers, n_{layer}, is:

$$n_{layer} = \frac{Vol_{catalyst}}{h'_{layer} \times A_{catalyst}} \tag{2.28}$$

This value of n_{layer} is then rounded to the nearest integer. In addition, there must be at least two catalyst layers.

The height of each catalyst layer is calculated using the estimated number of layers. This must result in the height of a catalyst layer, h_{layer}, to be within the standard industry range of 2.5 to 5.0 feet. The height of a catalyst layer is calculated from the following equation:

$$h_{layer} = \left[\frac{Vol_{catalyst}}{n_{layer} \times A_{catalyst}} \right] + 1 \tag{2.29}$$

where 1 foot is added to account for space required above and below the catalyst material

for module assembly.

The number of catalyst layers calculated above does not include any empty catalyst layers for the future installation of catalyst. An empty catalyst layer is recommended for use

with a catalyst management plan. The total number of catalyst layers includes all empty catalyst layers that will be installed:

$$n_{total} = n_{layer} + n_{empty} \tag{2.30}$$

The height of the SCR reactor, including the initial and future catalyst layers, the flow-rectifying layer, space for sootblowers and catalyst loading, but excluding the inlet and outlet ductwork and hoppers is determined from the equation:

$$h_{SCR} = n_{total} \times \left(c_1 + h_{layer}\right) + c_2 \tag{2.31}$$

where the constants are based on common industry practice of $c_1 = 7$ and $c_2 = 9$.

Estimating Reagent Consumption and Tank Size

The rate of reagent consumption or mass flow rate of the reagent, , generally expressed as pounds per hour (lb/hr), can be calculated using the inlet NOx in lb/MMBtu and heat input rate in MMBtu/hr.

$$\dot{m}_{reagent} = \frac{NO_{x_{in}}\ Q_B\ NSR\ \eta_{NO_x}\ M_{reagent}}{M_{NO_x}\ SR_T} \tag{2.32}$$

where the $M_{reagent}$ is the molecular weight of ammonia (17.03 gram per mole) and M_{NOx} is the molecular weight of NO_2 (46.01 gram per mole). The molecular weight of NO_2 is used because the NO_x emissions, NO_{xin}, are given in lb/MMBtu of NO_2.

For ammonia, the mass flow rate of the aqueous reagent solution, is given by:

$$\dot{m}_{sol} = \frac{\dot{m}_{reagent}}{C_{sol}} \tag{2.33}$$

where C_{sol} is the percent concentration of the aqueous reagent solution by weight.

The solution volume flow rate, q_{sol}, generally expressed as gallons per hour (gph), is:

$$q_{sol} = \frac{\dot{m}_{sol}}{\rho_{sol}} \quad v_{sol} \tag{2.34}$$

where ρ_{sol} is the density of the aqueous reagent solution in lbs per cubic foot, 56.0 lb/ft^3 for a 29% solution ammonia at 60°F. The specific volume of a 29% solution ammonia at 60°F is 7.481 gal.ft^3.

The total volume stored in the tank, or tanks, is based on the volume that the SCR system requires for operating a specified number of days. The volume stored on site for the number of operating days, t, is:

$$Tank\ Volume = q_{sol}\ t \tag{2.35}$$

Note that the tank volume is typically based on full-load operation, so the capacity factor is not included in Equation 2.36. A common on site storage requirement is for 14 days of SCR operation.

2.4 Cost Analysis

The cost-estimating methodology presented here provides a tool to estimate study-level costs for high-dust SCR systems. Actual selection of the most cost-effective option should be based on a detailed engineering study and cost quotations from the system suppliers. The costs presented here are expressed in 1998 dollars.

The cost estimating equations presented in this section are based on equations developed by The Cadmus Group, Bechtel Power, Inc. and SAIC in the draft EPA report, Selective Catalytic Reduction for NO_x Control on Coal-fired Boilers, [1]. These equations follow the costing methodology of Electric Power Research Institute (EPRI)[23]. In the EPRI method, both the purchased equipment cost (PEC) and direct installation cost are estimated together. This methodology is different from the EPA Air Pollution Control Cost Manual methodology, which estimates equipment costs and installation costs separately. Due to the limited availability of equipment cost data and installation cost data, the equations for SCR capital costs were not reformulated.

The capital and annual cost equations were developed for coal-fired wall and tan-

gential utility and industrial boilers with heat input rates ranging from 250 MMBtu/hr to 6000 MMBtu/hr (25 MW to 600 MW). The SCR system design is a high-dust configuration with one SCR reactor per combustion unit. It utilizes anhydrous ammonia as the reagent with an allowed ammonia slip in the range of 2 to 5 ppm.. The catalyst is a ceramic honeycomb with an operating life of 3 years at full load operations. The cost equations are sufficient for NO_x reduction efficiencies up to 90% . A correction factor for a new installation versus a retrofit installation is included to adjust the capital costs. [1]

The cost information presented in this report is based on using ceramic honeycomb catalyst for the base case. In general, more catalyst volume is required for an SCR system using plate catalyst, although the unit cost of plate catalyst is lower than honeycomb. Thus, any difference in capital cost is expected to be within the accuracy of a study-level cost estimate.

This report is based on the high-dust SCR system because it is the most common design. A low-dust configuration would cost somewhat less because the required catalyst volume is smaller and ash hoppers on the SCR reactor are not required. The cost methodology is valid for a low-dust SCR system because the cost reductions are expected to be within the range of uncertainty for study-level costs. The costs for the tail-end arrangement, however, cannot be estimated from this report because they are significantly higher than the high-dust SCR systems due to flue gas reheating requirements.

2.4.1 Total Capital Investment

Total Capital Investment (TCI) includes direct and indirect costs associated with purchasing and installing SCR equipment. Costs include the equipment cost (EC) for the SCR system itself, the cost of auxiliary equipment, direct and indirect installation costs, additional costs due to installation such as asbestos removal, costs for buildings and site preparation, offsite facilities, land, and working capital. In general, SCR does not require buildings, site preparation, offsite facilities, land, and working capital. A more detailed discussion of capital costs can be found in Section 1, Chapter 2 of this Manual.

Direct Capital Costs

Direct capital costs (DCC) include purchased equipment costs (PEC) such as SCR system equipment, instrumentation, sales tax and freight. This includes costs associated with field measurements, numerical modeling and system design. It also includes direct installation costs such as auxiliary equipment (e.g., ductwork, fans, compressor), foundations and supports, handling and erection, electrical, piping, insulation, painting, and asbestos removal. Table 2.2 presents a list of the equipment and installation requirements for an

aqueous ammonia-based, high-dust SCR system.

The direct capital cost equation is based on data for one representative boiler and several sensitivity cases [2]. The direct capital cost equation includes the affect of specific design parameters on the SCR cost through the use of adjustment factors. Adjustment factors are based on correlations
of the sensitivity cases as functions of the catalyst volume, reactor height, and ammonia flow rate [2]. The design parameters and their affect on the capital cost equation are listed below:

- NO_x removal (catalyst volume);
- Inlet NO_x (catalyst volume);
- Ammonia slip (catalyst volume);
- Sulfur content in coal (catalyst volume);
- Flue gas temperature (catalyst volume);
- Catalyst management plan (cost of the SCR reactor);
- New plant versus retrofit (cost of the ductwork, structural steel, foundations); and
- With and without an SCR bypass, e.g., for seasonal operation (cost of ductwork)

The equation for direct capital cost (DCC) including the adjustment factors in dollars is:

$$DC = Q_B\left[\frac{\$3{,}380}{\left(\frac{MMBtu}{hr}\right)} + f(h_{SCR}) + f(NH_3 rate) + f(new) + f(bypass)\right]\left(\frac{3500}{Q_B}\right)^{0.35} + f(Vol_{catalyst}) \qquad (2.36)$$

Note that in the capital cost equation only the catalyst volume factor is not multiplied by the boiler heat input scaling factor $(3500/Q_B)^{0.35}$. The catalyst cost in \$/ft^3 is simply based on the volume of catalyst used and does not incorporate an economy of scale. The design parameter adjustment factors are given in the equations listed below.

Adjustment for the SCR reactor height:

$$f(h_{SCR}) = \left[\frac{\$6.12}{\frac{ft-MMBtu}{hr}}\ h_{SCR}\right] - \frac{\$187.9}{\left(\frac{MMBtu}{hr}\right)} \qquad (2.37)$$

Adjustment for the ammonia flow rate:

$$f(NH_3 rate) = \left[\frac{\$411}{lb/hr}\ \frac{\dot{m}_{reag}}{Q_B}\right] - \frac{\$47.3}{\left(\frac{MMBtu}{hr}\right)} \qquad (2.38)$$

for a retrofit

$$f(new) = \frac{\$0}{\left(\frac{MMBtu}{hr}\right)} \tag{2.39}$$

for a new boiler

$$f(new) = \frac{-\$728}{\left(\frac{MMBtu}{hr}\right)} \tag{2.40}$$

Adjustment for installing an SCR bypass.
no bypass installed

$$f(bypass) = \frac{\$0}{\left(\frac{MMBtu}{hr}\right)} \tag{2.41}$$

bypass is installed

$$f(new) = \frac{\$127}{\left(\frac{MMBtu}{hr}\right)} \tag{2.42}$$

The capital cost for the initial charge of catalyst:

$$f\left(Vol_{catalyst}\right) = Vol_{catalyst}\ CC_{initial} \tag{2.43}$$

where $Vol_{catalyst}$ is in ft^3 and $CC_{initial}$ is the cost of the initial catalyst, currently estimated at 240 $/ft^3$ for a ceramic honeycomb catalyst.

Indirect Capital Costs

Indirect installation costs are those associated with installing and erecting the control system equipment but do not contribute directly to the physical capital of the installation. This generally includes general facilities and engineering costs such as construction and contractor fees, preproduction costs such as startup and testing, inventory capital and any process and project contingency costs. As explained in the Manual Section 1 Introduction, average values of indirect installation factors are applied to the direct capital cost estimate to obtain values for indirect installation costs. These costs are estimated as a percentage of the TCI. The equations for indirect installation costs are presented in Table 2.5.

Table 2.5: Capital Cost Factors for an SCR Application

Description	Data or Formula
Boiler Size (MMBtu/hr)	Q_B
NOx Removal Efficiency	0_{NOx}
Cost Year	December 1998
Total Direct Capital Costs ($)	A (see eq. 2.37-2.44)
Indirect Installation Costs	
General Facilities ($)	0.05 × A
Engineering and Home Office Fees ($)	0.10 × A
Process Contingency ($)	0.05 × A
Total Indirect Installation Costs ($)	$B = A \times (0.05 + 0.10 + 0.05)$
Project Contingency ($)	$C = (A + B) \times 0.15$
Total Plant Cost ($)	$D = A + B + C$
Allowance for Funds During Construction ($)	$E = 0$ (Assumed for SCR)
Royalty Allowance ($)	$F = 0$ (Assumed for SCR)
Preproduction Cost ($)	$G = 0.02 \times (D + E)$
Inventory Capital* ($)	$H = Vol_{reagent}$ (gal) × $Cost_{reagent}$ ($/gal)
Initial Catalyst and Chemicals ($)	$I = 0$ (Assumed for SCR)
Total Capital Investment (TCI) ($)	$TCI = D + E + F + G + H + I$

* Cost for ammonia stored at site, i.e., the first fill of the reagent tanks

The initial capital cost includes the cost of the initial reagent volume, where RC is the reagent cost in $ per gallon, given by the equation:

$$Initial\ Capital\ Cost = Vol_{reagent}\ RC \qquad (2.44)$$

2.4.1 Total Annual Costs

Total annual costs (TAC) consist of direct costs, indirect costs, and recovery credits. Direct annual costs are those proportional to the quantity of waste gas processed by the control system. Indirect(fixed) annual costs are independent of the operation of the control system and would be incurred even if it were shut down. No byproduct recovery credits are included because there are no salvageable byproducts generated from the SCR [2]. Each of these costs is discussed in the sections below. A more detailed discussion of annual costs can be found in Section 1, Chapter 2 of this Manual.

Design parameters are estimated using the maximum annual heat input rate of the boiler to ensure adequate sizing of the SCR system. Annual costs are calculated using the average heat input rate of the boiler and SCR system using CF_{total}. This ensures that annual costs are based on the actual operating conditions rather than the design case.

Direct Annual Costs

Direct annual costs (DAC) include variable and semivariable costs. Variable direct annual costs account for purchase of reagent and electrical power. Semivariable direct annual costs include operating and supervisory labor cost, maintenance cost, and catalyst replacement cost. These costs are discussed individually below. Equations for these variable cost items were derived in Reference [1].

Operating costs also result from small decreases in boiler efficiency due to operation of the economizer bypass. The economizer bypass operation depends on the flow rate of gas bypassed at full and partial loads and the boiler's capacity factor. Another operating cost is incurred for the steam or electric power used for compressed air, as required for the relatively infrequent operation of sootblowers. These operating costs are generally small and site-specific. Therefore, they are not discussed in this report.

$$DAC = \left(\begin{array}{c}Annual\\Maintenance\\Cost\end{array}\right) + \left(\begin{array}{c}Annual\\Reagent\\Cost\end{array}\right) + \left(\begin{array}{c}Annual\\Electricity\\Cost\end{array}\right) + \left(\begin{array}{c}Annual\\Water\\Cost\end{array}\right) + \left(\begin{array}{c}Annual\\Catalyst\\Cost\end{array}\right) \qquad (2.45)$$

Operating and Supervisory Labor

The SCR reactor is a stationary device with no moving parts. Further, the SCR system incorporates only a few pieces of rotating equipment (e.g., pumps, motors, etc.). Therefore, the existing plant staff can operate the SCR from an existing control room. In general, operation of an SCR system does not require any additional operating or supervisory labor.

Maintenance

The annual maintenance labor and material cost in dollars per year ($/yr), including nozzle tip replacement for the injectors, is assumed to be 1.5% of the Total Capital Investment (*TCI*) in dollars. The equation is given by:

$$Annual\ Maintenance\ Cost = 0.015\ TCI \qquad (2.46)$$

Reagent Consumption

The annual cost of ammonia purchase in $/yr is estimated using the ammonia volume flow rate, the capacity factor, and the cost of reagent in dollars per gallon, $Cost_{reag}$:

$$Annual\ Reagent\ Cost = q_{reagent}\ Cost_{reag}\ t_{op} \tag{2.47}$$

where

$$t_{ops} = CF_{plant}\frac{8760\ hr}{yr} \tag{2.47a}$$

Utilities

The electrical power consumption in kilowatts is estimated for SCR equipment, ammonia vaporization, and additional ID fan power [1]. It is based on linear regression of electrical power consumption data correlated to the uncontrolled NO_x concentration, NO_{xin}, ASR and the boiler heat input, Q_B.

$$Power = 0.105\ Q_B\left[NO_{x_{in}}\ \eta_{NO_x} + 0.5\ \left(\Delta P_{duct} + n_{total}\ \Delta P_{catalyst}\right)\right] \tag{2.48}$$

The term, $(NO_{xin} \times \eta_{NOx})$, accounts for the ammonia vaporization and dilution air blower. The terms $0.5 \times (\Delta P_{duct} + n_{total} \times \Delta P_{catalyst})$, account for the additional fan power required to overcome the additional pressure drop of the SCR and ductwork. Typical values of ΔP_{duct} range between 2 to 3 inches of water; $\Delta P_{catalyst}$ usually ranges from 0.75 to 1 inch of water per catalyst layer.

The annual cost of electricity is estimated from the equation:

$$Annual\ Electricity\ Cost = Power\ Cost_{elect}\ t_{op} \tag{2.49}$$

where $Cost_{elect}$ is the electricity cost in dollars per kilowatt-hour ($/kWh).

Catalyst Replacement

The catalyst life is a function of the catalyst activity and ammonia slip. As the catalyst activity decreases with time, the ammonia slip increase until it reaches the design limit and new catalyst must be added. Catalyst life is usually specified when purchasing

the catalyst. A catalyst layer is typically guaranteed for 16,000 to 24,000 operating hours. This cost methodology assumes a guaranteed catalyst life of 24,000 hours or 3 years.

If the SCR does not have an empty catalyst layer, all of the catalyst layers must be replaced at the end of 24,000 operating hours. This very conservative assumption has been used in the SCR costs developed in the References [2] and [3]. If the SCR includes a spare catalyst layer, then only one catalyst layer is replaced at the end of 24,000 hours. Most SCR designs include a spare catalyst layer. The cost for catalyst replacement in all the SCR reactors for a given boiler, n_{SCR}, is given by:

$$Catalyst\ Replacement\ Cost = n_{SCR}\ Vol_{catalyst}\ \frac{CC_{replace}}{R_{layer}} \tag{2.50}$$

where $Vol_{catalystt}$ is in ft^3 $CC_{replace}$ is the cost of catalyst in dollars per foot cubed ($/ft^3) and R_{layer} is a factor for catalyst replacement. $R_{layer} = 1$ for full replacement and $R_{layer} = n_{layer}$ for replacing 1 layer per year.

Because the catalyst is replaced every few years, the annual catalyst cost for all reactors is a function of the future worth of the catalyst given by:

$$Annual\ Catalyst\ Replacement\ Cost = \begin{pmatrix} Catalyst \\ Replacement \\ Cost \end{pmatrix} FWF \tag{2.51}$$

where *FWF* is the future worth factor. Future worth is used because the annual catalyst replacement cost is accrued starting in the first year of operation, while catalyst replacement purchases occur every few years. To account for the time value of money, the *FWF* amortizes the catalyst cost over the years preceding the actual catalyst purchase [21]. Because the money is allocated in advance of the purchase, the sum of the annual catalyst replacement costs is less than the purchase price of the catalyst. The future worth factor, *FWF* is given by:

$$FWF = i\left[\frac{1}{(1+i)^Y - 1}\right] \tag{2.52}$$

where i is the assumed interest rate and Y is the term in years.

The term, Y is given by the equation:

$$Y = \frac{h_{catalyst}}{h_{year}} \tag{2.53}$$

where $h_{catalyst}$ is the operating life of the catalyst in hours and h_{year} is the number of hours per year the SCR is operated. The value of Y estimated from the equation is then rounded to the nearest integer.

Indirect Annual Costs

In general, indirect annual costs (fixed costs) include the capital recovery cost, property taxes, insurance, administrative charges, and overhead. Capital recovery cost is based on the anticipated equipment lifetime and the annual interest rate employed. An economic lifetime of 20 years is assumed for the SCR system. The remaining life of the boiler may also be a determining factor for the system lifetime.

In many cases property taxes do not apply to capital improvements such as air pollution control equipment, therefore, for this analysis, taxes are assumed to be zero [19]. The cost of overhead for an SCR system is also considered to be zero. An SCR system is not viewed as risk-increasing hardware (e.g., a high energy device such as a boiler or a turbine). Consequently, insurance on an SCR system is on the order of a few pennies per thousand dollars annually [19]. The administrative charges, covering sales, research and development, accounting, and other home office expenses, incurred in operation of an SCR system are relatively insignificant for the cost estimation procedure presented here. Finally, there are two categories of overhead, payroll and plant. Payroll overhead includes expenses related to labor employed in operation and maintenance of hardware; whereas plant overhead accounts for items such as plant protection, control laboratories, and parking areas. Because this procedure assumes that no additional labor is needed in operation of an SCR system, payroll overhead is zero and plant overhead is considered to be negligible.

Using these assumptions, indirect annual costs in $/yr, $IDAC$, can be expressed as:

$$IDAC = CRF \; TCI \tag{2.54}$$

where *TCI* is the total investment and *CRF* is the capital recovery factor and defined by:

$$CRF = \frac{i}{\left(1-(1+i)^{-n}\right)} \tag{2.55}$$

where i is the interest rate, and n is the year the cost is incurred.

Total Annual Cost

The total annual cost (*TAC*) for owning and operating an SCR system is the sum of direct and indirect annual costs as given in the following equation:

$$Total\ Annual\ Cost = \begin{pmatrix} Direct \\ Annual \\ Cost \end{pmatrix} + \begin{pmatrix} Indirect \\ Annual \\ Cost \end{pmatrix} \tag{2.56}$$

The tons of NO_x removed annually are

$$NO_x\ Removed = NO_{x_{in}}\ \eta_{NO_x}\ Q_B\ t_{op} \tag{2.57}$$

where NO_x is in lb/MMBtu, Q_B is in MMBtu/hr and tops is in hours and a conversion of 2,000 lb/ton is required.

The cost in dollars per ton of NO_x removed per year is:

$$Cost\ Effectiveness = \frac{TAC}{NO_x\ Removed} \tag{2.58}$$

where TAC is in $/yr and NO_x removed is tons/yr.

2.5 Example Problem

An example problem, which calculates both the design parameters and capital and annual costs, is presented below. The design basis is a retrofit SCR system being applied to a 1000 MMBtu/hr, wall-fired, industrial boiler firing subbituminous coal. The following assumptions are made to perform the calculations:

Fuel High Heating Value	10,000 Btu/lb
Maximum Fuel Consumption Rate	1.0×10^5 lb/hr
Average Annual Fuel Consumption	4.38×10^8 lb
Number of SCR operating days	155 days
Plant Capacity Factor	50%
Uncontrolled NO_x Concentration	0.86 lb/MMBtu
Required Controlled NO_x Concentration	0.13 lb/MMBtu
Acceptable Ammonia Slip	2.0 ppm
Fuel Volume Flow Rate, Eastern Bituminous	484 ft^3/min per MMBtu/hr
Fuel Heating Value	12,696 Btu/lb
Fuel Sulfur Content	1.0% by weight
Fuel Ash Content	7.7% by weight
ASR	1.05
Stored Ammonia Concentration	29%
Number of Days of Storage for Ammonia	14 days
Pressure Drop for SCR Ductwork	3 inches w.g.
Pressure Drop for each Catalyst Layer	1 inch w.g.
Temperature at SCR Inlet	650°F

In addition to these assumptions, the estimated economic factors for the cost equations are:

Cost year	December, 1998
Equipment Life	20 years
Annual Interest Rate	7%
Catalyst Cost, Initial	240 $/$ft^3$
Catalyst Cost, Replacement	290 $/$ft^3$
Electrical Power Cost	0.05 $/kWh
29% Ammonia Solution Cost	0.101 $/lb
Operating Life of Catalyst	24,000 hours
Catalyst Layers	2 full, 1 empty

2.5.1 Design Parameter Example

Boiler Calculations

The boiler annual heat input rate, Q_B, is calculated from the High Heating Value for subbituminous coal given in Table 1.3 and the maximum fuel consumption rate, $\dot{m}_{fuel}$:

$$Q_B = \frac{10{,}000\,\frac{Btu}{lb}\;\;100{,}000\,\frac{lb}{hr}}{10^6\,\frac{Btu}{MMBtu}} = 1{,}000\;\frac{MMBtu}{lb}$$

The plant capacity factor is calculated from the maximum and annual average fuel consumption:

$$CF_{plant} = \frac{4.38\;10^8\;lb}{1\times10^5\left(\frac{lb}{hr}\right)\times 8760\,\frac{hr}{yr}} = 0.5 = 50\%$$

The SCR sytem capacity factor is calculated from themonths of SCR operation, 5 months:

$$CF_{SNCR} = \frac{155\;days}{365\;days} = 0.42 = 42\%$$

The total capacity factor including both plant and SCR capacity factors is given by:

$$CF_{total} = 0.5\times0.42 = 0.21 = 21\%$$

The flue gas flow rate using Equation (2.12) is:

$$q_{flue\;gas} = \frac{\frac{484\;ft^3}{min\text{-}\left(\frac{MMBtu}{hr}\right)}\times 1{,}000\,\frac{MMBtu}{hr}\times(460+650°F)}{(460+700°F)\times1} = 463{,}138\;acfm$$

The NO_x removal efficiency, η_{NOx}, is calculated from the inlet NO_x concentration and the required controlled NO_x concentration using Equation (2.16):

$$\eta_{NOx} = \frac{0.86 \frac{lb}{MMBtu} - 0.13\ MMBtu}{0.86 \frac{lb}{MMBtu}} = 0.85 = 85\%$$

SCR Reactor Calculations

The catalyst volume using Equation (2.20) and the equations for each adjustment factor is:

$Vol_{catalyst}$ = 2.81 × 1,000 MBtu/hr
× (0.2869 + (1.058 × 0.85)) (ηNO_x)
× (0.8524 + (0.3208 × 0.86)) (NO_{xin})
× (1.2835 - (0.0567 × 2.0)) (Slip)
× (0.9636 + (0.0455 × 1.0)) (Sulfur)
× (15.16 - (0.03937 × 650) + (0.0000274 × 650^2)) (Temperature)
= 5089 ft^3

The catalyst and SCR cross-sectional areas using Equations (2.26) and (2.27) are:

$$A_{catalyst} = \frac{463{,}138\ acfm}{\left(16 \frac{ft}{s}\right)\left(\frac{60s}{min}\right)} = 482\ ft^2$$

$$A_{SCR} = 1.15 \times 482\ ft^2 = 554\ ft^2$$

The length and width of the reactor using Equation (2.28) is:

$$l = w = (554)^{1/2} = 23.5\ ft$$

The first estimate of the number of catalyst layers using Equation (2.29) is:

$$n_{layer} = \frac{5089}{3.1 \times 482} = 3.4$$

Rounding this value gives, $n_{layer} = 3$.

Checking the actual catalyst height using Equation (2.29):

$$h_{layer} = \frac{5089\ ft^3}{3 \times 482\ ft} + 1 = 4.5$$

This value is between the design height limits of 2.5 and 5 feet.

The total number of catalyst layers is determined by Equation (2.30) with 1 empty catalyst layer:

$$n_{total} = 3 + 1 = 4$$

The SCR height, excluding the outlet duct and hoppers using Equation (2.31) is:

$$h_{SCR} = 4 \times (7 + 4.5) + 9 = 55\ ft$$

Reagent Calculations

The mass flow rate of the reagent is calculated using the molecular weight of the reagent, 17.03 g/mole and NO_2, 46.01g/mole. For an ASR of 1.05, the reagent mass flow rate is given by Equation (2.32):

$$\dot{m}_{reagent} = \frac{0.86\ \frac{lb}{MMBtu} \times 1{,}000\ \frac{MMBtu}{hr} \times 1.05 \times 17.03\ \frac{g}{mole}}{46.01\ \frac{g}{mole}} = 334\ \frac{lb}{hr}$$

The mass flow rate of 29% aqueous ammonia solution is given by Equation (2.33):

$$\dot{m}_{sol} = \frac{334\ \frac{lb}{hr}}{0.29} = 1{,}152\ \frac{lb}{hr}$$

The solution volume flow rate can then be calculated from Equation (2.34) where p is the density of the29% aqueous ammonia solution, 56.0 lb/ft^3 at 60° F and the specific volume of the solution is 7.481 gal/ft^3.

$$q_{sol} = \frac{1{,}152\ \frac{lb}{hr} \times 7.481\ \frac{gal}{ft^3}}{56.0\ \frac{lb}{ft^3}} = 154\ gph$$

The total volume stored in the tank, or tanks, is based on the volume that the SCR system requires for 14 days of operation. The on site storage requirement is given by Equation (2.35):

$$Tank\ Volume = 154\ gph \times 14\ (days) \times \left(\frac{24\ hr}{day}\right) = 51{,}744\ gal$$

The on site storage requirement for ammonia is 51,744 gallons per 14 days of operation.

2.5.2 Cost Estimation Example

Once the SCR system is sized, the capital and annual costs for the SCR system can be estimated. The DCC are estimated using the Equation (2.36):

$$DC = 1{,}000\ \frac{MMBtu}{hr} \times \left[3{,}380 + f(h_{SCR}) + f(NH_3) + f(new) + f(bypass)\right]$$

$$\times \left(\frac{3{,}500\ MMBtu/hr}{1{,}000\ MMBu/hr}\right)^{0.35} + (Vol_{catalyst})$$

where the funcions in units of dollars per MMBtu/hr are given by:

$$f(h_{SCR}) = \left[(6.12 \times 55) - 187.9\right] = \frac{\$\ 149}{MMBtu/hr}$$

$$f(NH^3) = \left[\frac{334\ \frac{lb}{hr}}{1000\ \frac{MMBtu}{hr}} \times \frac{\$411}{lb/hr}\right] - \frac{\$47.3}{\frac{MMBtu}{hr}} = 90\ \frac{\$}{\frac{MMBtu}{hr}}$$

$$f(new) = 0, for\ retrofit$$

$$f(bypass) = 0, for\ no\ bypass$$

The function $Vol_{catalyst}$ is the initial cost of the catalyst given by Equation (2.43):

$$f(Vol_{catalyst}) = 5089\ ft^3 \times \frac{\$\ 240}{ft^3} = \$1{,}221{,}360$$

Substituting the value obtained for the factors gives:

$$DC = 1{,}000\frac{MMBtu}{hr} \times [3{,}380 + 149 + 90 + 0 + 0]\frac{\$}{MMBtu} \times \left(\frac{3{,}500}{1{,}000}\right)^{0.35} + \$1{,}221{,}360 = \$6{,}832{,}000$$

Indirect capital costs are estimated from the direct capital costs using the following

Table 2.5: Indirect Capital Cost Estimations

Indirect Installation Costs	
General Facilities ($)	$6,832,000 × 0.05 = $341,600
Engineering and Home Office Fees ($)	$6,832,000 × 0.10 = $683,200
Process Contingency ($)	$6,832,000 × 0.05 = $341,600
Total Indirect Installation Costs ($)	$341,600+ $683,200 + $341,600 = $1,366,400
Project Contingency ($)	($6,832,000 + 1,366,400) × 0.15 = $1,229,760
Total Plant Cost ($)	$6,832,000 + $1,366,400 + $1,229,760 = $9,428,160
Preproduction Cost ($)	$9,428,160 × 0.02 = $188,562
Inventory Capital ($)	$0.101/lb × 1,152 lb/hr × 24hr/day × 14 days = $39,094
Total Capital Investment (TCI) ($)	$9,428,160 + $188,562 + $39,094 = $9,655,816

equations in Table 2.5:

Annual costs are based on the economic factors listed above. In addition, the SCR system is assumed to operate for 5 months of the year with a boiler loading of 50%, resulting in a total capacity factor of 27%. The annual variable costs are given by:

$$Maintenance\ Cost = \frac{0.015}{yr} \times 9{,}601{,}261 = \frac{\$\ 144{,}019}{yr}$$

An estimate for power consumption is given by:

$$Power = \frac{1000\ MMBtu/hr}{9.5\ \frac{MMBtu/hr}{kW}} [(0.86 \times 0.85) + 0.5\ (3 + 4 \times 1)] = 445\ kW$$

Electricity and Reagent Colution cost can then be estimated from Equation (2.49) and (2.47):

$$Electricity\ Cost = 445\ kW \times \frac{8{,}760\ hr}{yr} \times 0.27 \times \frac{0.05\ \$}{kWh} = \frac{\$\ 52{,}626}{yr}$$

$$Reagent\ Solution\ Cost = 1152\ \frac{lb}{hr} \times 8{,}760\ \frac{hr}{yr} \times 0.27 \times 0.101\ \frac{\$}{lb} = \frac{\$\ 275{,}196}{yr}$$

The catalyst is replaced every few years, the annual catalyst cost for all reactors is a function of the future worth of the catalyst given by:

$$FWF = 0.07 \times \frac{1}{(1 + 0.07)^6 - 1} = 0.14$$

It is assumed that 1 layer of catalyst is replaced per year, therefore, $R_{layer} = 3$. The operating life of the catalyst is given as 24,000 hours and the estimated SCR operation hours per year is 3,720 hours which gives the value of Y in Equation (2.53) as:

$$Y = \frac{24000\ hours}{3720\ hours} = 6.4 = 6$$

The cost of the catalyst replacement, for one layer replaced per year, is given by Equation (2.50) and (2.51):

$$Annual\ Catalyst\ Replace\ Cost = 0.14 \times \left[1 \times 5089\ ft^3 \times \frac{290\ \frac{\$}{ft^3}}{3} \right] = \frac{\$68{,}871}{yr}$$

The total variable direct annual cost, the sum of the cost of the reagent, electricity, and catalyst replacement, is given by Equation (2.45):

$$Total\ Variable\ Direct\ Cost = \frac{\$52{,}626}{yr} + \frac{\$275{,}196}{yr} + \frac{\$68{,}871}{yr} = \frac{\$396{,}693}{yr}$$

and the total direct annual cost is given by the sum of the maintenance and variable cost:

$$Total\ Direct\ Annual\ Cost = \frac{\$144{,}837}{yr} + \frac{\$396{,}693}{yr} = \frac{\$541{,}530}{yr}$$

In estimating total indirect annual costs, it is assumed that the property tax factor, F_{tax}, and the overhead factor, F_{ovhd}, are both zero. The capital recovery factor, *CRF*, is defined by:

$$CRF = \frac{0.07(1 + 0.07)^{20}}{(1 + 0.07)^{20} - 1} = 0.0944$$

and the indirect annual costs (IDAC) are calculated from Equation (2.55) and the TCI:

$$Indirect\ Annual\ Cost = \frac{0.0944}{yr} \times \$9{,}655{,}816 = \frac{\$911{,}412}{yr}$$

The total annual cost is the sum of the direct annual and indirect annual costs given by Equation (2.56)

$$Total\ Annual\ Cost = \frac{\$\ 541{,}530}{yr} + \frac{\$\ 911{,}412}{yr} = \frac{\$\ 1{,}452{,}942}{yr}$$

The annual cost in terms of NO_x removed can be calculated using the total annual cost and the tons of NO_x removed annually by Equation (2.58):

$$NOx\ removed = \frac{\frac{0.86\ lb}{MMBtu\ /\ hr} \times 0.85 \times 1{,}000 \frac{MMBtu}{hr} \times 0.27 \times 8{,}760 \frac{hr}{yr}}{2{,}000 \frac{lb}{ton}} = 864\ tons$$

and Equation (2.59)

$$Cost\ of\ NOx\ Removal = \frac{\frac{\$\ 1{,}452{,}492}{yr}}{864\ tons} = \frac{\$1{,}682}{tons}$$

References

[1] Selective Catalytic Reduction for NO_x Control on Coal-fired Boilers, Draft Report, prepared for the U.S. Environmental Protection Agency by The Cadmus Group, Inc., Bechtel Power Corporation, and Science Applications International Corporation. May 1998.

[2] Investigation of Performance and Cost of NO_x Controls as Applied to Group 2 Boilers. Revised Draft Report, prepared for the U.S. Environmental Protection Agency by The Cadmus Group, Inc., Bechtel Power Corporation, and Science Applications International Corporation. August 1996.

[3] Cost Estimates for Selected Applications of NO_x Control Technologies on Stationary Combustion Boilers, Draft Report and Responses to Comments on the Draft Report. prepared for the U.S. Environmental Protection Agency by The Cadmus Group, Inc., and Bechtel Power Corporation. June 1997.

[4] Institute of Clean Air Companies (ICAC). Selective Catalytic Reduction (SCR) Control of NO_x Emissions, Prepared by the ICAC SCR Committee, November 1997.

[5] U.S. EPA. Analyzing Electric Power Generation Under the CAAA. EPA Office of Air and Radiation, March 1998.

[6] Rummenhohl, V., H. Weiler, and W. Ellison. Experience Sheds Light on SCR O&M Issues. Power Magazine, September 1992.

[7] Pritchard, S., C. DiFrancesco, and T.R. von Alten. SCR Catalyst Performance Under Severe Operating Conditions. Presented at Power-Gen International '96 Conference and Exhibition, ASME, Orlando, FL. December 1996.

[8] Harmgart, S., M. Meierer, and J. Fahlke. In-Service Assessment of $DeNO_x$ Catalysts. Presented at the International Joint Power Generation Conference and Exposition '98, ASME, Baltimore, MD, 1998.

[9] Cochran, J.R., M.G. Gregory, and V. Rummenhohl. The Effect of Various Parameters on SCR System Cost. Presented at Power-Gen '93, the Fourth International Power Generation Exhibition & Conference, ASME, Dallas, Texas, November 1993.

[10] Electric Power research Institute. Technical Feasibility and Cost of Selective Catalytic Reduction (SCR) NO_x Control. EPRI GS-7266, Final Report, May 1991.

[11] Rosenberg, H.S. and J. H. Oxley. Selective Catalytic Reduction for NO_x Control at Coal-fired Power Plants. ICAC Forum '93, Controlling Air Toxics and NO_x Emissions, Baltimore, MD, February 24-26, 1993.

[12] Singer, J.G. (ed.) Combustion Fossil Power Systems. Third edition, Combustion Engineering, Inc., Windsor, CT, 1981.

[13] Sigling, R. and R. Johnson. Current Practices for Monitoring Ammonia Slip from SCR Processes. Presented at the 1998 Conference on Selective Catalytic and Non-catalytic Reduction for NO_x Control, U.S. Department of Energy, Federal Energy Technology Center, Pittsburgh, PA, May 21-22, 1998.

[14] Cho, S.M. and S. Z. Dubow. Design of a Selective Catalytic Reduction System for NO_x Abatement in a Coal-Fired Cogeneration Plant. Proceedings of the American Power Conference, Chicago, IL, pp. 717-722, April 1992.

[15] ANSI K-61.1-1989. Safety Requirements for the Storage and Handling of Anhydrous Ammonia. Compressed Gas Association, Arlington, VA, March 1989.

[16] Tanner Industries, Inc. Storage & Handling of Anhydrous Ammonia. Philadelphia, PA, Revision 5/96.

[17] Tanner Industries, Inc. Storage & Handling of Ammonium Hydroxide. Philadelphia, PA, Revision 10/95.

[18] Broske, D., G. Offen, and E. Cichanowicz. An Assessment of European Experience with Selective Catalytic Reduction for NO_x Control. Presented at the "NO_x Controls for Utility Boilers Workshop," Baltimore, MD, August 1998, ("Proceedings: 1998 NO_x Control Workshop," EPRI report number TR-1113561998.).

[19] Staudt, J.E. Status Report on NO_x Control Technologies and Cost Effectiveness for Utility Boilers. Published by Northeast States for Coordinated Air Use Management (NESCAUM) and Mid-Atlantic Regional Air Management Association (MARAMA), June 1998.

[20] S.C. Stultz and J. B. Kitto, eds. Steam, Its Generation and Use. 40th edition, Babcock& Wilcox, Inc., Barberton, Ohio, 1992.

[21] Green, D.W. and R.H. Perry, eds. Perry's Chemical Engineers' Handbook. 6th edition, McGraw Hill, Inc., 1984.

[22] Tonn, D.P. and T. A. Uysal. 2200 MW SCR Installation on New Coal-Fired

Project. Presented at the Institute of Clean Air Companies ICAC Forum '98, Durham, North Carolina, March 18-20, 1998.

[23] Electric Power Research Institute. Technical Assessment Guide. Vol. 1, Revision 7, 1993.

[24] Babb, Bob, Ed Angelini and Scot Pritchard. Implemenation of SCR System at TVA Paradise Unit 2, Presented at the Institute of Clean Air Companies ICAC Forum 2000, Washington D.C., March 23-24, 2000.

[25] Stenger, H. G., et al, Low Temperature Selective Catalytic Reduction of NO, Presented at the Institute of Clean Air Companies ICAC Forum 2000, Washington D.C., March 23-24, 2000.

EPA/452/B-02-001

Section 5

SO_2 and Acid Gas Controls

EPA/452/B-02-001

Section 5.2

Post-Combustion Controls

EPA/452/B-02-001

Chapter 1

WET SCRUBBERS FOR ACID GAS

Wiley Barbour
Roy Oommen
Gunseli Sagun Shareef
Radian Corporation
Research Triangle Park, NC 27709

William M. Vatavuk
Innovative Strategies and Economics Group, OAQPS
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

December 1995

Contents

1.0 Introduction 1-3
1.1 System Efficiencies and Performance 1-3

1.2 Process Description 1-4
1.2.1 Absorber System Configuration 1-4
1.2.2 Types of Absorption Equipment 1-5
1.2.3 Packed Tower Interals 1-6
1.2.4 Packed Tower Operation 1-9

1.3 Design Procedures 1-10
1.3.1 Determining Gas and Liquid Stream Conditions 1-11
1.3.2 Determining Absorption Factor 1-15
1.3.3 Determining Column Diameter 1-16
1.3.4 Determining Tower Height and Surface Area 1-18
1.3.5 Calculating Column Pressure Drop 1-20
1.3.6 Alternative Design Procedure 1-21

1.4 Estimating Total Capital Investment 1-23
1.4.1 Equipment Costs for Packed Towers 1-23
1.4.2 Installation Costs 1-25

1.5 Estimating Annual Cost 1-26
1.5.1 Direct Annual Costs 1-26
1.5.2 Indirect Annual Costs 1-28
1.5.3 Total Annual Costs 1-30

1.6 Example Problem 1-30
1.6.1 Required Information for Design 1-30
1.6.2 Determine Gas and Liquid Stream Properties 1-31
1.6.3 Calculate Absorption Factor 1-34
1.6.4 Estimate Column Diameter 1-34
1.6.5 Calculate Column Surface Area 1-36
1.6.6 Calculate Pressure Drop 1-37
1.6.7 Equipment Costs 1-37
1.6.8 Total Annual Cost 1-39
1.6.9 Alternate Example 1-43

1.7 Acknowledgements 1-43

References 1-44

Appendix A 1-46
Appendix B 1-50
Appendix C 1-54

1.0 Introduction

Gas absorbers are used extensively in industry for separation and purification of gas streams, as product recovery devices, and as pollution control devices. This chapter focuses on the application of absorption for pollution control on gas streams with typical pollutant concentrations ranging from 250 to 10,000 ppmv. Gas absorbers are most widely used to remove water soluble inorganic contaminants from air streams.[1, 2]

Absorption is a process where one or more soluble components of a gas mixture are dissolved in a liquid (i.e., a solvent). The absorption process can be categorized as physical or chemical. Physical absorption occurs when the absorbed compound dissolves in the solvent; chemical absorption occurs when the absorbed compound and the solvent react. Liquids commonly used as solvents include water, mineral oils, nonvolatile hydrocarbon oils, and aqueous solutions.[1]

1.1 System Efficiencies and Performance

Removal efficiencies for gas absorbers vary for each pollutant-solvent system and with the type of absorber used. Most absorbers have removal efficiencies in excess of 90 percent, and packed tower absorbers may achieve efficiencies as high as 99.9 percent for some pollutant-solvent systems.[1, 3]

The suitability of gas absorption as a pollution control method is generally dependent on the following factors: 1) availability of suitable solvent; 2) required removal efficiency; 3) pollutant concentration in the inlet vapor; 4) capacity required for handling waste gas; and, 5) recovery value of the pollutant(s) or the disposal cost of the spent solvent.[4]

Physical absorption depends on properties of the gas stream and solvent, such as density and viscosity, as well as specific characteristics of the pollutant(s) in the gas and the liquid stream (e.g., diffusivity, equilibrium solubility). These properties are temperature dependent, and lower temperatures generally favor absorption of gases by the solvent.[1] Absorption is also enhanced by greater contacting surface, higher liquid-gas ratios, and higher concentrations in the gas stream.[1]

The solvent chosen to remove the pollutant(s) should have a high solubility for the gas, low vapor pressure, low viscosity, and should be relatively inexpensive.[4] Water is the most common solvent used to remove inorganic contaminants; it is also used to absorb organic compounds having relatively high water solubilities. For organic compounds that have low water solubilities, other solvents such as hydrocarbon oils are used, though only in industries where large volumes of these oils are available (i.e., petroleum refineries and petrochemical plants).[5]

Pollutant removal may also be enhanced by manipulating the chemistry of the absorbing solution so that it reacts with the pollutant(s), e.g., caustic solution for acid-gas absorption vs. pure water as a solvent. Chemical absorption may be limited by the rate of reaction, although the rate limiting step is typically the physical absorption rate, not the chemical reaction rate.

1.2 Process Description

Absorption is a mass transfer operation in which one or more soluble components of a gas mixture are dissolved in a liquid that has low volatility under the process conditions. The pollutant diffuses from the gas into the liquid when the liquid contains less than the equilibrium concentration of the gaseous component. The difference between the actual concentration and the equilibrium concentration provides the driving force for absorption.

A properly designed gas absorber will provide thorough contact between the gas and the solvent in order to facilitate diffusion of the pollutant(s). It will perform much better than a poorly designed absorber.[6] The rate of mass transfer between the two phases is largely dependent on the surface area exposed and the time of contact. Other factors governing the absorption rate, such as the solubility of the gas in the particular solvent and the degree of the chemical reaction, are characteristic of the constituents involved and are relatively independent of the equipment used.

1.2.1 Absorber System Configuration

Gas and liquid flow through an absorber may be countercurrent, crosscurrent, or cocurrent. The most commonly installed designs are countercurrent, in which the waste gas stream enters at the bottom of the absorber column and exits at the top. Conversely, the solvent stream enters at the top and exits at the bottom. Countercurrent designs provide the highest theoretical removal efficiency because gas with the lowest pollutant concentration contacts liquid with the lowest pollutant concentration. This serves to maximize the average driving force for absorption throughout the column.[2] Moreover, countercurrent designs usually require lower liquid to gas ratios than cocurrent and are more suitable when the pollutant loading is higher.[3, 5]

In a crosscurrent tower, the waste gas flows horizontally across the column while the solvent flows vertically down the column. As a rule, crosscurrent designs have lower pressure drops and require lower liquid-to-gas ratios than both cocurrent and countercurrent designs. They are applicable when gases are highly soluble, since they offer less contact time for absorption.[2, 5]

In cocurrent towers, both the waste gas and solvent enter the column at the top of the tower and exit at the bottom. Cocurrent designs have lower pressure drops, are not subject to flooding limitations and are more efficient for fine (i.e., submicron) mist removal. Cocurrent designs are only efficient where large absorption driving forces are available. Removal efficiency is limited since the gas-liquid system approaches equilibrium at the bottom of the tower.[2]

1.2.2 Types of Absorption Equipment

Devices that are based on absorption principles include packed towers, plate (or tray) columns, venturi scrubbers, and spray chambers. This chapter focuses on packed towers, which are the most commonly used gas absorbers for pollution control. Packed towers are columns filled with packing materials that provide a large surface area to facilitate contact between the liquid and gas. Packed tower absorbers can achieve higher removal efficiencies, handle higher liquid rates, and have relatively lower water consumption requirements than other types of gas absorbers.[2] However, packed towers may also have high system pressure drops, high clogging and fouling potential, and extensive maintenance costs due to the presence of packing materials. Installation, operation, and wastewater disposal costs may also be higher for packed bed absorbers than for other absorbers.[2] In addition to pump and fan power requirements and solvent costs, packed towers have operating costs associated with replacing damaged packing.[2]

Plate, or tray, towers are vertical cylinders in which the liquid and gas are contacted in step-wise fashion on trays (plates). Liquid enters at the top of the column and flows across each plate and through a downspout (downcomer) to the plates below. Gas moves upwards through openings in the plates, bubbles into the liquid, and passes to the plate above. Plate towers are easier to clean and tend to handle large temperature fluctuations better than packed towers do.[4] However, at high gas flow rates, plate towers exhibit larger pressure drops and have larger liquid holdups. Plate towers are generally made of materials such as stainless steel, that can withstand the force of the liquid on the plates and also provide corrosion protection. Packed columns are preferred to plate towers when acids and other corrosive materials are involved because tower construction can then be of fiberglass, polyvinylchloride, or other less costly, corrosive-resistant materials. Packed towers are also preferred for columns smaller than two feet in diameter and when pressure drop is an important consideration.[3, 7]

Venturi scrubbers are generally applied for controlling particulate matter and sulfur dioxide. They are designed for applications requiring high removal efficiencies of submicron particles, between 0.5 and 5.0 micrometers in diameter.[4] A venturi scrubber employs a gradually converging and then diverging section, called the throat, to clean incoming gaseous streams. Liquid is either introduced to the venturi upstream of the throat or injected directly into the throat where it is atomized by the gaseous stream. Once the liquid is atomized, it collects particles from the gas and discharges from the venturi.[1] The high pressure drop through these systems results in high energy use, and the relatively short gas-liquid contact time restricts their application to highly soluble gases. Therefore, they are infrequently used for the control of volatile organic compound emissions in dilute concentration.[2]

Spray towers operate by delivering liquid droplets through a spray distribution system. The droplets fall through a countercurrent gas stream under the influence of gravity and contact the pollutant(s) in the gas.[7] Spray towers are simple to operate and maintain, and have relatively low

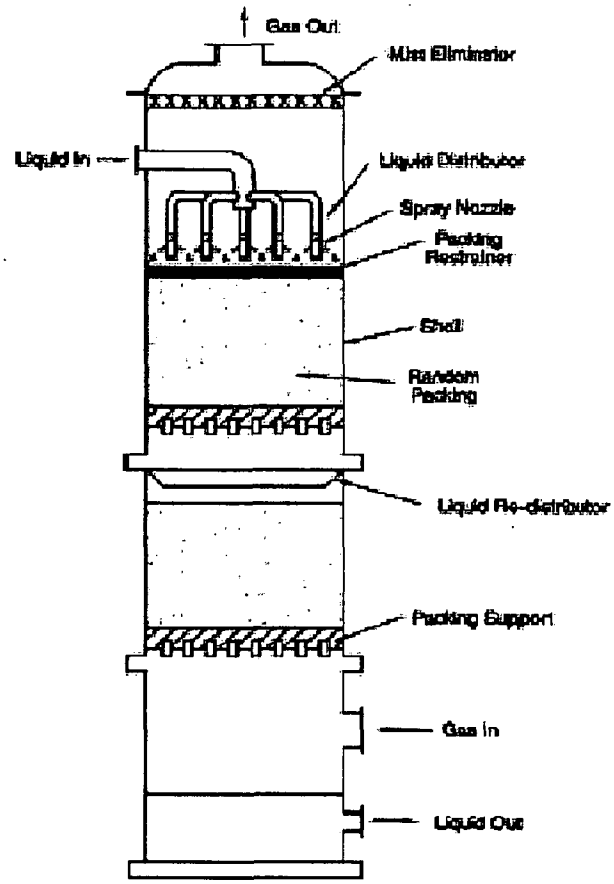


Figure 1.1: Packed Tower for Gas Absorption

energy requirements. However, they have the least effective mass transfer capability of the absorbers discussed and are usually restricted to particulate removal and control of highly soluble gases such as sulfur dioxide and ammonia. They also require higher water recirculation rates and are inefficient at removing very small particles.[2, 5]

1.2.3 Packed Tower Internals

A basic packed tower unit is comprised of a column shell, mist eliminator, liquid distributors, packing materials, packing support, and may include a packing restrainer. Corrosion resistant alloys or plastic materials such as polypropylene are required for column internals when highly corrosive solvents or gases are used. A schematic drawing of a countercurrent packed tower is shown in Figure 1.1. In this figure, the packing is separated into two sections. This configuration is more expensive than designs where the packing is not so divided.[5]

The tower shell may be made of steel or plastic, or a combination of these materials depending on the corrosiveness of the gas and liquid streams, and the process operating conditions. One alloy that is chemical and temperature resistant or multiple layers of different, less expensive materials may be used. The shell is sometimes lined with a protective membrane, often made from a corrosion resistant polymer. For absorption involving acid gases, an interior layer of acid resistant brick provides additional chemical and temperature resistance.[8]

At high gas velocities, the gas exiting the top of the column may carry off droplets of liquid as a mist. To prevent this, a mist eliminator in the form of corrugated sheets or a layer of mesh can be installed at the top of the column to collect the liquid droplets, which coalesce and fall back into the column.

A liquid distributor is designed to wet the packing bed evenly and initiate uniform contact between the liquid and vapor. The liquid distributor must spread the liquid uniformly, resist plugging and fouling, provide free space for gas flow, and allow operating flexibility.[9] Large towers frequently have a liquid redistributor to collect liquid off the column wall and direct it toward the center of the column for redistribution and enhanced contact in the lower section of packing.[4] Liquid redistributors are generally required for every 8 to 20 feet of random packing depth.[5, 10]

Distributors fall into two categories: gravitational types, such as orifice and weir types, and pressure-drop types, such as spray nozzles and perforated pipes. Spray nozzles are the most common distributors, but they may produce a fine mist that is easily entrained in the gas flow. They also may plug, and usually require high feed rates to compensate for poor distribution. Orifice-type distributors typically consist of flat trays with a number of risers for vapor flow and perforations in the tray floor for liquid flow. The trays themselves may present a resistance to gas flow.[9] However, better contact is generally achieved when orifice distributors are used.[3]

Packing materials provide a large wetted surface for the gas stream maximizing the area available for mass transfer. Packing materials are available in a variety of forms, each having specific characteristics with respect to surface area, pressure drop, weight, corrosion resistance, and cost. Packing life varies depending on the application. In ideal circumstances, packing will last as long as the tower itself. In adverse environments packing life may be as short as 1 to 5 years due to corrosion, fouling, and breakage.[11]

Packing materials are categorized as random or structured. Random packings are usually dumped into an absorption column and allowed to settle. Modern random packings consist of engineered shapes intended to maximize surface-to-volume ratio and minimize pressure drop.[2] Examples of different random packings are presented in Figure 1.2. The first random packings specifically designed for absorption towers were made of ceramic. The use of ceramic has declined because of their brittleness, and the current markets are dominated by metal and plastic. Metal packings cannot be used for highly corrosive pollutants, such as acid gas, and plastic packings are not suitable for high temperature applications. Both plastic and metal packings are generally limited to an unsupported depth of 20 to 25. At higher depths the weight may deform the packing.[10]

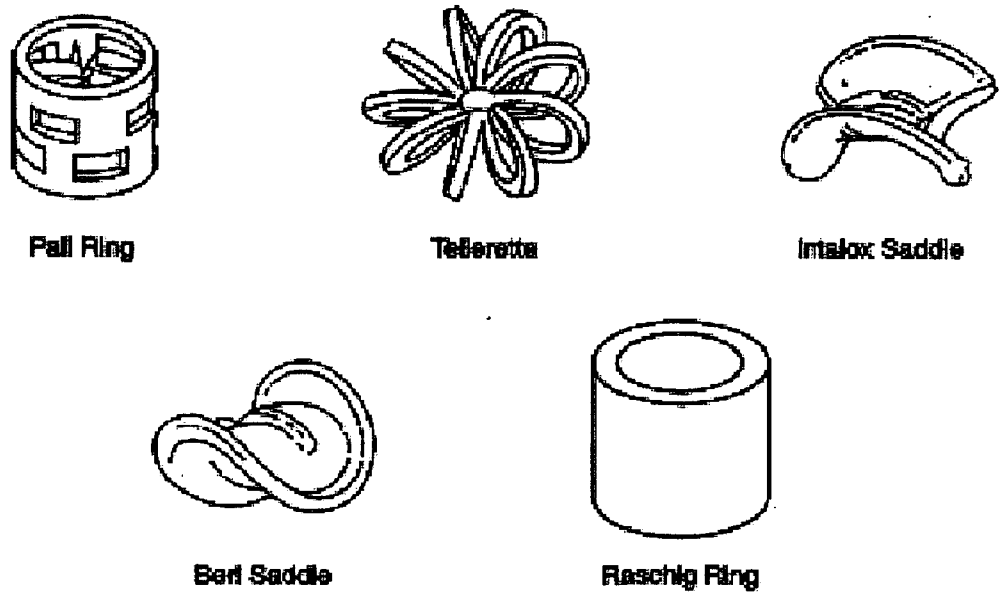


Figure 1.2: Random Packing Material

Structured packing may be random packings connected in an orderly arrangement, interlocking grids, or knitted or woven wire screen shaped into cylinders or gauze like arrangements. They usually have smaller pressure drops and are able to handle greater solvent flow rates than random packings.[4] However, structured packings are more costly to install and may not be practical for smaller columns. Most structured packings are made from metal or plastic.

In order to ensure that the waste gas is well distributed, an open space between the bottom of the tower and the packing is necessary. Support plates hold the packing above the open space. The support plates must have enough strength to carry the weight of the packing, and enough free area to allow solvent and gas to flow with minimum restrictions.[4]

High gas velocities can fluidize packing on top of a bed. The packing could then be carried into the distributor, become unlevel, or be damaged.[9] A packing restrainer may be installed at the top of the packed bed to contain the packing. The packing restrainer may be secured to the wall so that column upsets will not dislocate it, or a "floating" unattached weighted plate may be placed on top of the packing so that it can settle with the bed. The latter is often used for fragile ceramic packing.

1.2.4 Packed Tower Operation

As discussed in Section 1.2.1, the most common packed tower designs are countercurrent. As the waste gas flows up the packed column it will experience a drop in its pressure as it meets resistance from the packing materials and the solvent flowing down. Pressure drop in a column is a function of the gas and liquid flow rates and properties of the packing elements, such as surface area and free volume in the tower. A high pressure drop results in high fan power to drive the gas through the packed tower. and consequently high costs. The pressure drop in a packed tower generally ranges from 0.5 to 1.0 in. H_2O/ft of packing.[7]

For each column, there are upper and lower limits to solvent and vapor flow rates that ensure satisfactory performance. The gas flow rate may become so high that the drag on the solvent is sufficient to keep the solvent from flowing freely down the column. Solvent begins toaccumulate and blocks the entire cross section for flow, which increases the pressure drop and present the packing from mixing the gas and solvent effectively. When all the free volume in the packing is filled with liquid and the liquid is carried back up the column, the absorber is considered to be flooded.[4] Most packed towers operate at 60 to 70 percent of the gas flooding velocity, as it is not practical to operate a tower in a flooded condition.[7] A minimum liquid flow rate is also required to wet the packing material sufficiently for effective mass transfer to occur between the gas and liquid.[7]

The waste gas inlet temperature is another important scrubbing parameter. In general, the higher the gas temperature, the lower the absorption rate, and vice-versa. Excessively high gas temperatures also can lead to significant solvent loss through evaporation. Consequently, precoolers (e.g., spray chambers) may be needed to reduce the air temperature to acceptable levels.[6]

For operations that are based on chemical reaction with absorption, an additional concern is the rate of reaction between the solvent and pollutant(s). Most gas absorption chemical reactions are relatively fast and the rate limiting step is the physical absorption of the pollutants into the solvent. However, for solvent-pollutant systems where the chemical reaction is the limiting step, the rates of reaction would need to be analyzed kinetically.

Heat may be generated as a result of exothermal chemical reactions. Heat may also be generated when large amounts of solute are absorbed into the liquid phase, due to the heat of solution. The resulting change in temperature along the height of the absorber column may damage equipment and reduce absorption efficiency. This problem can be avoided by adding cooling coils to the column.[7] However, in those systems where water is the solvent, adiabatic saturation of the gas occurs during absorption due to solvent evaporation. This causes a substantial cooling of the absorber that offsets the heat generated by chemical reactions. Thus, cooling coils are rarely required with those systems.[5] In any event, packed towers may be designed assuming that isothermal conditions exist throughout the column.[7]

The effluent from the column may be recycled into the system and used again. This is usually the case if the solvent is costly, i.e., hydrocarbon oils, caustic solution. Initially, the recycle stream may go to a waste treatment system to remove the pollutants or the reaction product. Make-up solvent may then be added before the liquid stream reenters the column. Recirculation of the solvent requires a pump, solvent recovery system, solvent holding and mixing tanks, and any associated piping and instrumentation.

1.3 Design Procedures

The design of packed tower absorbers for controlling gas streams containing a mixture of pollutants and air depends on knowledge of the following parameters:

- Waste gas flow rate;
- Waste gas composition and concentration of the pollutants in the gas stream;
- Required removal efficiency;
- Equilibrium relationship between the pollutants and solvent; and
- Properties of the pollutant(s), waste gas, and solvent: diffusivity, viscosity, density, and molecular weight.

The primary objectives of the design procedures are to determine column surface area and pressure drop through the column. In order to determine these parameters, the following steps must be performed:

- Determine the gas and liquid stream conditions entering and exiting the column.
- Determine the absorption factor (AF).
- Determine the diameter of the column (D).
- Determine the tower height (H_{tower}) and surface area (S).
- Determine the packed column pressure drop (P).

To simplify the sizing procedures, a number of assumptions have been made. For example, the waste gas is assumed to comprise a two-component waste gas mixture (pollutant/air), where the pollutant consists of a single compound present in dilute quantities. The waste gas is assumed to behave as an ideal gas and the solvent is assumed to behave as an ideal solution. Heat effects associated with absorption are considered to be minimal for the pollutant concentrations

encountered. The procedures also assume that, in chemical absorption, the process is not reaction rate limited, i.e., the reaction of the pollutant with the solvent is considered fast compared to the rate of absorption of the pollutant into the solvent.

The design procedures presented here are complicated, and careful attention to units is required. Appendix A has a list of all design variables referred to in this chapter, along with the appropriate units.

1.3.1 Determining Gas and Liquid Stream Conditions

Gas absorbers are designed based on the ratio of liquid to gas entering the column (L_i/G_i), slope of the equilibrium curve (m), and the desired removal efficiency (η). These factors are calculated from the inlet and outlet gas and liquid stream variables:

- Waste gas flow rate, in actual cubic feet per minute (acfm), entering and exiting column (G_i and G_o, respectively);
- Pollutant concentration (lb-moles pollutant per lb-mole of pollutant free gas) entering and exiting the column in the waste gas (Y_i and Y_o, respectively);
- Solvent flow rate, in gallons per minute (gpm), entering and exiting the column (L_i and L_o, respectively); and
- Pollutant concentration (lb-moles pollutant per lb-mole of pollutant free solvent) entering and exiting the column in the solvent (X_i and X_o, respectively).

This design approach assumes that the inlet gas stream variables are known, and that a specific pollutant removal efficiency has been chosen as the design basis; i.e., the variables G_i, Y_i and η are known. For dilute concentrations typically encountered in pollution control applications and negligible changes in moisture content, G_i is assumed equal to G_o. If a once-through process is used, or if the spent solvent is regenerated by an air stripping process before it is recycled, the value of X_i will approach zero. The following procedures must be followed to calculate the remaining stream variables Y_o, L_i (and L_o), and X_o. A schematic diagram of a packed tower with inlet and outlet flow and concentration variables labeled is presented in Figure 1.3.

The exit pollution concentration, Y_o, may be calculated from using the following equation:

$$Y_o = Y_i \left(1 - \frac{\eta}{100}\right) \tag{1.1}$$

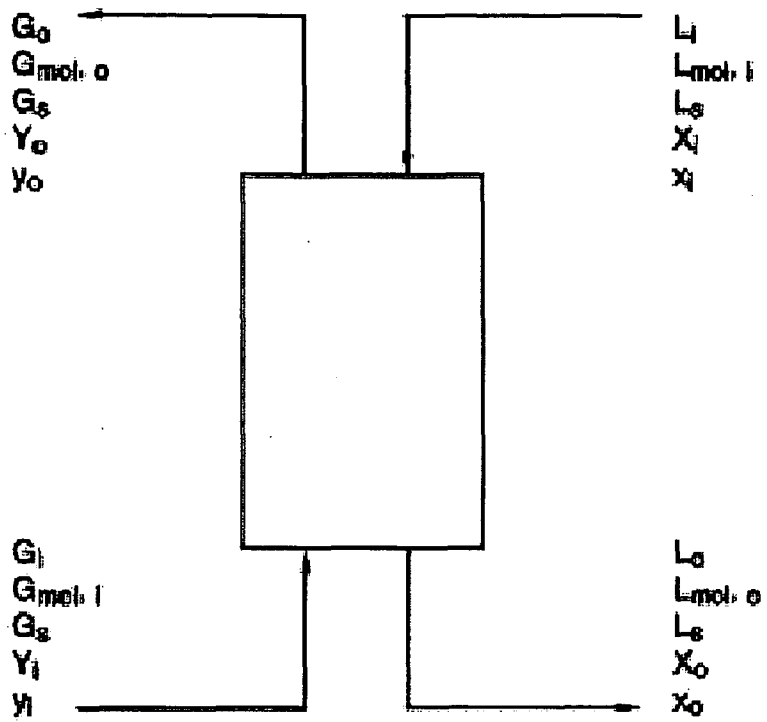


Figure 1.3: Schematic Diagram of Countercurrent Packed Bed Operation

The liquid flow rate entering the absorber, L_i (gpm), is then calculated using a graphical method. Figure 1.4 presents an example of an equilibrium curve and operating line. The equilibrium curve indicates the relationship between the concentration of pollutant in the waste gas and the concentration of pollutant in the solvent at a specified temperature. The operating line indicates the relation between the concentration of the pollutant in the gas and solvent at any location in the gas absorber column. The vertical distance between the operating line and equilibrium curve indicates the driving force for diffusion of the pollutant between the gas and liquid phases. The minimum amount of liquid which can be used to absorb the pollutant in the gas stream corresponds to an operating line drawn from the outlet concentration in the gas stream (Y_o) and the inlet concentration in the solvent stream (X_i) to the point on the equilibrium curve corresponding to the entering pollutant concentration in the gas stream (Y_i). At the intersection point on the equilibrium curve, the diffusional driving forces are zero, the required time of contact for the concentration change is infinite, and an infinitely tall tower results.

The slope of the operating line intersecting the equilibrium curve is equal to the minimum L/G ratio on a moles of pollutant-free solvent (L_s) per moles of pollutant-free gas basis G_s. in other words, the values L_s and G_s do not include the moles of pollutant in the liquid and gas streams. The values of L_s and G_s are constant through the column if a negligible amount of moisture is transferred

from the liquid to the gas phase. The slope may be calculated from the following equation:

$$\left(\frac{L_s}{G_s}\right)_{min} = \frac{Y_i - Y_o}{X_o^* - X_i} \tag{1.2}$$

where X^*_o would be the maximum concentration of the pollutant in the liquid phase if it were allowed to come to equilibrium with the pollutant entering the column in the gas phase, Y_i. The value of X^*_o is taken from the equilibrium curve. Because the minimum L_s/G_s, ratio is an unrealistic value, it must be multiplied by an adjustment factor, commonly between 1.2 and 1.5, to calculate the actual L/G ratio:[7]

$$\left(\frac{L_s}{G_s}\right)_{act} = \left(\frac{L_s}{G_s}\right)_{min} \times \text{(adjustment factor)} \tag{1.3}$$

The variable G_s may be calculated using the equation:

$$G_s = \frac{60\ \rho_G\ G_i}{MW_G\ (1 + Y_i)} \tag{1.4}$$

where 60 is the conversion factor from minutes to hours, MW_G is the molecular weight of the gas stream (lb/lb-mole), and ρ_G is the density of the gas stream (lb/ft^3). For pollutant concentrations typically encountered, the molecular weight and density of the waste gas stream are assumed to be equal to that of ambient air.

The variable L_s may then be calculated by:

$$L_s = \left(\frac{L_s}{G_s}\right)_{act} \times G_s \tag{1.5}$$

The total molar flow rates of the gas and liquid entering the absorber ($G_{mol,i}$ and $L_{mol,i}$) are calculated using the following equations:

$$G_{mol,i} = G_s\ (1 + Y_i) \tag{1.6}$$

$$L_{mol,i} = L_s\ (1 + X_i) \tag{1.7}$$

The volume flow rate of the solvent, L_i, may then be calculated by using the following relationship:

$$L_i = \frac{7.48\ L_{mol,i}\ MW_L}{60\ r_L} \tag{1.8}$$

where 60 is the conversion factor from minutes to hours, MW_L, is the molecular weight of the liquid stream (lb/lb-mole), ρ_L is the density of the liquid stream (lb/ft^3), and 7.48 is the factor used to convert cubic feet to gallons. If the volume change in the liquid stream entering and exiting the absorber is assumed to be negligible, then $L_i = L_o$.

Gas absorber vendors have provided a range for the L_i/G_i ratio for acid gas control from 2 to 20 gpm of solvent per 1000 cfm of waste gas.[12] Even for pollutants that are highly soluble in a solvent (i.e., HCl in water), the adjusted L_i/G_i ratio calculated using Equations 1.2 to 1.8 would be much lower than this range, because these equations do not consider the flow rate of the solvent required to wet the packing.

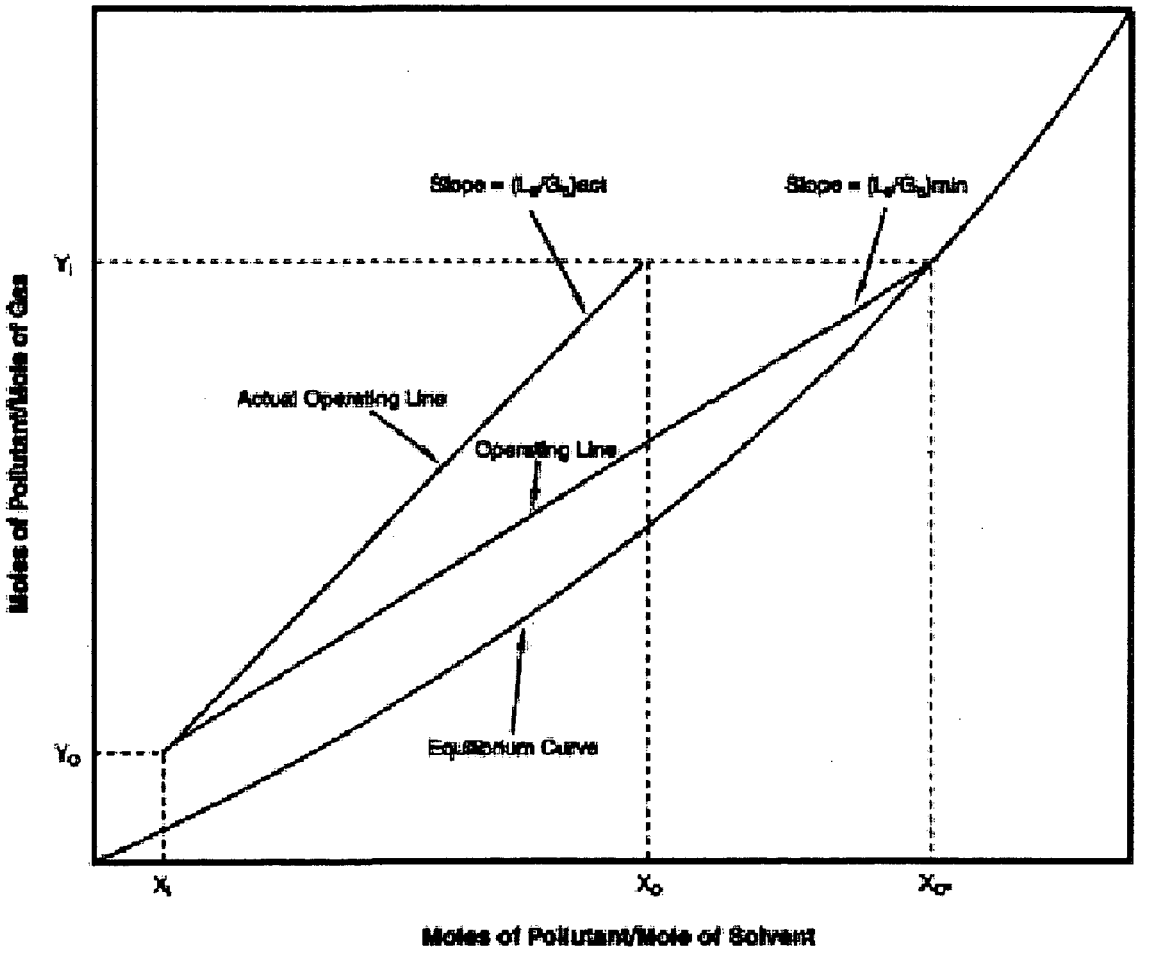


Figure 1.4: Minimum and Actual Liquid-to-Gas Ratio

Finally, the actual operating line may be represented by a material balance equation over the gas absorber:[4]

$$X_i L_s + Y_i G_s = X_o L_s + Y_o G_s \tag{1.9}$$

Equation 1.9 may then be solved for X_o:

$$X_o = \frac{Y_i - Y_o}{\left(\frac{L_s}{G_s}\right)} + X_i \tag{1.10}$$

1.3.2 Determining Absorption Factor

The absorption factor (AF) value is frequently used to describe the relationship between the equilibrium line and the liquid-to-gas ratio. For many pollutant-solvent systems, the most economical value for AF ranges around 1.5 to 2.0.[7] The following equation may be used to calculate AF:[4, 7]

$$AF = \frac{L_{mol,\,i}}{m\ G_{mol,\,i}} \tag{1.11}$$

where m is the slope of the equilibrium line on a mole fraction basis. The value of m may be obtained from available literature on vapor/liquid equilibrium data for specific systems. Since the equilibrium curve is typically linear in the concentration ranges usually encountered in air pollution control, the slope, m would be constant (or nearly so) for all applicable inlet and outlet liquid and gas streams. The slope may be calculated from mole fraction values using the following equation:[4]

$$m = \frac{y_o^* - y_i^*}{x_o - x_i} \tag{1.12}$$

where y_i^* and y_o^* are the mole fractions of the pollutant in the vapor phase in equilibrium with the mole fractions of the pollutant entering and exiting the absorber in the liquid, x_i and x_o, respectively. The slope of the equilibrium line in Figure 1.4 is expressed in terms of concentration values X_i, X_o, Y_i^*, and Y_o^*. These values may be converted to x_i, x_o, y_i^*, and y_o^* using the equations:

$$x_i = \frac{X_i}{1 + X_i} \tag{1.13}$$

$$x_o = \frac{X_o}{1 + X_o} \tag{1.14}$$

$$y_i^* = \frac{Y_i^*}{1 + Y_i^*} \tag{1.15}$$

$$y_o^* = \frac{Y_o^*}{1 + Y_o^*} \tag{1.16}$$

where the units for each of these variables are listed in Appendix A.

The absorption factor will be used to calculate the theoretical number of transfer units and the theoretical height of a transfer unit. First, however, the column diameter needs to be determined.

1.3.3 Determining Column Diameter

Once stream conditions have been determined, the diameter of the column may be estimated. The design presented in this section is based on selecting a fraction of the gas flow rate at flooding conditions. Alternatively, the column may be designed for a specific pressure drop (see Section 1.3.6.). Eckert's modification to the generalized correlation for randomly packed towers based on flooding considerations is used to obtain the superficial gas flow rate entering the absorber, $G_{sfr,i}$ (lb/sec-ft^2), or the gas flow rate per crossectional area based on the $L_{mol,i}/G_{mol,i}$ ratio calculated in Section 1.3.2.[10] The cross-sectional area (A) of the column and the column diameter (D) can then be determined from $G_{sfr,i}$. Figure 1.5 presents the relationship between $G_{sfr,i}$ and the $L_{mol,i}/G_{mol,i}$ ratio at the tower flood point. The Abscissa value (X axis) in the graph is expressed as:[10]

$$Abscissa = \left(\frac{L_{mol,i}}{G_{mol,i}}\right)\left(\frac{MW_L}{MW_G}\right)\sqrt{\frac{\rho_G}{\rho_L}} \tag{1.17}$$

The Ordinate value (Y axis) in the graph is expressed as:[10]

$$Ordinate = \frac{\left(G_{sfr,i}\right)^2 \Psi F_p \left(\frac{\mu_L}{2.42}\right)^{0.2}}{\rho_L \rho_G g_c} \tag{1.18}$$

where F_p is a packing factor, g_c is the gravitational constant (32.2), μ_L is the viscosity of the solvent (lb/ft-hr), 2.42 is the factor used to convert lb/ft-hr to centipoise, and Ψ is the ratio of the density of the scrubbing liquid to water. The value of F_p may be obtained from packing vendors (see Appendix B, Table 1.8).

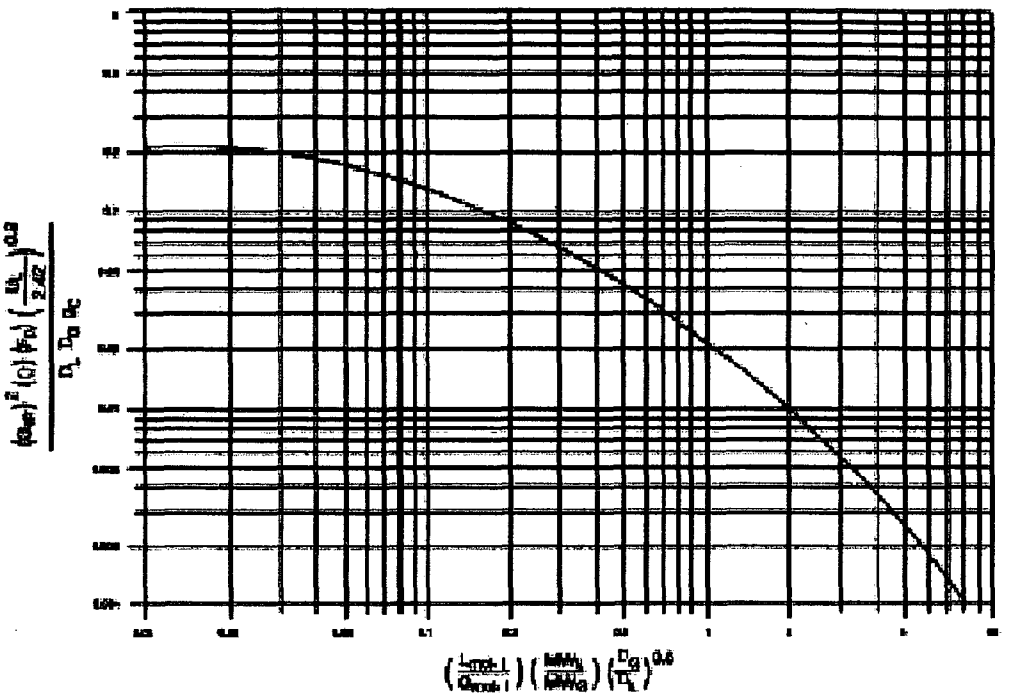


Figure 1.5: Eckert's Modification to the Generalized Correlation at Flooding Rate

After calculating the *Abscissa* value, a corresponding *Ordinate* value may determined from the floo ding curve. The *Ordinate* may also be calculated using the following equation:[10]

$$Ordinate = 10^{\left[-1.668 - 1.085\ (\log Abscissa) - 0.297\ (\log Abscissa)^2\right]} \tag{1.19}$$

Equation 1.18 may then be rearranged to solve for $G_{sfr,i}$:

$$G_{sfr,\,i} = \frac{\rho_l\ \rho_G\ g_c\ (Ordinate)}{F_p\ \Psi \left(\frac{\mu_L}{2.42}\right)^{0.2}} \tag{1.20}$$

The cross-sectional area of the tower (ft^2) is calculated as:

$$A = \frac{G_{mol,\,i}\ MW_G}{3{,}600\ G_{sfr,i}\ f} \tag{1.21}$$

where f is the flooding factor and 3600 is the conversion factor from hours to seconds. To prevent flooding, the column is operated at a fraction of $G_{sfr,i}$. The value of f typically ranges from 0.60 to

0.75.[7]

The diameter of the column (ft) can be calculated from the cross-sectional area, by:

$$D = \sqrt{\frac{4}{\pi} A} \tag{1.22}$$

If a substantial change occurs between inlet and outlet volumes (i.e., moisture is transferred from the liquid phase to the gas phase), the diameter of the column will need to be calculated at the top and bottom of the column. The larger of the two values is then chosen as a conservative number. As a rule of thumb, the diameter of the column should be at least 15 times the size of the packing used in the column. If this is not the case, the column diameter should be recalculated using a smaller diameter packing.[10]

The superficial liquid flow rate entering the absorber, $L_{sfr,i}$ (lb/hr-ft^2 based on the cross-sectional area determined in Equation 1.21 is calculated from the equation:

$$L_{sfr,i} = \frac{L_{mol,i}\, MW_L}{A} \tag{1.23}$$

For the absorber to operate properly, the liquid flow rate entering the column must be high enough to effectively wet the packing so mass transfer between the gas and liquid can occur. The minimum value of $L_{sfr,i}$ that is required to wet the packing effectively can be calculated using the equation:[7, 13]

$$\left(L_{sfr,i}\right)_{min} = MWR\ \rho_L\ a \tag{1.24}$$

where MWR is defined as the minimum wetting rate (ft^2/hr), and a is the surface area to volume ratio of the packing (ft^2/ft^3). An MWR value of 0.85 ft^2/hr is recommended for ring packings larger than 3 inches and for structured grid packings. For other packings, an MWR of 1.3 ft^2/hr is recommended.[7,13] Appendix B, Table 1.8 contains values of a for common packing materials.

If $L_{sfr,i}$ (the value calculated in Equation 1.23) is smaller than $(L_{sfr,})_{min}$ (the value calculated in Equation 1.24), there is insufficient liquid flow to wet the packing using the current design parameters. The value of $G_{sfr,i}$, and A then will need to be recalculated. See Appendix C for details.

1.3.4 Determining Tower Height and Surface Area

Tower height is primarily a function of packing depth. The required depth of packing (H_{pack}) is determined from the theoretical number of overall transfer units (N_{tu}) needed to achieve a specific removal efficiency, and the height of the overall transfer unit (H_{tu}):[4]

$$H_{pack} = N_{tu} H_{tu} \tag{1.25}$$

The number of overall transfer units may be estimated graphically by stepping off stages on the equilibrium-operating line graph from inlet conditions to outlet conditions, or by the following equation:[4]

$$N_{tu} = \frac{\ln\left[\left(\frac{y_i - mx_i}{y_o - mx_i}\right)\left(1 - \frac{1}{AF}\right) + \frac{1}{AF}\right]}{1 - \frac{1}{AF}} \tag{1.26}$$

where ln is the natural logarithm of the quantity indicated.

The equation is based on several assumptions: 1) Henry's law applies for a dilute gas mixture; 2) the equilibrium curve is linear from x_i to x_o; and 3) the pollutant concentration in the solvent is dilute enough such that the operating line can be considered a straight line.[4]

If $x_i \approx 0$ (i.e., a negligible amount of pollutant enters the absorber in the liquid stream) and $1/A \approx 0$ (i.e., the slope of the equilibrium line is very small and/or the L_{mol}/G_{mol} ratio is very large), Equation 1.26 simplifies to:

$$N_{tu} = \ln\left(\frac{y_i}{y_o}\right) \tag{1.27}$$

There are several methods that may be used to calculate the height of the overall transfer unit, all based on empirically determined packing constants. One commonly used method involves determining the overall gas and liquid mass transfer coefficients (K_G, K_L). A major difficulty in using this approach is that values for K_G and K_L are frequently unavailable for the specific pollutant-solvent systems of interest. The reader is referred to the book Random Packing and Packed Tower Design Applications in the reference section for further details regarding this method.[14]

For this chapter, the method used to calculate the height of the overall transfer unit is based on estimating the height of the gas and liquid film transfer units, H_L and H_G, respectively:[4]

$$H_{tu} = H_G + \frac{1}{AF} H_L \tag{1.28}$$

The following correlations may be used to estimate values for H_L and H_G:[13]

$$H_G = \left[\alpha \frac{\left(3{,}600\, f G_{sfr,i}\right)^{\beta}}{\left(L_{sfr,i}\right)^{\Gamma}}\right] \sqrt{\frac{\mu_G}{\rho_G D_G}} \tag{1.29}$$

$$H_L = \phi\left(\frac{L_{sfr,i}}{\mu_L}\right)^b \sqrt{\frac{\mu_L}{\rho_L D_L}} \qquad (1.30)$$

The quantity $\mu / \rho D$ is the Schmidt number and the variables β, b and Γ are packing constants specific to each packing type. Typical values for these constants are listed in Appendix B, Tables 1.9 and 1.10. The advantage to using this estimation method is that the packing constants may be applied to any pollutant-solvent system. One packing vendor offers the following modifications to Equations 1.29 and 1.30 for their specific packing:[15]

$$H_G = \left[\alpha \frac{\left(3{,}600\, f G_{sfr,i}\right)^\beta}{\left(L_{sfr,i}\right)^\Gamma}\right] \sqrt{\frac{\mu_G}{\rho_G D_G}} \left(\frac{\mu_L^\Gamma}{\mu_G^\beta}\right) \qquad (1.31)$$

$$H_L = \phi\left(\frac{L_{sfr,i}}{\mu_L}\right)^b \sqrt{\frac{\mu_L}{\rho_L D_L}} \left(\frac{T}{286}\right)^{-4.255} \qquad (1.32)$$

where T is the temperature of the solvent in Kelvin.

After solving for H_{pack} using Equation 1.25, the total height of the column may be calculated from the following correlation:[16]

$$H_{tower} = 1.40\, H_{pack} + 1.02\, D + 2.81 \qquad (1.33)$$

Equation 1.33 was developed from information reported by gas absorber vendors, and is applicable for column diameters from 2 to 12 feet and packing depths from 4 to 12 feet. The surface area (S) of the gas absorber can be calculated using the equation:[16]

$$S = \pi D \left(H_{tower} + \frac{D}{2}\right) \qquad (1.34)$$

Equation 1.34 assumes the ends of the absorber are flat and circular.

1.3.5 Calculating Column Pressure Drop

Pressure drop in a gas absorber is a function of $G_{sfr,i}$ and properties of the packing used. The pressure drop in packed columns generally ranges from 0.5 to 1 inch of H_2O per foot of packing. The absorber may be designed for a specific pressure drop or pressure drop may be estimated using Leva's correlation:[7, 10]

$$\Delta P = c10\left(\frac{j L_{sfr,i}}{3{,}600}\right) \frac{\left(f G_{sfr,i}\right)^2}{\rho_G} \qquad (1.35)$$

The packing constants c and j are found in Appendix B, Table 1.11, and 3600 is the conversion factor from seconds to hours. The equation was originally developed for air-water systems. For other liquids, $L_{sfr,i}$ is multiplied by the ratio of the density of water to the density of the liquid.

1.3.6 Alternative Design Procedure

The diameter of a column can be designed for a specific pressure drop, rather than being determined based on a fraction of the flooding rate. Figure 1.6 presents a set of generalized correlations at various pressure drop design values. The *Abscissa* value of the graph is similar to Equation 1.17:[10]

$$Abscissa = \left(\frac{L_{mol,i}}{G_{mol,i}}\right)\left(\frac{MW_L}{MW_G}\right)\sqrt{\frac{\rho_G}{\rho_L - \rho_G}} \tag{1.36}$$

The *Ordinate* value is expressed as:[10]

$$Ordinate = \frac{\left(G_{sfr,i}\right)^2 F_p \left(\frac{\mu_L}{2.42}\right)^{0.1}}{\left(\rho_L - \rho_G\right)\rho_G g_c} \tag{1.37}$$

For a calculated *Abscissa* value, a corresponding *Ordinate* value at each pressure drop can be read off Figure 1.6 or can be calculated from the following equation:[10]

$$\begin{aligned} Ordinate = \exp[k_0 + k_1(\ln Abscissa) + k_2(\ln Abscissa)^2 \\ + k_3(\ln Abscissa)^3 + k_4(\ln Abscissa)^4] \end{aligned} \tag{1.38}$$

The constants k_0, k_1, k_2, k_3, and k_4, are shown below for each pressure drop value.

Table 1.1: Values of Constants k_0 through k_4 for Various Pressure Drops

Δ P (inches water/ ft packing)	k_0	k_1	k_2	k_3	k_4
0.05	-6.3205	-06080	-0.1193	-0.0068	0.0003
0.10	-5.5009	-0.7851	-0.1350	0.0013	0.0017
0.25	-5.0032	-0.9530	-0.1393	0.0126	0.0033
0.50	-4.3992	-0.9940	-0.1698	0.0087	0.0034
1.00	-4.0950	-1.0012	-0.1587	0.0080	0.0032
1.50	-4.0256	-0.9895	-0.0830	0.0324	0.0053

Equation 1.37 can be solved for $G_{sfr,i}$.

$$G_{sfr,i} = \sqrt{\frac{(\rho_L - \rho_G)\,\rho_G\, g_c(Ordinate)}{F_p \left(\frac{\mu_L}{2.42}\right)^{0.1}}} \tag{1.39}$$

The remaining calculations to estimate the column diameter and $L_{sfr,i}$ are the same as presented in Section 1.3.3, except the flooding factor (f) is not used in the equations. The flooding factor is not required because an allowable pressure drop that will not cause flooding is chosen to calculate the diameter rather than designing the diameter at flooding conditions and then taking a fraction of that value.

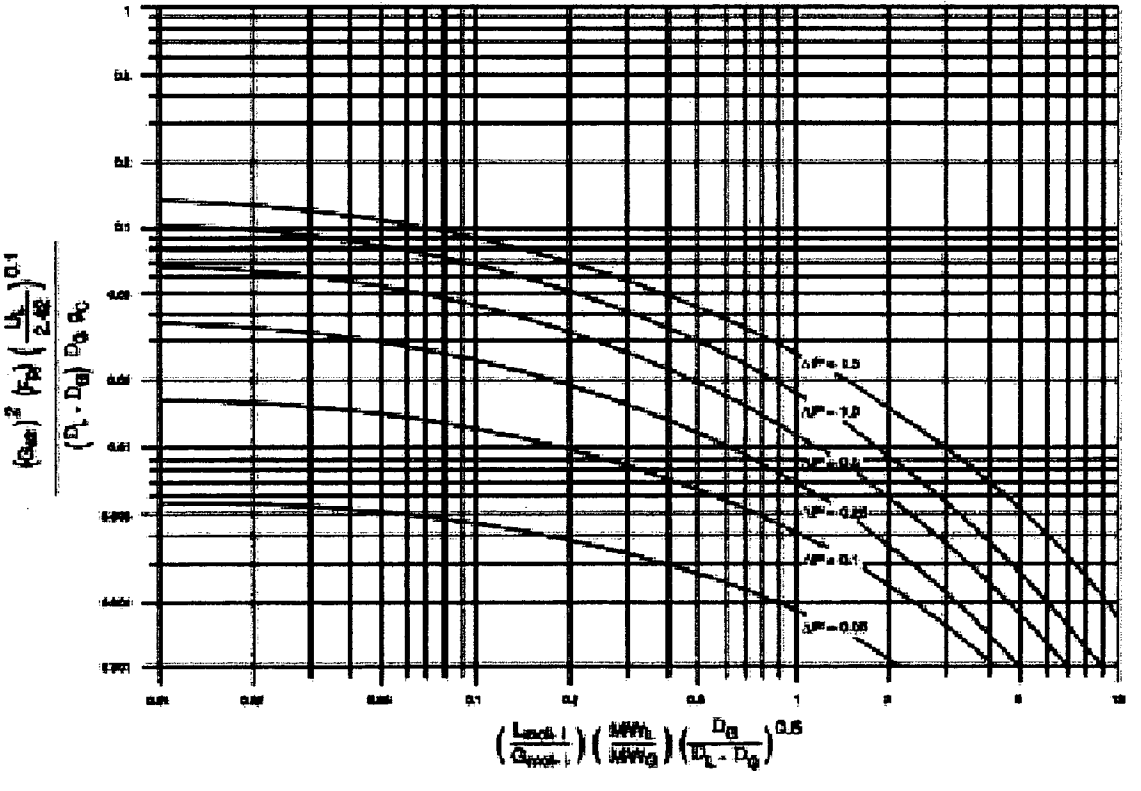


Figure 1.6: Generalized Pressure Drop Correlations [10]

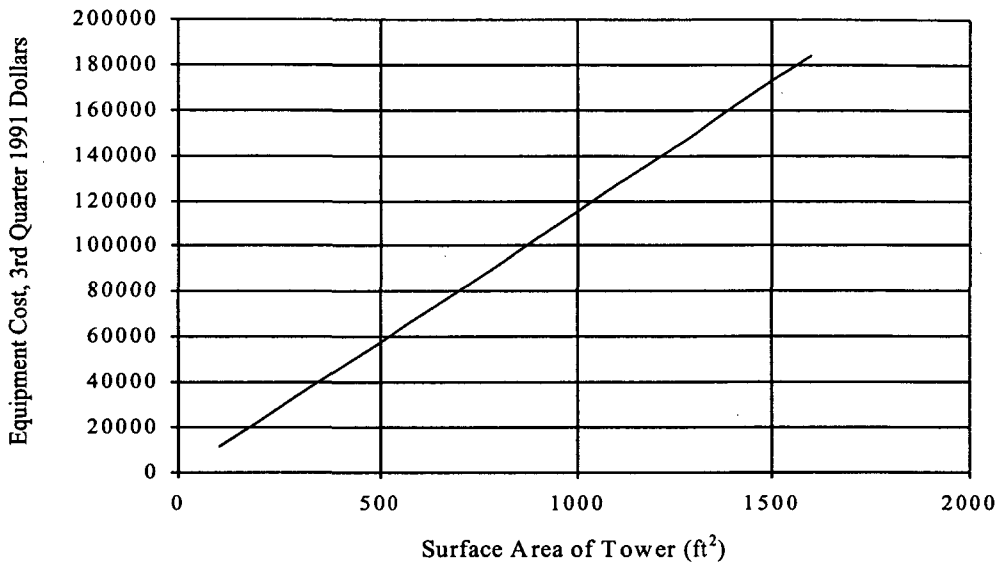


Figure 1.7: Packed Tower Equipment Cost [16]

1.4 Estimating Total Capital Investment

This section presents the procedures and data necessary for estimating capital costs for vertical packed bed gas absorbers using countercurrent flow to remove gaseous pollutants from waste gas streams. Equipment costs for packed bed absorbers are presented in Section 1.4.1, with installation costs presented in Section 1.4.2.

Total capital investment, *TCI*, includes equipment cost, *EC*, for the entire gas absorber unit, taxes, freight charges, instrumentation, and direct and indirect installation costs. All costs are presented in **third quarter 1991 dollars**[1]. The costs presented are study estimates with an expected accuracy of ± 30 percent. It must be kept in mind that even for a given application, design and manufacturing procedures vary from vendor to vendor, so costs vary. All costs are for new plant installations; no retrofit cost considerations are included.

1.4.1 Equipment Costs for Packed Towers

Gas absorber vendors were asked to supply cost estimates for a range of tower dimensions (i.e., height, diameter) to account for the varying needs of different applications. The equipment for which they were asked to provide costs consisted of a packed tower absorber made of fiberglass reinforced plastic (FRP), and to include the following equipment components:

- absorption column shell;
- gas inlet and outlet ports;
- liquid inlet port and outlet port/drain;
- liquid distributor and redistributor;
- two packing support plates;
- mist eliminator;
- internal piping;
- sump space; and
- platforms and ladders.

The cost data the vendors supplied were first adjusted to put them on a common basis, and then were regressed against the absorber surface area (S). The equation shown below is a linear regression of cost data provided by six vendors.[16, 12]

$$Total\ Tower\ Cost\ (\$) = 115\ S \qquad (1.40)$$

where S is the surface area of the absorber, in ft^2. Figure 1.7 depicts a plot of Equation 1.40. This equation is applicable for towers with surface areas from 69 to 1507 ft^2 constructed of FRP. Costs for towers made of materials other than FRP may be estimated using the following equation:

$$TTC_M = CF \times TTC \qquad (1.41)$$

where TTC_M is the total cost of the tower using other materials, and TTC is the total tower cost as estimated using Equation 1.40. The variable CF is a cost factor to convert the cost of an FRP gas

Table 1.2: Random Packing Costs[a]

Nominal Diameter (inches)	Construction Material	Packing Type	Packing cost ($/ft³) <100 ft³	>100 ft³
1	304 stainless steel	Pall rings, Rasching rings, Ballast rings	70-109	65-99
1	Ceramic	Rasching rings, Berl saddles	33-44	26-36
1	Polypropylene	Tri-Pak®, Pall rings, Ballast rings, Flexisaddles, Berl saddles, Rasching rings	141-37	12-34
2	Ceramic	Tri-Pac®, Lanpac®, Flexiring, Flexisaddle	13-32	10-30
2	Polypropylene	Tellerette , Ballast rings	3-20	5-19
3.5	304 stainless steel	Tri-pack®, Lanpac®, Ballast rings 30	27	
3.5	Polypropylene		6-14	6-12

[a] Provided by packing vendors. [17]

® Denotes registered trademark.

absorber to an absorber fabricated from another material. Ranges of cost factors provided by vendors are listed for the following materials of construction:[12]

304 Stainless steel:	1.10 - 1.75
Polypropylene:	0.80 - 1.10
Polyvinyl chloride:	0.50 - 0.90

Auxiliary costs encompass the cost of all necessary equipment not included in the absorption column unit. Auxiliary equipment includes packing material, instruments and controls, pumps, and fans. Cost ranges for various types of random packings are presented in Table 1.2. The cost of structured packings varies over a much wider range. Structured packings made of stainless steel range from $\$45/ft^3$ to $\$405/ft^3$, and those made of polypropylene range from $\$65/ft^3$ to $\$350/ft^3$.[17]

Similarly, the cost of instruments and controls varies widely depending on the complexity required. Gas absorber vendors have provided estimates ranging from $1,000 to $10,000 per column. A factor of 10 percent of the *EC* will be used to estimate this cost in this chapter. (see eq. 1.42, below.) Design and cost correlations for fans and pumps will be presented in a chapter on auxiliary equipment elsewhere in this manual. However, cost data for auxiliaries are available from the literature (see reference [18], for example).

The total equipment cost (EC) is the sum of the component equipment costs, which includes tower cost and the auxiliary equipment cost.

$$EC = TTC + Packing\ Cost + Auxiliary\ Equipment \tag{1.42}$$

The purchased equipment cost (PEC) includes the cost of the absorber with packing and its auxiliaries (*EC*), instrumentation (0.10 *EC*), sales tax (0.03 *EC*), and freight (0.05 *EC*). The PEC is calculated from the following factors, presented in Section 1 of this manual and confirmed from the gas absorber vendor survey conducted during this study:[12, 19],

$$PEC = (1+ 0.10+0.03+0.05)EC = 1.18\ EC \tag{1.43}$$

1.4.2 Installation Costs

The total capital investment, *TCI*, is obtained by multiplying the purchased equipment cost, *PEC*, by the total installation factor:

$$TCI = 2.20\ PEC \tag{1.44}$$

The factors which are included in the total installation factor are also listed in Table 1.3.[19] The factors presented in Table 1.3 were confirmed from the gas absorber vendor survey.

1.5 Estimating Annual Cost

The total annual cost (TAC) is the sum of the direct and indirect annual costs.

1.5.1 Direct Annual Costs

Direct annual costs (DC) are those expenditures related to operating the equipment, such as labor and materials. The suggested factors for each of these costs are shown in Table 1.4. These factors were taken from Section 1 of this manual and were confirmed from the gas absorber vendor survey. The annual cost for each item is calculated by multiplying the number of units used annually (i.e., hours, pounds, gallons, kWh) by the associated unit cost.

Operating labor is estimated at ½-hour per 8-hour shift. The supervisory labor cost is estimated at 15 percent of the operating labor cost. Maintenance labor is estimated at 1/2-hour per 8-hour shift. Maintenance materials costs are assumed to equal maintenance labor costs.

Solvent costs are dependent on the total liquid throughput, the type of solvent required, and the fraction of throughput wasted (often referred to as blow-down). Typically, the fraction of solvent wasted varies from 0.1 percent to 10 percent of tire total solvent throughput.[12] For acid gas systems, the amount of solvent wasted is determined by the solids content, with bleed off occurring when solids content reaches 10 to 15 percent to prevent salt carry-over.[12]

The total annual cost of solvent (C_s) is given by:

$$C_s = L_i \, WF \left(60 \, \frac{min}{hr}\right) \begin{pmatrix} annual \\ operating \\ hours \end{pmatrix} \begin{pmatrix} solvent \\ unit \, cost \end{pmatrix} \tag{1.45}$$

where WF is the waste (make-up) fraction, and the solvent unit cost is expressed in terms of $/gal.

The cost of chemical replacement (C_c) is based on the annual consumption of the chemical and can be calculated by:

$$C_c = \left(\frac{lbs \, chemical \, used}{hr}\right) \begin{pmatrix} annual \\ operating \\ hours \end{pmatrix} \begin{pmatrix} chemical \\ unit \, cost \end{pmatrix} \tag{1.46}$$

where the chemical unit cost is in terms of $/lb.

Table 1.3: Capital Cost Factors for Gas Absorbers [19]

Cost Item	Factor
Direct Costs	
Purchased equipment costs	
Absorber + packing + auxiliary equipment [a], EC	As estimated, A
Instrumentation [b]	0.10 A
Sales taxes	0.03 A
Freight	0.05 A
Purchased equipment cost, PEC	B = 1.18 A
Direct installation costs	
Foundations & supports	0.12 B
Handling & erection	0.40 B
Electrical	0.01 B
Piping	0.30 B
Insulation	0.01 B
Painting	0.01 B
Direct installation costs	0.85 B
Site preparation	As required, SP
Buildings	As required, Bldg.
Total Direct Costs, DC	1.85 B + SP + Bldg.
Indirect Costs (installation)	
Engineering	0.10 B
Construction and field expenses	0.10 B
Contractor fees	0.10 B
Start-up	0.01 B
Performance test	0.01 B
Contingencies	0.03 B
Total Indirect Costs, IC	0.35 B
Total Capital Investment = DC + IC	2.20 B + SP + Bldg.

[a] Includes the initial quantity of packing, as well as items normally not included with the unit supplied by vendors, such as ductwork, fan, piping, etc.
[b] Instrumentation costs cover pH monitor and liquid level indicator in sump.

Table 1.4: Suggested Annual Cost Factors for Gas Absorber Systems

Cost Item	Factor
Direct Annual Costs, DC	
Operating labor [a]	
Operator	1/2 hour per shift
Supervisor	15% of operator
Operating materials [b]	Application specific
Solvent	(throughput/yr) x (waste fraction)
Chemicals	Based on annual consumption
Wastewater disposal	(throughput/yr) x (waste fraction)
Maintenance [a]	
Labor	1/2 hour per shift
Material	100% of maintenance labor
Electricity	(consumption rate) x (hours/yr) x (unit cost)
Fan	
Pump	
Indirect Annual Costs, IC	
Overhead	60% of total labor and material costs
Administrative charges	2% of Total Capital Investment
Property tax	1% of Total Capital Investment
Insurance	1% of Total Capital Investment
Capital recovery [c]	0.1098 x Total Capital Investment
Total Annual Cost	DC + IC

[a] These factors were confirmed by vendor contacts.
[b] If system does not use chemicals (e.g., caustic), this quantity is equal to annual solvent consumption.
[c] Assuming a 15-year life at 7%. See Chapter 2

Solvent disposal (C_{ww}) costs vary depending on geographic location. type of waste disposed of, and availability of on-site treatment. Solvent disposal costs are calculated by:

$$C_{ww} = L_i\ WF \left(60\ \frac{min}{hr}\right) \begin{pmatrix} annual \\ operating \\ hours \end{pmatrix} \begin{pmatrix} solvent \\ disposal\ cost \end{pmatrix} \qquad (1.47)$$

where the solvent disposal costs are in terms of $/gal of waste solvent.

The electricity costs associated with operating a gas absorber derive from fan requirements to overcome the pressure drop in the column, ductwork, and other parts of the control system, and pump requirements to recirculate the solvent. The energy required for the fan can be calculated using Equation 1.48:

$$Energy_{fan} = \frac{1.17 \times 10^4\ G_i\ DP}{\varepsilon} \qquad (1.48)$$

where Energy (in kilowatts) refers to the energy needed to move a given volumetric flow rate of air (acfm), G_i is the waste gas flow rate entering the absorber, P is the total pressure drop through the system (inches of H_2O) and is the combined fan-motor efficiency. Values for typically range from 0.4 to 0.7. Likewise, the electricity required by a recycle pump can be calculated using Equation 1.49:

$$Energy_{pump} = \frac{(0.746)\ (2.52 \times 10^{-1})\ L_i\ (pressure)}{\varepsilon} \qquad (1.49)$$

where 0.746 is the factor used to convert horsepower to kW, pressure is expressed in feet of water, and is the combined pump-motor efficiency.

The cost of electricity (C_e) is then given by:

$$C_e = Energy_{fan\ +\ pump} \begin{pmatrix} annual \\ operating \\ hours \end{pmatrix} \begin{pmatrix} cost\ of \\ electricity \end{pmatrix} \qquad (1.50)$$

where cost of electricity is expressed in units of $/KW-hr.

1.5.2 Indirect Annual Costs

Indirect annual costs (IC) include overhead, taxes, insurance, general and administrative (G&A), and capital recovery costs. The suggested factors for each of these items also appear in

Table 1.4. Overhead is assumed to be equal to 60 percent of the sum of operating, supervisory, and maintenance labor, and maintenance materials. Overhead cost is discussed in Section 1 of this manual.

The system capital recovery cost, *CRC*, is based on an estimated 15-year equipment life. (See Section 1 of this manual for a discussion of the capital recovery cost.) For a 15-year life and an interest rate of 7 percent, the capital recovery factor is 0.1098 The system capital recovery cost is then estimated by:

$$CRC = 0.1098\ TCI \tag{1.51}$$

G&A costs, property tax, and insurance are factored from total capital investment, typically at 2 percent, 1 percent, and 1 percent, respectively.

1.5.3 Total Annual Cost

Total annual cost (TAC) is calculated by adding the direct annual costs and the indirect annual costs.

$$TAC = DC + IC \tag{1.52}$$

1.6 Example Problem

The example problem presented in this section shows how to apply the gas absorber sizing and costing procedures presented in this chapter to control a waste gas stream consisting of HCl and air. This example problem will use the same outlet stream parameters presented in the thermal incinerator example problem found in Section 3.2, Chapter 2 of this manual. The waste gas stream entering the gas absorber is assumed to be saturated with moisture due to being cooled in the quench chamber. The concentration of HCl has also been adjusted to account for the change in volume.

1.6.1 Required Information for Design

The first step in the design procedure is to specify the conditions of the gas stream to be controlled and the desired pollutant removal efficiency. Gas and liquid stream parameters for this example problem are listed in Table 1.5.

The quantity of HCl can be written in terms of lb-moles of HCl per lb-moles of pollutant-free-gas (Y_i) using the following calculation:

$$Y_i = \frac{0.001871}{1 - 0.001871} = 0.00187 \frac{\text{lb} - \text{moles HCL}}{\text{lb - mole pollutant free gas}}$$

The solvent, a dilute aqueous solution of caustic, is assumed to have the same physical properties as water.

1.6.2 Determine Gas and Liquid Stream Properties

Once the properties of the waste gas stream entering the absorber are known. the properties of the waste gas stream exiting the absorber and the liquid streams entering and exiting the absorber need to be determined. The pollutant concentration in the entering liquid (X_i) is assumed to be zero. The pollutant concentration in the exiting gas stream (Y_o) is calculated using Equation 1.1 and a removal efficiency of 99 percent.

$$Y_o = 0.00187\left(1 - \frac{99}{100}\right) = 0.0000187$$

The liquid flow rate entering the column is calculated from the L_s/G_s ratio using Equation 1.2. Since Y_i, Y_o, and X_i are defined, the remaining unknown, X_o^*, is determined by consulting the equilibrium curve. A plot of the equilibrium curve-operating line graph for an HCl-water system is presented in Figure 1.8. The value of X_o^* is taken at the point on the equilibrium curve where Y_i intersects the curve. The value of Y_i intersects the equilibrium curve at an X value of 0.16.

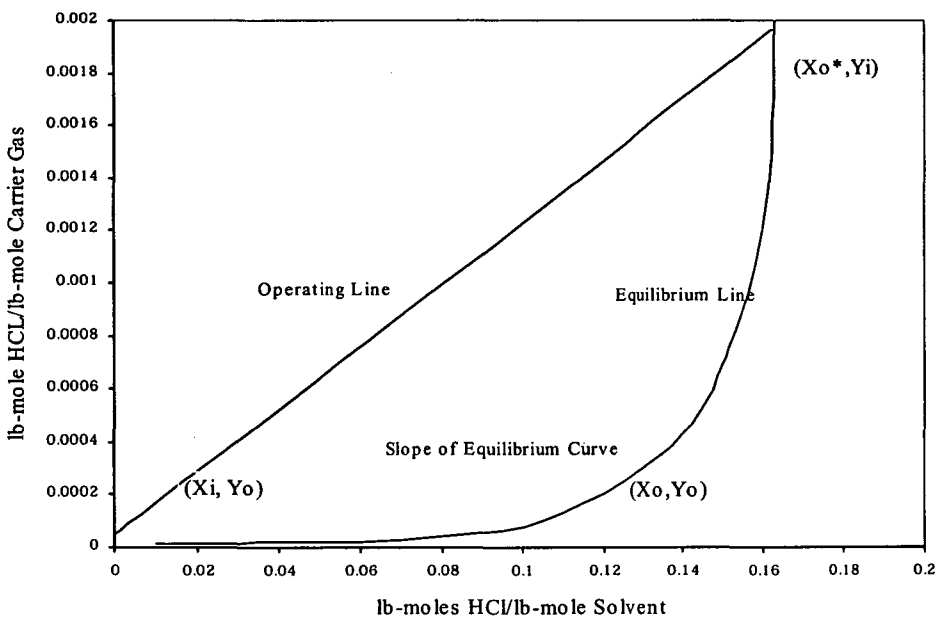


Figure 1.8: Equilibrium Curve Operating Line for the HCl-Water System [7]

Table 1.5: Example Problem Data

Parameters	Values
Stream Properties	
Waste Gas Flow Rate Entering Absorber	21,377 scfm (22,288 acfm)
Temperature of Waste Gas Stream	100°F
Pollutant in Waste Gas	HCl
Concentration of HCl Entering Absorber in Waste Gas	1871 ppmv
Pollutant Removal Efficiency	99% (molar basis)
Solvent	Water with caustic in solution
Density of Waste Gas [a]	0.0709 lb/ft^3
Density of Liquid [7]	62.4 lb/ft^3
Molecular Weight of Waste Gas[a]	29 lb/lb-mole
Molecular Weight of Liquid [7]	18 lb/lb-mole
Viscosity of Waste Gas[a]	0.044 lb/ft-hr
Viscosity of Liquid [7]	2.16 lb/ft-hr
Minimum Wetting Rate [7]	1.3 ft^2/hr
Pollutant Properties [b]	
Diffusivity of HCl in Air	0.725 ft^2/hr
Diffusivity of HCl in Water	1.02 x 10^{-4} ft^2/hr
Packing Properties [c]	
Packing type	2-inch ceramic Raschig rings
Packing factor: Fp	65
Packing constant: α	3.82
Packing constant: β	0.41
Packing constant: γ	0.45
Packing constant: φ	0.0125
Packing constant: b	0.22
Surface Area to Volume Ratio	28

[a] Reference [7], at 100°F
[b] Appendix 9A.
[c] Appendix 9B.

The operating line is constructed by connecting two points: (X_i, Y_o) and (X_o^*, Y_i). The slope of the operating line intersecting the equilibrium curve, $(L_s/G_s)min$, is:

$$\left(\frac{L_s}{G_s}\right)_{min} = \left(\frac{0.00187 - 0.0000187}{0.16 - 0}\right) = 0.0116$$

The actual L_s/G_s ratio is calculated using Equation 1.3. For this example, an adjustment factor of 1.5 will be used.

$$G_s = \frac{\left(60\,\frac{min}{hr}\right)\left(0.0709\,\frac{lb}{ft^3}\right)\left(22{,}288\ acfm\right)}{\left(29\,\frac{lb}{lb-mole}\right)(1 + 0.00187)} = 3{,}263\,\frac{lb-moles}{hr}$$

The flow rate of the solvent entering the absorber may then be calculated using Equation 1.5.

$$L_s = 0.0174\left(3{,}263\,\frac{lb-moles}{hr}\right) = 56.8\,\frac{lb-moles}{hr}$$

The values of $G_{mol,i}$ and $L_{mol,i}$ are calculated using Equations 1.6 and 1.7, respectively:

$$G_{mol,i} = \left(3{,}263\,\frac{lb-moles}{hr}\right)(1 + 0.00187) = 3{,}269\,\frac{lb-moles}{hr}$$

$$L_{mol,i} = \left(56.8\,\frac{lb-moles}{hr}\right)(1 + 0) = 56.8\,\frac{lb-moles}{hr}$$

The pollutant concentration exiting the absorber in the liquid is calculated using Equation 1.10.

$$x_o = \frac{0.00187 - 0.0000187}{0.0174} = \frac{0.106\ lb-moles\ HCL}{lb-mole\ solvent}$$

1.6.3 Calculate Absorption Factor

The absorption factor is calculated from the slope of the equilibrium line and the $L_{mol,i}/G_{mol,i}$ ratio. The slope of the equilibrium curve is based on the mole fractions of x_i, x_o, y_i^*, and y_o^*, which are calculated from X_i, X_o, Y_i^* and Y_o^* from Figure 1.8. From Figure 1.8, the value of Y_o^* in equilibrium with the X_o value of 0.106 is 0.0001. The values of Y_i^* and X^i are 0. The mole fraction values are calculated from the concentration values using Equations 1.13 through 1.16.

$$x_o = \frac{0.106}{1 + 0.106} = 0.096$$

$$y_o^* = \frac{0.0001}{1 + 0.0001} = 0.0001$$

The slope of the equilibrium fine from x_i to x_o is calculated from Equation 1.12:

$$m = \frac{0.0001 - 0}{0.096 - 0} = 0.00104$$

Since HCl is very soluble in water, the slope of the equilibrium curve is very small. The absorption factor is calculated from Equation 1.11.

$$AF = \frac{0.0174}{0.00104} = 17$$

1.6.4 Estimate Column Diameter

Once the inlet and outlet stream conditions are determined, the diameter of the gas absorber may be calculated using the modified generalized pressure drop correlation presented in Figure 1.5. The *Abscissa* value from the graph is calculated from Equation 1.17:

$$Abcissa = 0.0174\left(\frac{18}{29}\right)\sqrt{\frac{0.0709}{62.4}} = 0.000364$$

Since this value is outside the range of Figure 1.5, the smallest value (0.01) will be used as a default value. The *Ordinate* is calculated from Equation 1.19.

$$Ordinate = 10^{\left[-1.668 - 1.085(\log 0.01) - 0.297(\log 0.01)^2\right]} = 0.207$$

The superficial gas flow rate, $G_{sfr,i}$, is calculated using Equation 1.20. For this example calculation,

2-inch ceramic Rasching rings are selected as the packing. The packing factors for Raching rings are listed in Appendix B.

$$G_{sfr,i} = \sqrt{\frac{(0.207)\ (62.4)\ \left(0.0709\ \frac{lb}{ft^3}\right)\left(32.2\ \frac{ft}{sec^2}\right)}{(65)\ (1)\ (0.893)^{0.2}}} = 0.681\ \frac{lb}{sec - ft^2}$$

Once $G_{sfr,i}$ is determined, the cross-sectional area of the column is calculated using Equation 1.21.

$$A = \frac{\left(3{,}263\ \frac{lb-mol}{hr}\right)\left(29\ \frac{lb}{lb-mol}\right)}{\left(3600\ \frac{sec}{hr}\right)\left(0.681\ \frac{lb}{sec-ft^2}\right)(0.7)} = 55.1\ ft^2$$

The superficial liquid flow rate is determined using Equation 1.23.

$$L_{sfr,i} = \frac{\left(56.8\ \frac{lb-mol}{hr}\right)\left(18\ \frac{lb}{lb-mol}\right)}{55.1\ ft^2} = 18.6\ \frac{lb}{hr - ft^2}$$

At this point, it is necessary to determine if the liquid flow rate is sufficient to wet the packed bed. The minimum value of $L_{sfr,i}$ is calculated using Equation 1.24. The packing constant (a) is found in Appendix B.

$$\left(L_{sfr,i}\right)_{min} = \left(1.3\ \frac{ft^2}{hr}\right)\left(62.4\ \frac{lb}{ft^3}\right)\left(28\ \frac{ft^2}{ft^3}\right) = 2{,}271\ \frac{lb}{hr - ft^2}$$

The $L_{sfr,i}$ value calculated using the L/G ratio is far below the minimum value needed to wet the packed bed. Therefore, the new value, $(L_{sfr,i})_{min}$ will be used to determine the diameter of the absorber. The calculations for this revised diameter are shown in Appendix C. Appendix C shows that the cross-sectional area of the column is calculated to be 60 ft^2, $L_{mol,i}$ is 7572, and $G_{sfr,i}$ is 0.627 lb/sec-ft^2. (The diameter of the column is then calculated using Equation 1.22)

$$D = \sqrt{\frac{(4)\ (60\ ft^2)}{\Pi}} = 8.74\ ft$$

The value of X_o is then:

$$x_o = \frac{0.00187 - 0.0000187}{\frac{7{,}572}{3{,}263}} = 0.0008$$

Expressed in terms of mole fraction:

$$x_o = \frac{0.0008}{1 - 0.0008} = 0.0008$$

The value of y_o in equilibrium with x_o cannot be estimated accurately. However, the value will approach zero, and the value of AF will be extremely large:

$$AF = \frac{7{,}572}{(3{,}263)\ (\approx 0)} \rightarrow \infty$$

1.6.5 Calculate Column Surface Area

Since $x_i = 0$ and AF is large, Equation 1.26 will be used to calculate the number of transfer units:

$$n_{tu} = ln\left(\frac{0.00187}{0.0000187}\right) = 4.61$$

The height of a transfer unit is calculated from , AF, H_L, and H_G. The values of H_G and H_L are calculated from Equations 1.29 and 1.30:

$$H_G = \frac{3.82\left[(3{,}600)\ (0.7)\ (0.627)\right]^{0.41}}{2{,}271^{0.45}} \sqrt{\frac{0.044}{(0.725)\ (0.0709)}} = 2.24 \text{ ft}$$

$$H_L = 0.0125\left(\frac{2{,}271}{2.16}\right)^{0.22} \sqrt{\frac{2.16}{(0.000102)\ (62.4)}} = 1.06 \text{ ft}$$

The height of the transfer unit is calculated using Equation 1.28:

$$H_{tu} = (2.24 \text{ ft}) + \frac{1}{\infty}(1.06 \text{ ft}) = 2.24 \text{ ft}$$

The depth of packing is calculated from Equation 1.25.

$$H_{pack} = N_{tu} \times H_{tu} = (4.61)\ (2.24\ \text{ft}) = 10.3\ \text{ft}$$

The total height of the column is calculated from Equation 1.33:

$$H_{tower} = 1.40\ (10.3) + 1.02\ (8.74) + 2.81 = 26.1\ \text{ft}$$

The surface area of the column is calculated using Equation 1.34:

$$s = (3.14)\ (8.74)\left(26.1 + \frac{8.74}{2}\right) = 836\ \text{ft}^2$$

1.6.6 Calculate Pressure Drop

The pressure drop through the column is calculated using Equation 1.35.

$$\Delta P = (0.24)\ 10^{\frac{(0.17)\ (2,271)}{3,600}} \frac{\left[(0.7)\ (0.627)\right]^2}{0.0709}$$

$$= 0.83 \text{ inches water/foot packing}$$

The total pressure drop (through 10.3 feet of packing) equals 8.55 inches of water.

1.6.7 Equipment Costs

Once the system sizing parameters have been determined, the equipment costs can be calculated. For the purpose of this example, a gas absorber constructed of FRP will be costed using Equation 1.40.

$$\text{TTC}(\$) = 115(836) = \$96,140$$

The cost of 2-inch ceramic Raschig rings can be estimated from packing cost ranges presented in Section 1.5. The volume of packing required is calculated as:

$$\text{Volume of packing} = (60\ \text{ft}^2)(10.3\ \text{ft}) = 618\ \text{ft}^3$$

Using the average of the cost range for 2-inch ceramic packings, the total cost of packing is:

$$\text{Packing cost} = (\$20/\text{ft}^3)(618\ \text{ft}^3) = \$12{,}360$$

For this example problem, the cost of a pump will be estimated using vendor quotes. First, the flow rate of solvent must be converted into units of gallons per minute:

$$L(\text{gpm}) = \left(2{,}271 \frac{\text{lb}}{\text{h - ft}^2}\right)(60\ \text{ft}^2)\left(\frac{\text{gal}}{8.34\ \text{lb}}\right)\left(\frac{\text{hr}}{60\ \text{min}}\right) = 272\ \text{gpm}$$

The average price for a FRP pump of this size is $16/gpm at a pressure of 60 ft water, based on the vendor survey.[12] Therefore, the cost of the recycle pump is estimated as:

$$C_{pump} = (272\ \text{gpm})\left(\frac{\$16}{\text{gpm}}\right) = \$4{,}350\ \text{gpm}$$

For this example, the cost for a fan (FRP, backwardly-inclined centrifugal) can be calculated using the following equation:[18]

$$C_{fan} = 57.9d^{1.38}$$

where d is the impeller (wheel) diameter of the fan expressed in inches. For this gas flow rate and pressure drop, an impeller diameter of 33 inches is needed. At this diameter, the cost of the fan is:

$$C_{motor} = 104(hp)^{0.821}$$

The cost of a fan motor (three-phase, carbon steel) with V-belt drive, belt guard, and motor starter can be computed as follows:[18]

$$C_{motor} = 104\ (42.6)^{0.821} = \$2{,}260$$

As will be shown in Section 1.6.8, the electricity consumption of the fan is 32.0kW. Converting to horsepower, we obtain a motor size of 42.6 hp. The cost of the fan motor is:

$$Energy_{fan} = \frac{(1.17 \times 10^{-4})\ (22{,}288)\ (8.55)}{0.70} = 32.0\ kw$$

The total auxiliary equipment cost is:

$$\$4,350 + \$7,210 + \$2,260 = \$13,820$$

The total equipment cost is the sum of the absorber cost, the packing cost, and the auxiliary equipment cost:

$$EC = 96,140 + 12,360 + 13,820 = \$122,320$$

The purchased equipment cost including instrumentation, controls, taxes, and freight is estimated using Equation 1.43:

$$PEC = 1.18(122,320) = \$144,340$$

The total capital investment is calculated using Equation 1.44:

$$TCI = 2.20(144,340) = \$317,550 \quad \$318,000$$

1.6.8 Total Annual Cost

Table 1.6 summarizes the estimated annual costs using the suggested factors and unit costs for the example problem.

Direct annual costs for gas absorber systems include labor, materials, utilities, and wastewater disposal. Labor costs are based on 8,000 hr/year of operation. Supervisory labor is computed at 15 percent of operating labor, and operating and maintenance labor are each based on 1/2 hr per 8-hr shift.

The electricity required to run the fan is calculated using Equation 1.48 and assuming a combined fan-motor efficiency of 70 percent:

$$Energy_{fan} = \frac{\left(1.17 \times 10^{-4}\right)\ (22,288)\ (8.55)}{0.70} = 32.0\ \ kw$$

The energy required for the liquid pump is calculated using Equation 1.49. The capital cost of the pump was calculated using data supplied by vendors for a pump operating at a pressure of 60 feet of water. Assuming a pressure of 60 ft of water a combined pump-motor efficiency of 70 percent:

$$Energy_{pump} = \frac{(0.746)\ \left(2.52 \times 10^{-4}\right)\ (272)\ (60)\ (1)}{0.70} = 4.4\ \ kw$$

The total energy required to operate the auxiliary equipment is approximately 36.4 kW.

Table 1.6: Annual Costs for Packed Tower Absorber Example Problem

Cost Item	Calculations	Cost
Direct Annual Costs, DC		
Operating Labor Operator	$\frac{0.5\text{hr}}{\text{shift}} \times \frac{\text{shift}}{8\text{hr}} \times \frac{8{,}000\text{hr}}{\text{yr}} \times \frac{\$15.64}{\text{hr}}$	$7,820
Supervisor	15% of operator = 0.15 × 7,820	1,170
Operating materials Solvent (water)	$7.16\ \text{gpm} \times \frac{60\ \text{min}}{\text{hr}} \times \frac{8{,}000\text{hr}}{\text{yr}} \times \frac{\$0.20}{1000\text{gal}}$	690
Caustic Replacement	$\frac{3.06\text{lb-mole}}{\text{hr}} \times \frac{62\text{lb}}{\text{lb-mole}} \times \frac{8{,}000\text{hr}}{\text{yr}} \times \frac{\text{ton}}{2000\text{lb}} \times \frac{1}{0.76\ \text{ton}} \times \frac{\$300}{}$	299,560
Wastewater disposal	$7.16\text{gpm} \times \frac{60\ \text{min}}{\text{hr}} \times \frac{8{,}000\ \text{hr}}{\text{yr}} \times \frac{\$3.80}{100\text{gal}}$	13,060
Maintenance Labor	$\frac{0.5}{\text{shift}} \times \frac{\text{shift}}{8\text{hr}} \times \frac{8{,}000\text{hr}}{\text{yr}} \times \frac{\$17.21}{\text{hr}}$	8,610
Material	100% of maintenance labor	8,610
Electricity	$36.4\text{kw} \times \frac{8{,}000\text{hr}}{\text{yr}} \times \frac{\$0.0461}{\text{kWh}}$	13,420
Total DC		$352,940
Indirect Annual Costs, IC		
Overhead	60% of total labor and maintenance material: = 0.6(7,820 + 1,170 + 8,610 + 8,610)	15,730
Administrative charges	2% of Total Capital Investment = 0.02($317,550)	6,350
Property tax	1% of Total Capital Investment = 0.01($317,550)	3,180
Insurance	1% of Total Capital Investment = 0.01($317,550)	3,180
Capital recovery[a]	0.1315 × $317,550	41,760
Total IC		$70,200
Total Annual Cost (rounded)		**$423,000**

[a] The capital recovery cost factor, CRF, is a function of the absorber equipment life and the opportunity cost of the capital (i.e., interest rate). For this example, assume a 15-year equipment life and a 10% interest rate.

The cost of electricity, C_e, is calculated using Equation 1.50 and with the cost per kWh shown in Table 1.6.

$$C_e = (36.4\text{kW})(8{,}000\text{ h/yr})(\$0.0461/\text{kWh}) = \$13{,}420/\text{yr}$$

The costs of solvent (water), wastewater disposal, and caustic are all dependent on the total system throughput and the fraction of solvent discharged as waste. A certain amount of solvent will be wasted and replaced by a fresh solution of water and caustic in order to maintain the system's pH and solids content at acceptable levels. Based on the vendor survey, a maximum solids content of 10 percent by weight will be the design basis for this example problem.[12] The following calculations illustrate the procedure used to calculate how much water and caustic are needed, and how much solvent must be bled off to maintain system operability.

From previous calculations, $L_{mol,i}$ = 7,572 lb-moles/hr. The mass flow rate is calculated as:

$$L_{mass} = \left(7{,}572\,\frac{\text{lb}-\text{mole}}{\text{hr}}\right)\left(18\,\frac{\text{lb}}{\text{lb}-\text{mole}}\right) = 136{,}300\,\frac{\text{lb}}{\text{hr}}$$

With $G_{mol,i}$ at 3,263 lb-moles/hr, the mass flow rate of the gas stream is calculated as:

$$G_{mass} = \left(3{,}263\,\frac{\text{lb}-\text{mole}}{\text{hr}}\right)\left(29\,\frac{\text{lb}}{\text{lb}-\text{mole}}\right) = 94{,}800\,\frac{\text{lb}}{\text{hr}}$$

The amount of HCl in the gas stream is calculated on a molar basis as follows:

$$G_{mass,\text{HCL}} = \left(3{,}263\,\frac{\text{lb}-\text{mole}}{\text{hr}}\right)\left(1874\,\frac{\text{ppmv}}{1\times10^{6}}\right) = 6.12\,\frac{\text{lb - molHCL}}{\text{hr}}$$

On a mass basis:

$$G_{mass,\text{HCL}} = \left(6.12\,\frac{\text{lb - molHCL}}{\text{hr}}\right)\left(36.5\,\frac{\text{lb}}{\text{lb - mole}}\right) = 223.4\,\frac{\text{lb HCL}}{\text{hr}}$$

For this example problem, the caustic is assumed to be Na_2O, with one mole of caustic required for neutralizing 2 moles of HCl. Therefore, 3.06 lb-moles/hr of caustic are required.

The unit cost of a 76 percent solution of Na_2O is given in Table 1.6. The annual cost is calculated from:

$$C_c = \left(3.06 \frac{\text{lb - moles}}{\text{hr}}\right)\left(62 \frac{\text{lb}}{\text{lb - mole}}\right)\left(\frac{8{,}000 \text{ hr}}{\text{yr}}\right)\left(\frac{\text{ton}}{2{,}000 \text{ lb}}\right)\left(\frac{1}{0.76}\right)\left(\frac{\$300}{\text{ton}}\right)$$

$$= \$299{,}560 \text{ yr}$$

Mass of the salt formed in this chemical reaction, NaCl, is calculated as:

$$Mass_{NaCl} = \left(223.4 \frac{\text{lb - HCL}}{\text{hr}}\right)\left(\frac{\text{lb - mole}}{36.5 \text{ lb HCL}}\right)\left(\frac{1 \text{ lb - mole NaCl}}{\text{lb - mole HCL}}\right)\left(\frac{58.5 \text{ lb NaCl}}{\text{lb - mole NaCl}}\right)$$

$$= 358.1 \frac{\text{lb NaCl}}{\text{hr}}$$

If the maximum concentration of NaCl in the wastewater (ww) is assumed to be 10 weight percent, the wastewater volume flow rate is calculated as:

$$Wastewater_{flowrate} = \left(358.1 \frac{\text{lb NaCl}}{\text{hr}}\right)\left(\frac{1 \text{ lb ww}}{0.1 \text{ lb NaCl}}\right)\left(\frac{\text{gal ww}}{8.34 \text{ lb ww}}\right)\left(\frac{1 \text{ hr}}{60 \text{ min}}\right)$$

$$= 7.16 \text{ gpm}$$

where 8.34 is the density of the wastewater.

The cost of wastewater disposal is:[1]

$$C_{ww} = (7.16 \text{ gpm})\left(\frac{60 \text{ min}}{\text{lhr}}\right)\left(8{,}000 \frac{\text{hr}}{\text{yr}}\right)\left(\frac{\$3.80}{1{,}000 \text{ gal}}\right) = \frac{\$13{,}060}{\text{yr}}$$

The cost of solvent (water) is:

$$C_s = (7.16 \text{ gpm})\left(\frac{60 \text{ min}}{\text{lhr}}\right)\left(8{,}000 \frac{\text{hr}}{\text{yr}}\right)\left(\frac{\$0.20}{1{,}000 \text{ gal}}\right) = \frac{\$690}{\text{yr}}$$

[1]Because the wastewater stream contains only NaCl, it probably will not require pretreatment before discharge to a municipal wastewater treatment facility. Therefore, the wastewater disposal unit cost shown here is just a sewer usage rate. This unit cost ($3.80/1,000 gal) is the average of the rates charged by the seven largest municipalities in North Carolina.[20] These rates range from approximately $2 to $6/1,000 gal. This wide range is indicative of the major differences among sewer rates throughout the country. Indirect annual costs include overhead, administrative charges, property tax, insurance, and capital recovery. Total annual cost is estimated using Equation 1.52. For this example case, the total annual cost is estimated to be $423,000 per year (Table 1.6).

1.6.9 Alternate Example

In this example problem the diameter of a gas absorber will be estimated by defining a pressure drop. A pressure drop of 1 inch of water per foot of packing will be used in this example calculation. Equation 1.38 will be used to calculate the ordinate value relating to an abscissa value. If the $L_{mol,i}/G_{mol,i}$ ratio is known, the *Abscissa* can be calculated directly. The *Ordinate* value is then:

$$\begin{aligned} Ordinate &= \exp\,[-4.0950-1.00121n(0.0496)-0.1587(1n\ 0.0496)^2 + \\ &\quad 0.0080(1n\ 0.0496)^3 + 0.0032(1n\ 0.0496)^4] \\ &= 0.084 \end{aligned}$$

The value of G_{sfr} is calculated using Equation 1.39.

$$G_{sfr,i} = \sqrt{\frac{(62.4 - 0.0709)\ (0.0709)\ (32.2)\ (0.084)}{65\ (0.893)^{0.1}}} = 0.43\,\frac{lb}{ft^2 - sec}$$

The remaining calculations are the same as in Section 1.3.4, except the flooding factor is not used in the equations.

1.7 Acknowledgments

The authors gratefully acknowledge the following companies for contributing data to this chapter:

- Air Plastics, Inc. (Cincinnati, OH)
- April, Inc. (Teterboro, NJ)
- Anderson 2000, Inc. (Peachtree City, GA)
- Calvert Environmental (San Diego, CA)
- Ceilcote Air Pollution Control (Berea, OH)
- Croll-Reynolds Company, Inc. (Westfield, NJ)
- Ecolotreat Process Equipment (Toledo, OH)
- Glitsch, Inc. (Dallas, TX)
- Interel Corporation (Englewood, CO)
- Jaeger Products, Inc. (Spring, TX)
- Koch Engineering Co., Inc. (Wichita, KS)
- Lantec Products, Inc. (Agoura Hills, CA)
- Midwest Air Products Co., Inc. (Owosso, MI)
- Monroe Environmental Corp., (Monroe, MI)
- Norton Chemical Process Products (Akron, OH)

References

[1] *Control Technologies for Hazardous Air Pollutants,* Office of Research and Development, U.S. Environmental Protection Agency, Research Triangle Par,, North Carolina, Publication No. EPA 625/6-91-014.

[2] McInnes, R., K. Jameson, and D. Austin, "Scrubbing Toxic Inorganics", *Chemical Engineering,* September 1990, pp. 116-121.

[3] Letter from Jose L. Bravo of Jaeger Products, Inc., to William M. Vatavuk, U. S. Environmental Protection Agency, June 8, 1992.

[4] Treybal, Robert E., *Mass Transfer Operations* (Third edition), McGraw-Hill Book Company, New York, 1980.

[5] Letter from Jack D. Brady of Anderson 2000, Inc., to William M. Vatavuk, U.S. Environmental Protection Agency, June 9, 1992.

[6] Letter from S. Raymond Woll of Air Products, Inc., to William M. Vatavuk, U.S. Environmental Protection Agency, June 25, 1992.

[7] Perry, R.H. and C.H. Chilton, Eds., *Chemical Engineers' Handbook* (Sixth edition), McGraw-Hill Book Company, New York, 1984.

[8] Crowe, Charles R., and D. Cooper, "Brick/Membrane Linings Pass the Acid Test", *Chemical Engineering,* July 1988, pp. 83-86.

[9] Harrison, Mark E., and John J. France, "Distillation Column Troubleshooting, Part 2: Packed Columns", *Chemical Engineering*, April 1989, pp. 121-128.

[10] Coker, A.K., "Understanding the Basics of Packed-Column Design", *Chemical Engineering Progress,* November 1991, pp. 93-99.

[11] Telephone conversation between Roy Oommen, Radian Corporation and Gerald Nealon, Norton Process Equipment, April 4, 1992.

[12] Gas absorber questionnaire responses from nine gas absorber vendors to Radian Corporation August-December, 1991.

[13] Buonicore, A.J., and L. Theodore, *Industrial Control Equipment for Gaseous Pollutants, Volume I,* CRC Press, Inc., Cleveland, Ohio, 1975.

[14] Strigle, Ralph F., *Random Packings and Packed Towers, Design Applications,* Gulf Publishing Company, Houston, Texas, 1987.

[15] Questionnaire response from K.C. Lang of Lantec Products, Inc. To R.V. Oommen, Radian Corporation, August 23, 1991.

[16] Memorandum from Vatavuk, W.M. of U.S. Environmental Protection Agency to Martha Smith, U.S. EPA, March 27, 1992.

[17] Packing vendor questionnaire responses from seven packing vendors to Radian Corporation, August 1991 - January 1992.

[18] Vatavuk, W.M., "Pricing Equipment for Air-Pollution Control", *Chemical Engineering,* May 1990, pp. 126-130.

[19] Vatavuk, W.M., and R.B. Neveril, "Estimating Costs of Pollution Control Systems, Part II: Factors for Estimating Capital and Operating Costs", *Chemical Engineering*, November 3, 1980, pp. 157-162.

[20] Telephone conversation between William M. Vatavuk, U.S. Environmental Protection Agency, and Cindy Kling, City of Raleigh, N.C., July 16, 1992.

[21] "Air Pollution Engineering Manual" (AP-40), (Second Edition), Danielson, John A., Los Angeles County Air Pollution Control District, CA, May 1973.

---

[1] For information on escalating these prices to more current dollars, refer to the EPA report *Escalation Indexes for Air Pollution Control Costs* and updates thereto, all of which are installed on the OAQPS Technology Transfer Network (CTC Bulletin Board).

Appendix A

Properties of Pollutants

Table 1.7: List of Design Variables

	Variable	Symbol	Units
▸	Surface to volume ratio of packing	a	ft^2/ft^3
	Cross-sectional area of absorber	A	ft^2
	Abscissa value from plot of generalized press drop correlation	Abscissa	—
	Absorption factor	AF	—
	Diameter of absorber	D	feet
▸	Diffusivity of pollutant in gas	D_G	ft^2/hr
▸	Diffusivity of pollutant in liquid	D_L	ft^2/hr
▸	Flooding factor	f	—
▸	Packing factor	F_p	—
▸	Waste gas flow rate entering absorber	G_i	acfm
	Waste gas flow rate exiting absorber	G_o	acfm
	Waste gas molar flow rate entering absorber	G_{mol}	lb-moles/h
	Molar flow rate of pollutant free gas	G_s	lb-moles/h
	Waste gas superficial flow rate entering absorber	$G_{sfr,i}$	$lb/sec\text{-}ft^2$
	Height of gas transfer unit	H_G	feet
	Height of liquid transfer unit	H_L	feet
	Height of overall transfer unit	H_{tu}	feet
	Height of packing	H_{pack}	feet
	Height of absorber	H_{tower}	feet
	Pressure drop constants	k_0, k_1, k_2, k_3, k_4	—
	Liquid rate entering absorber	L_i	gpm
	Liquid rate exiting absorber	L_o	gpm
	Liquid molar flow rate entering absorber	$L_{mol,i}$	lb-moles/h
	Molar flow rate of pollutant free solvent	$L_?$	lb-moles/h
	Liquid superficial flow rate entering absorber	$L_{sfr,i}$	$lb/hr\text{-}ft^2$
	Slope of equilibrium line	m	—
▸	Molecular weight of gas stream	MW_G	lb/lb-mole
▸	Molecular weight of the liquid stream	MW_L	lb/lb-mole
▸	Minimum wetting rate	MWR	ft^2/hr
	Number of overall transfer units	N_{tu}	—
	Ordinate value from plot of generalized pressure drop correlation	Ordinate	—
	Surface area of absorber	S	ft^2
▸	Temperature of solvent	T	K

Table 1.7: List of Design Variables (continued)

	Variable	Symbol	Units
	Maximum pollutant concentration in liquid phase in equilibrium with pollutant entering column in gas phase	X^*_o	$\frac{\text{lb-moles pollutant}}{\text{lb-moles pollutant free solvent}}$
	Pollutant concentration exiting absorber in liquid	X_o	$\frac{\text{lb-moles pollutant}}{\text{lb-moles pollutant free solvent}}$
	Mole fraction of pollutant entering absorber in waste gas	y_i	$\frac{\text{lb-moles pollutant}}{\text{lb-mole of total gas}}$
	Mole fraction of pollutant in gas phase in equilibrium with mole fraction of pollutant entering in the liquid phase	y^*_i	$\frac{\text{lb-moles pollutant}}{\text{lb-mole of total gas}}$
	Mole fraction of pollutant exiting scrubber in waste gas	y_o	$\frac{\text{lb-moles pollutant}}{\text{lb-mole of total gas}}$
	Mole fraction of pollutant in gas phase in equilibrium with mole fraction of pollutant exiting in the liquid phase	y^*_o	$\frac{\text{lb-moles pollutant}}{\text{lb-mole of total gas}}$
▸	Pollutant concentration entering scrubber in waste gas	Y_i	$\frac{\text{lb-moles pollutant}}{\text{lb-moles pollutant free gas}}$
	Pollutant concentration entering scrubber in equilibrium with concentration in liquid phase	Y^*_i	$\frac{\text{lb-moles pollutant}}{\text{lb-moles pollutant free gas}}$
	Pollutant concentration exiting scrubber in waste gas	Y_o	$\frac{\text{lb-moles pollutant}}{\text{lb-moles pollutant free gas}}$
▸	Pollutant removal efficiency	η	%
	Pollutant concentration exiting scrubber in equilibrium with concentration in liquid phase	Y_o	$\frac{\text{lb-moles pollutant}}{\text{lb-mole of total gas}}$
▸	Density of waste gas stream	ρ_G	lb/ft^3
▸	Density of liquid stream	ρ_L	lb/ft^3
▸	Viscosity of waste gas	μ_G	lb/ft-hr
▸	Viscosity of solvent	μ_L	lb/ft-hr
	Ratio of solvent density to water	Ψ	—
	Pressure drop	ΔP	inches H_2O/feet of packing
▸	Packing factors	$a, \alpha, \phi, b, \beta, \gamma, c, j$	—

▸ Denotes required input data.

Table 1.8: Physical Properties of Common Pollutants[a]

Pollutant	Molecular Weight $\left(\frac{lb}{lb-mole}\right)$	Diffusivity in Air at 25°C (cm^2/sec)	Diffusivity in Water at 20°C (cm^2/sec)x10^5
Ammonia	17	0.236	1.76
Methanol	32	0.159	1.28
Ethyl Alcohol	46	0.119	1.00
Propyl Alcohol	60	0.100	0.87
Butyl Alcohol	74	0.09	0.77
Acetic Acid	60	0.133	0.88
Hydrogen Chloride	36	0.187	2.64
Hydrogen Bromide	36	0.129	1.93
Hydrogen Fluoride	20	0.753	3.33

[a] Diffusivity data taken from Reference [7, 21].

Appendix B

Packing Characteristics

Table 1.9: Packing Factors for Various Packings[3,7,10,13]

Packing Type	Construction Level	Nominal Diameter (inches)	Fp	a
Raschig rings	ceramic	1/2	640	111
		5/8	380	100
		3/4	255	80
		1	160	58
		1 1/2	95	38
		2	65	28
		3	37	
Raschig rings	metal	1/2	410	118
		5/8	290	
		3/4	230	72
		1	137	57
		1 1/2	83	41
		2	57	31
		3	32	21
Pall rings	metal	5/8	70	131
		1	48	66
		1 1/2	28	48
		2	20	36
		3 1/2	16	
Pall rings	polypropylene	5/8	97	110
		1	52	63
		1 1/2	32	39
		2	25	31
Berl saddles	ceramic	1/2	240	142
		3/4	170	82
		1	110	76
		1 1/2	65	44
		2	45	32
Intalox saddles	ceramic	1/2	200	190
		3/4	145	102
		1	98	78
		1 1/2	52	60
		2	40	36
		3	22	
Tri-Packs®	plastic	2	16	48
		3 1/2	12	38

Table 1.10: Packing Constants Used to Estimate H_G [1, 3, 7, 13]

Packing Type	Size (inches)	Packing Constants			Applicable Range[a]	
		α	β	γ	G_{sfr}	L_{sfr}
Raschig Rings	3/8	2.32	0.45	0.47	200-500	500-1,500
	1	7.00	0.39	0.58	200-800	400-500
	1	6.41	0.32	0.51	200-600	500-4,500
	1 1/2	1.73	0.38	0.66	200-700	500-1,500
	1 1/2	2.58	0.38	0.40	200-700	1,500-4,500
	2	3.82	0.41	0.45	200-800	500-4,500
Berl Saddles	1/2	32.4	0.30	0.74	200-700	500-1,500
	1/2	0.81	0.30	0.24	200-700	1,500-4,500
	1	1.97	0.36	0.40	200-800	400-4,500
	1 1/2	5.05	0.32	0.45	200-1,000	400-4,500
Partition Rings	3	640	0.58	1.06	150-900	3,000-10,000
LanPac®	2.3	7.6	0.33	-0.48	400-3,000	500-8,000
Tri-Packs®	2	1.4	0.33	0.40	100-900	500-10,000
	3 1/2	1.7	0.33	0.45	100-2,000	500-10,000

[a]Units of lb/hr-ft^2

Table 1.11: Packing Constants Used to Estimate H_L [1, 3, 13]

Packing Type	Size (inches)	Packing Constants		Applicable Range[a]
		φ	b	L^a_{sfr}
Raschig Rings	3/8	0.00182	0.46	400-15,000
	1	0.00357	0.35	400-15,000
	1 1/2	0.0100	0.22	400-15,000
	2 1/2	0.0111	0.22	400-15,000
	2	0.0125	0.22	400-15,000
Berl Saddles	1/2	0.00666	0.28	400-15,000
	1	0.00588	0.28	400-15,000
	1 1/2	0.00625	0.28	400-15,000
Partition Rings	3	0.0625	0.09	3,000-14,000
LanPac®	2.3	0.0039	0.33	500-8,000
	3.5	0.0042	0.33	500-8,000
Tri-Packs®	2	0.0031	0.33	500-10,000
	3 1/2	0.0040	0.33	500-10,000

[a]Units of lb/hr-ft^2

Table 1.12: Packing Constants Used to Estimate Pressure Drop [1, 7, 13]

Packing Type	Construction Material	Nominal Diameter (inches)	*c*	*j*
Raschig rings	ceramic	1/2	3.1	0.41
		3/4	1.34	0.26
		1	0.97	0.25
		1 1/4	0.57	0.23
		1 1/2	0.39	0.23
		2	0.24	0.17
Raschigrings	metal	5/8	1.2	0.28
		1	0.42	0.21
		11/2	0.29	0.20
		2	0.23	0.135
Pallrings	metal	5/8	0.43	0.17
		1	0.15	0.16
		11/2	0.08	0.15
		2	0.06	0.12
Berlsaddles	ceramic	1/2	1.2	0.21
		3/4	0.62	0.17
		1	0.39	0.17
		11/2	0.21	0.13
Intaloxsaddles	ceramic	1/2	0.82	0.20
		3/4	0.28	0.16
		1	0.31	0.16
		11/2	0.14	0.14

Appendix C

Minimum Wetting Rate Analysis

Minimum Wetting Rate Analysis

As explained in the design procedures, the liquid flow rate entering the column must be high enough to effectively wet the packing. If the liquid flow rate, as determined theoretically in Equation 1.23, is lower than the flow rate dictated by the minimum wetting rate, calculated in Equation 1.24, then the packing will not be wetted sufficiently to ensure mass transfer between the gas and liquid phases. The minimum liquid flow rate should then be used as a default value. The superficial gas flow rate, $G_{sfr,}$, and cross-sectional area of the column must then be recalculated to account for the increased liquid flow rate. The approach is outlined below

- The value of $L_{mol,i}$ must be recalculated from the value of $(L_{sfr,i})_{\min}$ using the equation:

$$L_{mol,i} = \frac{\left(L_{sfr,i}\right)_{\min} A}{(MW)_L}$$

The value of A (the cross-sectional area of the absorber column) is the only unknown in the equation.

- The *Abscissa* value is calculated in terms of A by substituting the new $L_{mol,i}$ into Equation 1.17.

- The value of $G_{sfr,i}$ is recalculated by rearranging Equation 1.21, with A as the only unknown.

- The *Ordinate* value is calculated in terms of A from the new $G_{sfr,i}$ using the Equation 1.18.

- An iterative process is used to determine A, *Abscissa*, and *Ordinate*. Values of A are chosen and the *Abscissa* and *Ordinate* values are calculated. The *Ordinate* value corresponding to the *Abscissa* value is determined from Figure 1.5 (or Equation .19), and this value is compared to the *Ordinate* value calculated using Equation 1.18. This process is continued until both *Ordinate* values are equal.

Step 1: The first step is to recalculate the liquid flow rate. The liquid molar flow rate may be calculated using Equation 1.23.

$$L_{mol,i} = \left(2{,}271 \frac{\text{lb}}{\text{hr} - \text{ft}^2}\right)\left(\frac{\text{lb} - \text{mole}}{18\ \text{lb}}\right) A = \left(126.2\ \frac{\text{lb} - \text{mole}}{\text{hr} - \text{ft}^2}\right) A$$

Step 2: The *Abscissa* value from Figure 1.5, and presented in Equation 1.17, is calculated as:

$$Abscissa = \frac{\left(126.2\ \frac{\text{lb-mole}}{\text{hr-ft}^2}\right)A}{3{,}263\ \frac{\text{lb-mole}}{\text{hr}}}\left(\frac{18}{29}\right)\sqrt{\frac{0.0709}{62.4}}$$

Step 3: The value of $G_{sfr.i}$ is then recalculated in terms of the cross-sectional area of the column.

$$G_{sfr} = \frac{\left(3{,}263\ \frac{\text{lb-mole}}{\text{hr}}\right)\left(29\ \frac{\text{lb}}{\text{lb-mole}}\right)}{\left(3{,}600\ \frac{\text{sec}}{\text{hr}}\right)(0.7)A} = \frac{37.6}{A}$$

Step 4: The ordinate value from Figure 1.5, and presented in Equation 1.18, is calculated as:

Step 5: At this point the simplest solution is an iterative approach. Choose a value for A, calculate the *Abscissa* value using Equation 1.53, and find the corresponding *Ordinate* value off the flooding curve in Figure 1.5 (or use Equation 1.19 to calculate the *Ordinate* value). Compare the calculated *Ordinate* value from Equation 1.54 to the value obtained from the graph or from Equation 1.19. By continuing this process until the *Ordinate* values converge the value of A is determined to be 60 ft^2. The following table illustrates the intermediate steps in the calculational process.

Assumed Value of A	Abscissa Calculated From Equation 9.53	Ordinate Calculated From Equation 9.19	Ordinate Calculated From Equation 9.54
65	0.0526	0.1714	0.1493
62	0.0503	0.1740	0.1642
60	0.0485	0.1757	0.1752

The value of G_{sfr} is then:

$$G_{sfr} = \frac{37.6}{60} = 0.627 \frac{\text{lb}}{\text{sec} - \text{ft}^2}$$

The liquid molar flow rate is:

$$L_{mol,i} = (126.2)\ (60) = 7{,}572 \frac{\text{lb} - \text{mole}}{\text{hr}}$$

The diameter and height of the column using the results of this calculation are presented in the first Example Problem.

EPA/452/B-02-001

Section 6

Particulate Matter Controls

EPA/452/B-02-001

Chapter 1

Baghouses and Filters

James H. Turner
Research Triangle Institute
Research Triangle Park, NC 22709

John D. McKenna
John C. Mycock
Arthur B. Nunn
ETS, Inc.
Roanoke, VA 24018-4394

William M. Vatavuk
Innovative Strategies and Economics Group, OAQPS
U.S. Environmental Protection Agency
Research Triangle Park, NC 22711

December 1998

Contents

1.1 Introduction 1-4

1.2 Process Description 1-5
1.2.1 Shaker Cleaning 1-5
1.2.2 Reverse-air Cleaning 1-6
1.2.3 Pulse-jet Cleaning 1-7
1.2.3.1 Caged Filters 1-7
1.2.3.2 Cartridge Filters 1-9
1.2.4 Sonic Cleaning 1-10
1.2.5 Auxiliary Equipment 1-12
1.2.6 Fabric Filtration Theory 1-13
1.2.6.1 Reverse Air/Shake Deflate Baghouses 1-15
1.2.6.2 Pulse-Jet Baghouses 1-17

1.3 Design Procedure 1-20
1.3.1 Gas-to-Cloth Ratio 1-20
1.3.1.1 Gas-to-Cloth Ratio From Similar Applications 1-20
1.3.1.2 Gas-to-Cloth Ratio From Manufacturer's Methods 1-21
1.3.1.3 Gas-to-Cloth Ratio From Theoretical/Empirical Equations 1-26
1.3.2 Pressure Drop 1-30
1.3.3 Particle Characteristics 1-31
1.3.4 Gas Stream Characteristics 1-31
1.3.4.1 Temperature 1-31
1.3.4.2 Pressure 1-31
1.3.5 Equipment Design Considerations 1-33
1.3.5.1 Pressure or Suction Housings 1-33
1.3.5.2 Standard or Custom Construction 1-33
1.3.5.3 Filter Media 1-34

1.4 Estimating Total Capital Investment 1-35
1.4.1 Equipment Cost 1-35
1.4.1.1 Bare Baghouse Costs 1-35
1.4.1.2 Bag Costs 1-41
1.4.1.3 Auxiliary Equipment 1-41
1.4.2 Total Purchased Cost 1-41
1.4.3 Total Capital Investment 1-43

1.5 Estimating Total Annual Costs 1-44
1.5.1 Direct Annual Cost 1-44
1.5.1.1 Gas-to-Cloth Ratio From Manufacturer's Methods 1-44
1.5.1.2 Gas-to-Cloth Ratio From Theoretical/Empirical Equations 1-44
1.5.1.3 Maintenance 1-44
1.5.1.4 Replacement Parts 1-46
1.5.1.5 Electricity 1-47
1.5.1.6 Fuel 1-47
1.5.1.7 Water 1-47
1.5.1.8 Compressed Air 1-48
1.5.1.9 Dust Disposal 1-48

1.5.2 Indirect Annual Cost 1-48
1.5.3 Recovery Credits 1-49
1.5.4 Total Annual Cost 1-49

1.6 Example Problem 1-49

1.7 Acknowledgements 1-56

References 1-56

1.1 Introduction

A fabric filter unit consists of one or more isolated compartments containing rows of fabric bags in the form of round, flat, or shaped tubes, or pleated cartridges. Particle-laden gas passes up (usually) along the surface of the bags then radially through the fabric. Particles are retained on the upstream face of the bags, and the cleaned gas stream is vented to the atmosphere. The filter is operated cyclically, alternating between relatively long periods of filtering and short periods of cleaning. During cleaning, dust that has accumulated on the bags is removed from the fabric surface and deposited in a hopper for subsequent disposal.

Fabric filters collect particles with sizes ranging from submicron to several hundred microns in diameter at efficiencies generally in excess of 99 or 99.9 percent. The layer of dust, or dust cake, collected on the fabric is primarily responsible for such high efficiency. The cake is a barrier with tortuous pores that trap particles as they travel through the cake. Gas temperatures up to about 500°F, with surges to about 550°F can be accommodated routinely in some configurations. Most of the energy used to operate the system appears as pressure drop across the bags and associated hardware and ducting. Typical values of system pressure drop range from about 5 to 20 inches of water. Fabric filters are used where high-efficiency particle collection is required. Limitations are imposed by gas characteristics (temperature and corrosivity) and particle characteristics (primarily stickiness) that affect the fabric or its operation and that cannot be economically accommodated.

Important process variables include particle characteristics, gas characteristics, and fabric properties. The most important design parameter is the air- or gas-to-cloth ratio (the amount of gas in ft^3/min that penetrates one ft^2 of fabric) and the usual operating parameter of interest is pressure drop across the filter system. The major operating feature of fabric filters that distinguishes them from other gas filters is the ability to renew the filtering surface periodically by cleaning. Common furnace filters, high efficiency particulate air (HEPA) filters, high efficency air filters (HEAFs), and automotive induction air filters are examples of filters that must be discarded after a significant layer of dust accumulates on the surface. These filters are typically made of matted fibers, mounted in supporting frames, and used where dust concentrations are relatively low. Fabric filters are usually made of woven or (more commonly) needlepunched felts sewn to the desired shape, mounted in a plenum with special hardware, and used across a wide range of dust concentrations.

Another type of fabric filter developed in the 1970s and 1980s is the electrostatically enhanced filter. Pilot plant baghouses employing this technology have shown substantially lower pressure drops than conventional filter designs. Further, some cost analyses have shown that electrostatically enhanced baghouses could have lower lifetime costs than convention baghouses. The purpose of this chapter, however, is to focus only on currently available commercial filters. Readers interested in electrostatically enhanced filtration may consult such references as Van Osdell et al. [1] , Viner et al. [2] , or Donovan [3].

1.2 Process Description

In this section, the types of fabric filters and the auxiliary equipment required are discussed first from a general viewpoint. Then, fabric filtration theory as applied to each type of filter is discussed to lay a foundation for the sizing procedures. Fabric filters can be categorized by several means, including type of cleaning (shaker, reverse-air, pulse-jet), direction of gas flow (from inside the bag towards the outside or vice versa), location of the system fan (suction or pressure), or size (low, medium, or high gas flow quantity). Of these four approaches, the cleaning method is probably the most distinguishing feature. Fabric filters are discussed in this section based on the type of cleaning employed.

1.2.1 Shaker Cleaning

For any type of cleaning, enough energy must be imparted to the fabric to overcome the adhesion forces holding dust to the bag. In shaker cleaning, used with inside-to-outside gas flow, energy transfer is accomplished by suspending the bag from a motor-driven hook or framework that oscillates. Motion may be imparted to the bag in several ways, but the general effect is to create a sine wave along the fabric. As the fabric moves outward from the bag centerline during portions of the wave action, accumulated dust on the surface moves with the fabric. When the fabric reaches the limit of its extension, the patches of dust have enough inertia to tear away from the fabric and descend to the hopper.

For small, single-compartment baghouses, usually operated intermittently, a lever attached to the shaker mechanism may be operated manually at appropriate intervals, typically at the end of a shift. In multi-compartment baghouses, usually operated continuously, a timer or a pressure sensor responding to system pressure drop initiates bag shaking automatically. The compartments operate in sequence so that one compartment at a time is cleaned. Forward gas flow to the compartment is stopped, dust is allowed to settle, residual gas flow stops, and the shaker mechanism is switched on for several seconds to a minute or more. The settling and shaking periods may be repeated, then the compartment is brought back on-line for filtering. As a result of no forward flow through the compartment, the baghouse collecting area must be increased to compensate for that portion being out of service at any time for cleaning. Figure 1.1 illustrates a shaker-cleaned baghouse.

Parameters that affect cleaning include the amplitude and frequency of the shaking motion and the tension of the mounted bag. The first two parameters are part of the baghouse design and generally are not changed easily. The tension is set when bags are installed. Typical values are about 4 Hz for frequency and 2 to 3 inches for amplitude (half-stroke).[4] Some installations allow easy adjustment of bag tension, while others require that the bag be loosened and reclamped to its attaching thimble.

Compared with reverse-air cleaned bags (discussed below) the vigorous action of shaker systems tends to stress the bags more, which requires heavier and more durable fabrics. In the United States, woven fabrics are used almost exclusively for shaker cleaning.[5] European practice allows the use of felted fabrics at somewhat higher filtering velocities. These higher velocities allow construction of a smaller baghouse, which requires less capital. However, the higher velocities lead to higher pressure drop, which increases operating costs. For any given application, an economic balance exists that must often be found by estimating costs for both types of fabric. Significant research has been done with shaker baghouses and the woven fabrics used in them, and many shaker baghouses remain in service. However, the majority of newly erected baghouses are pulse jets. Where baghouses larger than typical pulse jets are required, they are often custom-built, reverse-air units. The pulse-jet baghouses have become popular because they occupy less space than the equivalent shaker baghouse and are perceived as being less expensive. For high-temperature applications using glass bags, longer bag life may be expected than would be found with shaker baghouses.

1.2.2 Reverse-air Cleaning

When glass fiber fabrics were introduced, a gentler means of cleaning the bags, which may be a foot in diameter and 30 feet in length, was needed to prevent premature degradation. Reverse-air cleaning was developed as a less intensive way to impart energy to the bags. In reverse-air cleaning, gas flow to the bags is stopped in the compartment being cleaned and reverse (outside-in) air flow is directed through the bags. This reversal of gas flow gently collapses the bags toward their centerlines, which causes the cake to detach from the fabric surface. The detachment is caused by shear forces developed between the dust and fabric as the latter changes its shape. Metal caps to support the bag tops are an integral part of the bag as are several sewn-in rings that encircle the bags to prevent their complete collapse during cleaning. Without these rings, falling collected dust tends to choke the bag as the fabric collapses in on itself while cleaning. As with multi-compartment shaker baghouses, a similar cycle takes place in reverse-air baghouses of stopping forward gas flow and allowing dust to settle before cleaning action begins. Also, as with shaker baghouses, extra filtering capacity must be added to reverse-air baghouses to compensate for that portion out of service for cleaning at any time. Some reverse-air baghouses employ a supplemental shaker system to assist cleaning by increasing the amount of energy delivered to the bag.

The source of reverse air is generally a separate system fan capable of supplying clean, dry air for one or two compartments at a gas-to-cloth ratio as high or higher than that of the forward gas flow. Figure 1.2 illustrates a reverse-air cleaned baghouse.

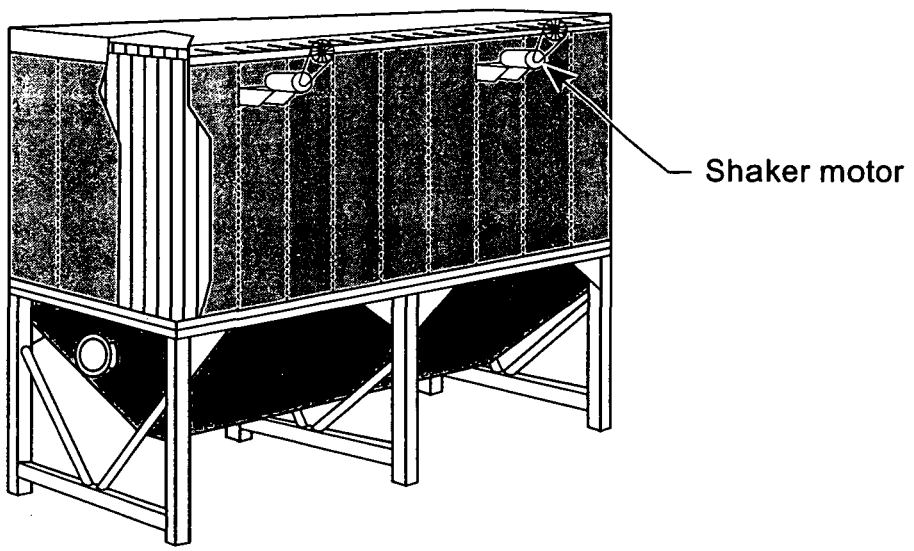


Figure 1.1: Typical Shaker Baghouse
(Courtesy of North Carolina State University)

1.2.3 Pulse-jet Cleaning

An advantage of pulse-jet cleaning compared to shaker or reverse-air baghouses is the reduction in baghouse size (and capital cost) allowed by using less fabric because of higher gas-to-cloth ratios and, in some cases, by not having to build an extra compartment for off-line cleaning. However, the higher gas-to-cloth ratios cause higher pressure drops that increase operating costs. This form of cleaning uses compressed air to force a burst of air down through the bag and expand it violently. As with shaker baghouses, the fabric reaches its extension limit and the dust separates from the bag. Air escaping through the bag carries the separated dust away from the fabric surface. In pulse jets, however, filtering gas flows are opposite in direction when compared with shaker or reverse-air baghouses (i.e., outside-in). Figure 1.3 illustrates a pulse-jet cleaned baghouse.

1.2.3.1 Caged Filters

In conventional pulse-jet baghouses, bags are mounted on wire cages to prevent collapse while the dusty gas flows from outside the bag to the inside during filtration. Instead of attaching both ends of the bag to the baghouse structure, the bag and cage assembly generally is attached only at the top. The bottom end of the assembly tends to move in the turbulent gas flow during filtration and may rub other bags, which accelerates wear.

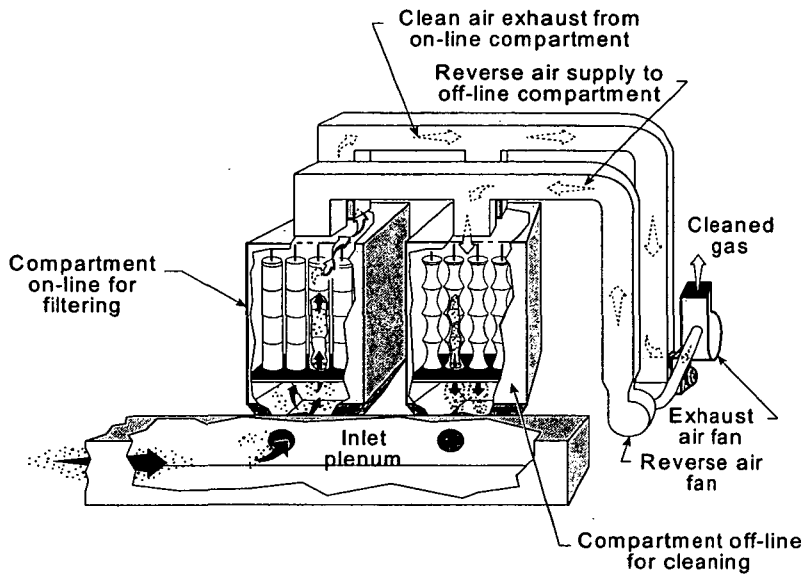


Figure 1.2: Typical Reverse-Air Baghouse
(Courtesy of North Carolina State University)

Often, pulse-jet baghouses are not compartmented. Bags are cleaned one row at a time when a timer initiates the burst of cleaning air through a quick-opening valve. A pipe across each row of bags carries the compressed air. The pipe has a nozzle above each bag so that cleaning air exits directly into the bag. Some systems direct the air through a short venturi that is intended to entrain additional cleaning air. The pulse opposes and interrupts forward gas flow for only a few tenths of a second. However, the quick resumption of forward flow redeposits most of the dust back on the clean bag or on adjacent bags. This action has the disadvantage of inhibiting dust from dropping into the hopper, but the advantage of quickly reforming the dust cake that provides efficient particle collection.

To increase filter area in the same volume of baghouse, star-shaped and pleated (in cross section) bag/cage configurations have been developed. The bag/cage combination is designed as a unit to be installed similarly to a standard bag and cage unit. Such units can be used as replacements for standard bags and cages when additional fabric area is needed, or may be used in original designs. Normal pulse cleaning is used, i.e., no special changes to the cleaning equipment are required. Costs for star-shaped bags and cages are about three to three-and-a-half times normal bags and cages.

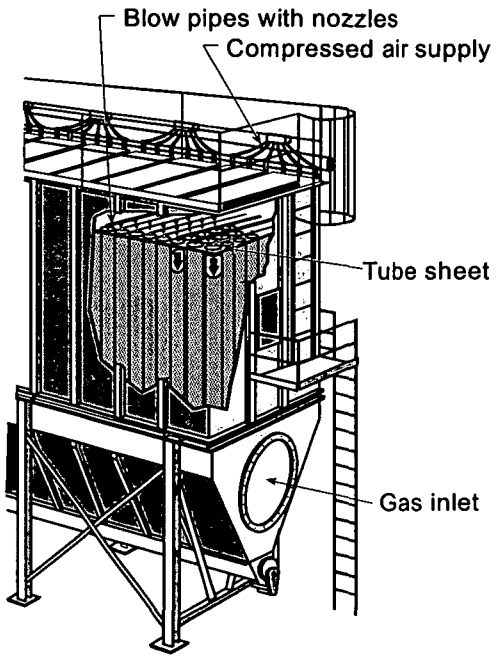


Figure 1.3: Typical Pulse-Jet Baghouse
(Courtesy of North Carolina State University)

1.2.3.2 Cartridge Filters

Further increases in filter area per unit of baghouse volume are obtained by using finely pleated filter media supported on a wire framework. This cartridge can be mounted vertically as a nearly direct replacement for standard bags and cages in existing baghouses, or mounted horizontally in original designs. When used as a direct replacement for standard bags and cages, retrofit costs for one case are 70 % of the cost of building a new baghouse.[6] Cleaning of early cartridge baghouse designs is by typical pulse equipment using a blow pipe across a row of cartridges. More recent designs use individual air valves for each pair of cartridges.

One type of cartridge[7] contains an inner supporting core surrounded by the pleated filter medium and outer supporting mesh. One end of the cartridge is open, which allows gas passing through the filter from the outside to exit to a clean air plenum. Cleaning air is pulsed through the same open end, but in a reverse direction from the gas being cleaned. The other end of the cartridge is closed by an end cap. The manufacturing process requires

strong, rigid joints where the end caps attach to the filter medium and cores. Epoxy or polyurethane plastics are used to seal the medium against the end caps. The cartridge is held tightly in place against a mounting plate surrounding the hole that connects it to the clean air plenum. Horizontal cartridges are typically mounted in tandem with a gasket seal between them. If not properly mounted or if the gasket material is not of high quality, leakage will occur after repeated cleaning pulses.

Filter media for cartridges may be paper, spunbonded monofilament plastics (polyester is predominant), or nonwoven fabrics. Cartridges may be from 6 in. to 14 in. in diameter and 16 in. to 36 in. in length. The filtering surface is from about 25 ft^2 to 50 ft^2 for cartridges with nonwoven fabrics, about three to four times as much with spunbondeds, and more than six times as much with paper. A typical cartridge may have 36 ft^2 of nonwoven fabric, 153 ft^2 of spunbonded fabric, or 225 ft^2 of paper. Pleat spacing is important for two reasons: closer spacing increases filter area for a specific cartridge volume, but closer spacing increases the likelihood of dust permanently bridging the bottoms of the pleats and reducing available filtering area. For nonagglomerating dusts of small particle size, (up to a few micrometers) and benign characteristics for paper, the cartridge may have 12 pleats/in. to 16 pleats/in. Nonwovens under more difficult conditions may have 4 pleats/in. to 8 pleats/in. Pleat depth is 1 in. to 3 in. Pleat arrangement and available volume of cleaning air determine the cleanability of the media for a specific dust. An advantage of paper media is their ability to collect particles less than 2.5 μm in diameter with high efficiency. Overall efficiency can be 99.999+ percent. Nonwoven media may be an order of magnitude less efficient. However, even glass fiber bags in reverse-air baghouses on combustion sources can collect 2.5 μm particles with 99.9 percent efficiency.

Cartridge filters are limited in temperature by the adhesives that seal the media to the end caps. Operating temperatures of 200 °F are common, with temperature capability to 350 °F soon to be marketed. Figure 1.4 illustrates a cartridge collector.

1.2.4 Sonic Cleaning

Because reverse-air cleaning is a low-energy method compared with shaking or pulse-jet cleaning, additional energy may be required to obtain adequate dust removal. Shaking, as described above, is one such means of adding energy, but another is adding vibrational energy in the low end of the acoustic spectrum. Sonic horns powered by compressed air are a typical means of applying this energy. The horns (1 to several per compartment for large baghouses) typically operate in the range of 125 to 550 Hz (more frequently in the 125 to 160 Hz range) and produce sound pressures of 120 to 145 db. When properly applied, sonic energy can reduce the mass of dust on bags considerably, but may also lead to increased dust

penetration through the fabric. Increased penetration reduces the efficiency of the baghouse. Sonic horns are effective as supplemental equipment for some applications that require added energy for adequate cleaning, Occasionally sonic horns are used as the only source of cleaning energy.

Horn construction includes a horn-shaped outlet attached to an inlet chamber containing a diaphragm. Compressed air at 45 to 75 psig enters the chamber, vibrates the diaphragm, and escapes through the horn. Sound waves leaving the horn contact and vibrate dust-containing fabric with sufficient energy to loosen or detach patches of dust that fall through the bag to the hopper below. Compressed air consumption varies from 45 to 75 scfm depending on the size of the horn. Horns can be flange mounted through the baghouse siding with the flange at either the outlet end of the horn or at the inlet chamber. The horns also can be suspended inside the baghouse structure.

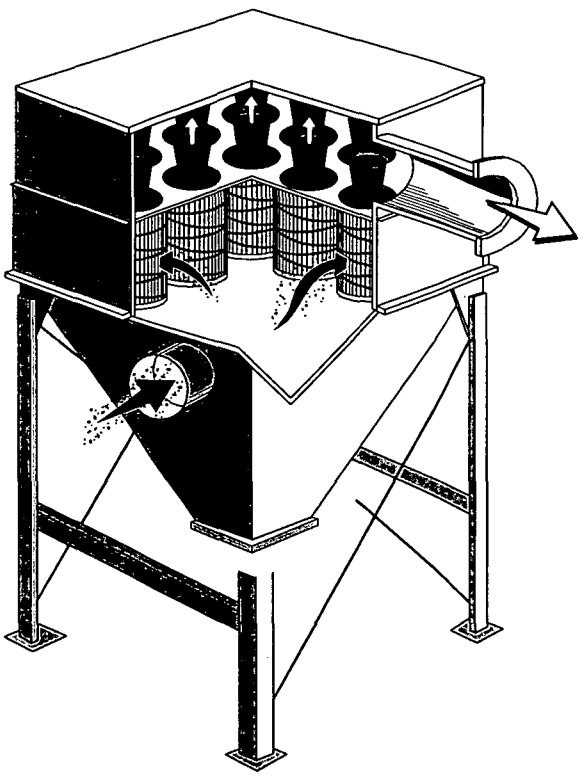


Figure 1.4: Typical Vertical-Mount Cartridge Baghouse
(Courtesy of North Carolina State University)

An example of sonic horn usage is a 10-compartment, reverse-air baghouse cleaning combustion gases at 835,000 acfm. Bags being cleaned are 12 in. in diameter and 35 ft in length. Each compartment has a horn mounted in each of the four corners and angled towards the center of the compartment. Compartments are cleaned every 30 minutes with reverse air for 1 minute and sonic horns for 30 seconds during the reverse-air cleaning. The horns operate at 75 psig and consume 65 scfm of compressed air. For baghouses requiring less intensive cleaning, the cleaning cycle might be extended to 1 hour or more.

For a 6-compartment baghouse requiring 1 horn per compartment, the system investment for horns was $13,500 (the BHA Group). The installed horns operated at 125 Hz and used 75 scfm of compressed air at 75 psig. In this case, each horn cleaned 8,500 ft^2 of fabric. The same size horn can clean up to 15,000 ft^2 of fabric.

1.2.5 Auxiliary Equipment

The typical auxiliary equipment associated with fabric filter systems is shown in Figure 1.5. Along with the fabric filter itself, a control system typically includes the following auxiliary equipment: a capture device (i.e., hood or direct exhaust connection); ductwork; dust removal equipment (screw conveyor, etc.); fans, motors, and starters; and a stack. In addition, spray chambers, mechanical collectors, and dilution air ports may be needed to precondition the gas before it reaches the fabric filter. Capture devices are usually hoods or direct exhaust couplings attached to a process vessel. Direct exhaust couplings are less common, requiring sweep air to be drawn through the process vessel, and may not be feasible in some processes. Ductwork (including dampers) is used to contain, and regulate the flow of, the exhaust stream as it moves from the emission source to the control device and stack. Spray chambers and dilution air ports decrease the temperature of the pollutant stream to protect the filter fabric from excessive temperatures. When a substantial portion of the pollutant loading consists of relatively large particles (more than about 20 μm), mechanical collectors such as cyclones are used to reduce the load on the fabric filter. Fans provide motive power for air movement and can be mounted before (pressure baghouse) or after (suction baghouse) the filter. Stacks, when used, vent the cleaned stream to the atmosphere. Screw conveyors are often used to remove captured dust from the bottom of the hoppers under the fabric filter and (if used) mechanical collector. Air conveying (pneumatic) systems and direct dumping into containers are also used as alternate means for dust removal from the hoppers.

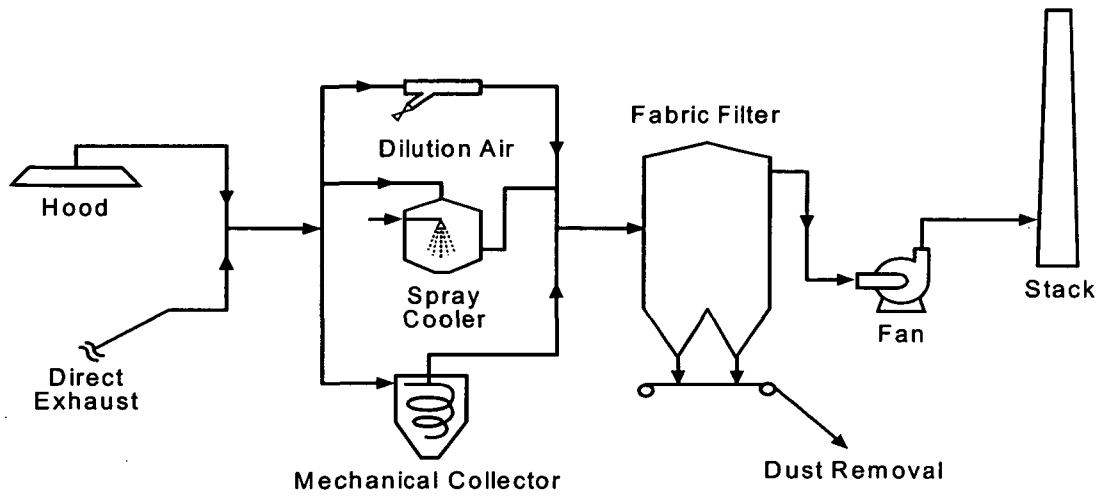


Figure 1.5: Typical alternative auxiliary equipment items used with fabric filter control systems.

1.2.6 Fabric Filtration Theory

The key to designing a baghouse is to determine the face velocity that produces the optimum balance between pressure drop (operating cost that increases as pressure drop increases) and baghouse size (capital cost that decreases as the baghouse size is reduced). Baghouse size is reduced as the face velocity (or gas-to-cloth ratio) is increased. However, higher gas-to-cloth ratios cause higher pressure drops. Major factors that affect design gas-to-cloth ratio, discussed in Section 1.3, include particle and fabric characteristics and gas temperature.

Although collection efficiency is another important measure of baghouse performance, a properly designed and well run baghouse will generally have an extemely high particulate matter (PM) collection efficiency (i.e., 99.9+ percent). Baghouses are particularly effective for collecting small particles. For example, tests of baghouses on two utility boilers[8],[9] showed efficiencies of 99.8 percent for particles 10 μm in diameter and 99.6 percent to 99.9 percent for particles 2.5 μm in diameter. Because high efficiency is assumed, the design process focuses on the pressure drop.

Pressure drop occurs from the flow through inlet and outlet ducts, from flow through the hopper regions, and from flow through the bags. The pressure drop through the baghouse compartment (excluding the pressure drop across the bags) depends largely on the baghouse design and ranges from 1 to 2 inches of H_2O[3] in conventional designs and up to about 3 inches of H_2O in designs having complicated gas flow paths. This loss can be kept to a minimum

(i.e., 1 inch of H_2O or less) by investing in a flow modeling study of the proposed design and modifying the design in accordance with the study results. A study of this sort would cost on the order of $70,000 (in 1998).

The pressure drop across the bags (also called the tube-sheet pressure drop) can be as high as 10 inches of H_2O or more. The tube-sheet pressure drop is a complex function of the physical properties of the dust and the fabric and the manner in which the baghouse is designed and operated. The duct and hopper losses for a specific configuration are constant and can be minimized effectively by changing the configuration through proper design based on a knowledge of the flow through the baghouse.[1]

Fabric filtration is a batch process that has been adapted to continuous operation. One requirement for a continuously operating baghouse is that the dust collected on the bags must be removed periodically. Shaker and reverse-air baghouses normally use woven fabric bags, run at relatively low face velocities, and have cake filtration as the major particle removal mechanism. That is, the fabric merely serves as a substrate for the formation of a dust cake that is the actual filtration medium. Pulse-jet baghouses generally use felt fabric and run with a high gas-to-cloth ratio (about double that of shaker or reverse-air baghouses). The felt fabric may play a much more active role in the filtration process. This distinction between cake filtration and fabric filtration has important implications for the rate of pressure loss across the filter bags. The theoretical description and design process for cake filtration is quite different from that for fabric filtration. Fabric selection is aided by bench-scale filtration tests to investigate fabric effects on pressure drop, cake release during cleaning, and collection efficiency. These tests cost less than one-tenth the cost of flow modeling. Electrical properties of the fabric , such as resistivity and triboelectric order (the fabric's position in a series from highly electropositive to highly electronegative as determined from its charge under a specific triboelectrification procedure), may be measured to aid in fabric selection. Although their effects are generally poorly understood, electrical/ electrostatic effects influence cake porosity and particle adhesion to fabrics or other particles.[10][11][12] Knowledge of the effects can lead to selection of fabrics that interact favorably regarding dust collection and cleaning.

The following sections display the general equations used to size a baghouse, beginning with the reverse air/shake deflate type of baghouse.

[1]A procedure for estimating duct pressure losses is given in Section 2 ("Hoods, Ductwork, and Stacks") of this Manual.

1.2.6.1 Reverse Air/Shake Deflate Baghouses

The construction of a baghouse begins with a set of specifications including average pressure drop, total gas flow, and other requirements; a maximum pressure drop may also be specified. Given these specifications, the designer must determine the maximum face velocity that can meet these requirements. The standard way to relate baghouse pressure drop to face velocity is given by the relation:

$$\Delta P(\theta) = S_{sys}(\theta) V_{f\ (avg.)} \tag{1.1}$$

where

$\Delta P(\theta)$ = the pressure drop across the filter, a function of time, θ(in. H_2O)
$S_{sys}(\theta)$ = system drag, a function of time [in. H_2O/(ft/min)]
$V_{f\ (avg.)}$ = average (i.e., design) face velocity or G/C, constant (ft/min)

For a multi-compartment baghouse, the system drag, which accounts for most of the drag from the inlet flange to the outlet flange of the baghouse, is determined as a combination of resistances representative of several compartments. For the typical case where the pressure drop through each compartment is the same, and where the filtering area per compartment is equal, it can be shown that:[13]

$$S_{sys}(\theta) = \left[\frac{1}{M}\sum_{i=1}^{M}\frac{1}{S_i(\theta)}\right]^{-1} = \frac{1}{\frac{1}{M}\sum_{i=1}^{M}\frac{1}{S_i(\theta)}} = \frac{M}{\sum_{i=1}^{M}\frac{1}{S_i(\theta)}} \tag{1.2}$$

where

M = number of compartments in the baghouse
$S_i(\theta)$ = drag across compartment i

The compartment drag is a function of the amount of dust collected on the bags in that compartment. Dust load varies nonuniformly from one bag to the next, and within a given bag there will also be a variation of dust load from one area to another. For a sufficiently small area, j, within compartment i, it can be assumed that the drag is a linear function of dust load:

$$S_{i,j}(\theta) = S_e + K_2\, W_{i,j}(\theta) \tag{1.3}$$

where

S_e = drag of a dust-free filter bag [in. H_2O/(ft/min)]
K_2 = dust cake flow resistance {[in. H_2O/(ft/min)]/(lb/ft^2)}
$W_{i,j}(\theta)$ = dust mass per unit area of area j in compartment i, "areal density" (lb/ft^2)

If there are N different areas of equal size within compartment i, each with a different drag $S_{i,j}$ then the total drag for compartment i can be computed in a manner analogous to Equation 1.2:

$$S_i(\theta) = \frac{N}{\sum \frac{1}{S_{i,j}(\theta)}} \tag{1.4}$$

The constants S_e and K_2 depend upon the fabric and the nature and size of the dust. The relationships between these constants and the dust and fabric properties are not understood well enough to permit accurate predictions and so must be determined empirically, either from prior experience with the dust/fabric combination or from laboratory measurements. The dust mass as a function of time is defined as:

$$W_{i,j}(\theta) = W_r + \int_0^{\theta} C_{in} V_{i,j}(\theta) d\theta \tag{1.5}$$

where

W_r = dust mass per unit area remaining on a "clean" bag (lb/ft^2)
C_{in} = dust concentration in the inlet gas (lb/ft^3)
$V_{i,j}(\theta)$ = face velocity through area j of compartment i (ft/min)

The inlet dust concentration and the filter area are assumed constant. The face velocity, (gas-to-cloth ratio) through each filter area j and compartment i changes with time, starting at a maximum value just after clearing and steadily decreasing as dust builds up on the bags. The individual compartment face velocities are related to the average face velocity by the expression:

$$V_{avg} = \frac{\sum_i \sum_j V_{i,j}(\theta) A_{i,j}}{\sum_i \sum_j A_{i,j}} \tag{1.6}$$

$$= \frac{\sum_i \sum_j V_{i,j}(\theta)}{M}$$

(for M compartments with equal area)

Equations 1.1 through 1.6 reveal that there is no explicit relationship between the design face velocity and the tube-sheet pressure drop. The pressure drop for a given design can only be determined by the simultaneous solution of Equations 1.1 through 1.5, with Equation 1.6 as a constraint on that solution. Solving the equations requires an iterative procedure: begin with a known target for the average pressure drop, propose a baghouse design (number of compartments, length of filtration period, etc.), assume a face velocity that will yield that pressure drop, and solve the system of Equations 1.1 through 1.6 to verify that the calculated pressure drop equals the target pressure drop. If not, repeat the procedure with new parameters until the specified face velocity yields an average pressure drop (and maximum pressure drop, if applicable) that is sufficiently close to the design specification. Examples of the iteration procedure's use are given in reference [13].

1.2.6.2 Pulse-Jet Baghouses

The distinction between pulse-jet baghouses using felts and reverse-air and shaker baghouses using woven fabrics is basically the difference between cake filtration and composite dust/fabric filtration (noncake filtration). This distinction is more a matter of convenience than physics, as either type of baghouse can be designed for a specific application. However, costs for the two types will differ depending on application- and size-specific factors. Some pulse jets remain on-line at all times and are cleaned frequently. Others are taken off-line for cleaning at relatively long intervals. The longer a compartment remains on-line without cleaning, the more its composite dust/fabric filtration mechanism changes to cake filtration. Therefore, a complete model of pulse-jet filtration must account for the depth filtration occurring on a relatively clean pulse-jet filter, the cake filtration that inevitably results from prolonged periods on-line, and the transition period between the two regimes. When membrane fabrics are used, filtration takes place primarily at the surface of the membrane, which acts similarly to a cake. The following analysis has not been tested against membrane fabrics.

Besides the question of filtration mechanism, there is also the question of cleaning method. If the conditions of an application require that a compartment be taken off-line for cleaning, the dust removed falls into the dust hopper before forward gas flow resumes. If conditions allow a compartment to be cleaned while on-line, only a small fraction of the dust removed from the bag falls into the hopper. The remainder of the dislodged dust will be redeposited (i.e., "recycled") on the bag by the forward gas flow. The redeposited dust layer

has different pressure drop characteristics than the freshly deposited dust. The modeling work that has been done to date focuses on the on-line cleaning method. Dennis and Klemm[14] proposed the following model of drag across a pulse-jet filter:

$$S = S_e\left(K_2\right)_c W_c + K_2 W_o \tag{1.7}$$

where

S	=	drag across the filter
S_e	=	drag of a just-cleaned filter
$(K_2)_c$	=	specific dust resistance of the recycling dust
W_c	=	areal density of the recycling dust
K_2	=	specific dust resistance of the freshly deposited dust
W_o	=	areal density of the freshly deposited dust

This model has the advantage that it can easily account for all three regimes of filtration in a pulse-jet baghouse. As in Equations 1.1 to 1.6, the drag, filtration velocity and areal densities are functions of time, θ. For given operating conditions, however, the values of S_e, $(K_2)_c$, and W_c may be assumed to be constant, so that they can be grouped together:

$$\Delta P = (PE)_{\Delta w} + K_2 W_o V_f \tag{1.8}$$

where

ΔP	=	pressure drop (in. H_2O)
V_f	=	filtration velocity (ft/min)
$(PE)_{\Delta w}$	=	$[S_e + (K_2)_c W_c] V_f$

Equation 1.8 describes the pressure drop behavior of an individual bag. To extend this single bag result to a multiple-bag compartment, Equation 1.7 would be used to determine the individual bag drag and total baghouse drag would then be computed as the sum of the parallel resistances. Pressure drop would be calculated as in Equation 1.1. It seems reasonable to extend this analysis to the case when the dust is distributed unevenly on the bag and then apply Equation 1.7 to each area on the bag, followed by an equation analogous to 1.4 to compute the overall bag drag. The difficulty in following this procedure is that one must assume values for W_c for each different area to be modeled.

The disadvantage of the model represented by Equations 1.7 and 1.8 is that the constants, S_e, $(K_2)_c$, and W_c, cannot be predicted at this time. Consequently, correlations of laboratory data must be used to determine the value of $(PE)_{\Delta w}$. For the fabric-dust combination of Dacron felt and coal fly ash, Dennis and Klemm[14] developed an empirical relationship between $(PE)_{\Delta w}$, the face velocity, and the cleaning pulse pressure. This relationship (converted from metric to English units) is as follows:

$$(PE)_{\Delta w} = 6.08 V_f P_j^{-0.65} \qquad (1.9)$$

where

V_f = face velocity, (ft/min)
P_j = pressure of the cleaning pulse (usually 60 to 100 psig; see Section 5.4.1)

This equation is essentially a regression fit to a limited amount of laboratory data and should not be applied to other dust/fabric combinations. The power law form of Equation 1.9 may not be valid for other dusts or fabrics. Consequently, more data should be collected and analyzed before the model represented by Equation 1.9 can be used for rigorous sizing purposes.

Another model that shows promise in the prediction of noncake filtration pressure drop is that of Leith and Ellenbecker[15] as modified by Koehler and Leith.[16] In this model, the tube-sheet pressure drop is a function of the clean fabric drag, the system hardware, and the cleaning energy. Specifically:

$$\Delta P = \frac{1}{2}\left[P_s + K_1 V_f - \sqrt{\left(P_s - K_1 V_f\right)^2 - 4 W_o \frac{K_2}{K_3}} \right] + K_v V_f^2 \qquad (1.10)$$

where

P_s = maximum static pressure achieved in the bag during cleaning
K_1 = clean fabric resistance
V_f = face velocity
K_2 = dust deposit flow resistance
K_3 = bag cleaning efficiency coefficient
K_v = loss coefficient for the venturi at the inlet to the bag

Comparisons of laboratory data with pressure drops computed from Equation 1.10 [15,16] are in close agreement for a variety of dust/fabric combinations. The disadvantage of Equation 1.10 is that the constants K_1, K_2, and K_3 must be determined from laboratory measurements. The most difficult one to mine is the K_3 value, which can only be found by making measurements in a pilot-scale pulse-jet baghouse. A limitation of laboratory measurements is that actual filtration conditions cannot always be adequately simulated. For example, a redispersed dust may not have the same size distribution or charge characteristics as the original dust, thereby yielding different values of K_1, K_2, and K_3 than would be measured in an operating baghouse.

1.3 Design Procedures

The design procedure requires estimating a gas-to-cloth ratio that is compatible with fabric selection and cleaning type. Fabric selection for composition depends on gas and dust characteristics; fabric selection for construction (woven or felt) largely depends on type of cleaning. Estimating a gas-to-cloth ratio that is too high, compared to a correctly estimated gas-to-cloth ratio, leads to higher pressure drops, higher particle penetration (lower collection efficiency), and more frequent cleaning that leads to reduced fabric life. Estimating a gas-to-cloth ratio that is too low increases the size and cost of the baghouse unnecessarily. Each of the parameters for design is discussed below.

1.3.1 Gas-to-Cloth Ratio

The gas-to-cloth ratio is difficult to estimate from first principles. However, shortcut methods of varying complexity allow rapid estimation. Three methods of increasing difficulty follow. For shaker and reverse-air baghouses, the third method is best performed with publicly available computer programs. Although pulse-jet baghouses have taken a large share of the market, they are not necessarily the least costly type for a specific application. Costing should be done for pulse-jet baghouses at their application-specific gas-to-cloth ratios and for reverse-air or shaker baghouses at their application-specific gas-to-cloth ratios.

The methods outlined below pertain to conventional baghouses. Use of electrostatic stimulation may allow a higher gas-to-cloth ratio at a given pressure drop; thus a smaller baghouse structure and fewer bags are needed. Viner and Locke[17] discuss cost and performance models for electrostatically stimulated fabric filters; however, no data are available for full-scale installations. Use of extended area bag configurations (star-shaped bags or pleated media cartridges) do not allow significant changes in gas-to-cloth ratios, but do allow installation of more fabric in a given volume.

1.3.1.1 Gas-to-Cloth Ratio From Similar Applications

After a fabric has been selected, an initial gas-to-cloth ratio can be determined using Table 1.1. Column 1 shows the type of dust; column 2 shows the gas-to-cloth ratios for woven fabric; and column 3 shows gas-to-cloth ratios for felted fabrics. Notice that these values are all "net" gas-to-cloth ratios, equal to the total actual volumetric flow rate in cubic feet per minute divided by the net cloth area in square feet. This ratio, in units of feet per minute, affects pressure drop and bag life as discussed in Section 1.2. The net cloth area is determined by dividing the exhaust gas flow rate in actual cubic feet per minute (acfm) by the design gas-to-cloth ratio. For an intermittent-type baghouse that is shut down for cleaning, the net cloth area is also the total, or gross, cloth area. However, for continuously operated shaker and reverse-air filters,

the area must be increased to allow the shutting down of one or more compartments for cleaning. Continuously operated, compartmented pulse-jet filters that are cleaned off line also require additional cloth to maintain the required net area when cleaning. Table 1.2 provides a guide for adjusting the net area to the gross area, which determines the size of a filter requiring off-line cleaning.

1.3.1.2 Gas-to-Cloth Ratio From Manufacturer's Methods

Manufacturers have developed nomographs and charts that allow rapid estimation of the gas-to-cloth ratio. Two examples are given below, one for shaker-cleaned baghouses and the other for pulse-jet cleaned baghouses.

For shaker baghouses, Table 1.3 gives a factor method for estimating the ratio. Ratios for several materials in different operations are presented, but are modified by factors for particle size and dust load. Directions and an example are included. Gas-to-cloth ratios for reverse-air baghouses would be about the same or a little lower compared to the Table 1.3 values.

Table 1.1: Gas-to-Cloth Ratios for Baghouse/Fabric Combinations[a,b]
(actual ft^3/min)/(ft^2 of net cloth area)

Dust	Shaker/Woven Fabric Reverse-Air/Woven Fabric	Pulse Jet/Felt Fabric Reverse-Air/Felt Fabric
Alumina	2.5	8
Asbestos	3.0	10
Bauxite	2.5	8
Carbon Black	1.5	5
Coal	2.5	8
Cocoa, Chocolate	2.8	12
Clay	2.5	9
Cement	2.0	8
Cosmetics	1.5	10
Enamel Frit	2.5	9
Feeds, Grain	3.5	14
Feldspar	2.2	9
Fertilizer	3.0	8
Flour	3.0	12
Fly Ash	2.5	5
Graphite	2.0	5
Gypsum	2.0	10
Iron Ore	3.0	11
Iron Oxide	2.5	7
Iron Sulfate	2.0	6
Lead Oxide	2.0	6
Leather Dust	3.5	12
Lime	2.5	10
Limestone	2.7	8
Mica	2.7	9
Paint Pigments	2.5	7
Paper	3.5	10
Plastics	2.5	7
Quartz	2.8	9
Rock Dust	3.0	9
Sand	2.5	10
Sawdust (Wood)	3.5	12
Silica	2.5	7
Slate	3.5	12
Soap, Detergents	2.0	5
Spices	2.7	10
Starch	3.0	8
Sugar	2.0	13
Talc	2.5	5
Tobacco	3.5	
Zinc Oxide	2.0	

[a]Reference[18]

[b]Generally safe design values; application requires consideration of particle size and grain loading.

Table 1.2: Approximate Guide to Estimate Gross Cloth Area From Net Cloth Area[a]

Net Cloth Area (ft^2)	Multiplier to Obtain Gross Cloth Area (ft^2)	
1-4,000	Multiply by	2
4,001-12,000	"	1.5
12,001-24,000	"	1.25
24,001-36,000	"	1.17
36,001-48,000	"	1.125
48,001-60,000	"	1.11
60,001-72,000	"	1.10
72,001-84,000	"	1.09
84,001-96,000	"	1.08
96,001-108,000	"	1.07
108,001-132,000	"	1.06
132,001-180,000	"	1.05
above 180,001	"	1.04

[a]Reference[19]

For pulse-jet baghouses, which normally operate at two or more times the gas-to-cloth ratio of reverse-air baghouses, another factor method[20] has been modified with equations to represent temperature, particle size, and dust load:

$$V = 2.878\ A\ B\ T^{-0.2335} L^{-0.06021} (0.7471 + 0.0853 \ln D) \qquad (1.11)$$

where

V = gas-to-cloth ratio (ft/min)
A = material factor, from Table 5.4
B = application factor, from Table 5.4
T = temperature, (°F, between 50 and 275)
L = inlet dust loading (gr/ft^3, between 0.05 and 100)
D = mass mean diameter of particle (μm, between 3 and 100)

For temperatures below 50°F, use $T = 50$ but expect decreased accuracy; for temperatures above 275°F, use $T = 275$. For particle mass mean diameters less than 3 μm, the value of D is 0.8, and for diameters greater than 100 μm, D is 1.2. For dust loading less than 0.05 gr/ft^3, use $L = 0.05$; for dust loading above 100 gr/ft^3, use $L = 100$. For horizontal cartridge baghouses, a similar factor method can be used. Table 1.5 provides the factors.

Table 1.3: Manufacturer's Factor Method for Estimating Gas-to-cloth Ratios for Shaker Baghouses

A — 4/1 RATIO

Material	Operation
Cardboard	1
Feeds	2, 3, 4, 5, 6, 7
Flour	2, 3, 4, 5, 6, 7
Grain	2, 3, 4, 5, 6, 7
Leather Dust	1, 7, 8
Tobacco	1, 4, 6, 7
Supply Air	13
Wood, Dust, Chips	1, 6, 7

3/1 RATIO

Material	Operation
Asbestos	1, 7, 8
Aluminum Dust	1, 7, 8
Fibrous Mat'l	1, 4, 7, 8
Cellulose Mat'l	1, 4, 7, 8
Gypsum	1, 3, 5, 6, 7
Lime (Hydrated)	2, 4, 6, 7
Perlite	2, 4, 5, 6
Rubber Chem.	4, 5, 6, 7, 8
Salt	2, 3, 4, 5, 6, 7
Sand*	4, 5, 6, 7, 9, 15
Iron Scale	1, 7, 8
Soda Ash	4, 6, 7
Talc	3, 4, 5, 6, 7
Machining Operation	1, 8

2.5/1 RATIO

Material	Operation
Alumina	2, 3, 4, 5, 6
Carbon Black	4, 5, 6, 7
Cement	3, 4, 5, 6, 7
Coke	2, 3, 5, 6
Ceramic Pigm. Clay and Brick Dust	4, 5, 6, 7
	2, 4, 6, 12
Coal	2, 3, 6, 7, 12
Kaolin	4, 5, 7
Limestone	2, 3, 4, 5, 6, 7
Rock, Ore Dust	2, 3, 4, 5, 6, 7
Silica	2, 3, 4, 5, 6, 7
Sugar	3, 4, 5, 6, 7

2/1 RATIO

Material	Operation
Ammonium Phosphate Fertilizer	2, 3, 4, 5, 6, 7
Diatomaceous Earth	4, 5, 6, 7
Dry Petrochem.	2, 3, 4, 5, 6, 7, 14
Dyes	2, 3, 4, 5, 6, 7
Fly Ash	10
Metal Powders	2, 3, 4, 5, 6, 7, 14
Plastics	2, 3, 4, 5, 6, 7, 14
Resins	2, 3, 4, 5, 6, 7, 14
Silicates	2, 3, 4, 5, 6, 7, 14
Starch	6, 7
Soaps	3, 4, 5, 6, 7

1.5/1 RATIO

Material	Operation
Activated Carbon	2, 4, 5, 6, 7
Carbon Black	11, 14
Detergents	2, 4, 5, 6, 7
Metal Fumes, Oxides and other Solid Dispersed Products	10, 11

CUTTING - 1
CRUSHING - 2
PULVERIZING - 3
MIXING - 4
SCREENING - 5
STORAGE - 6
CONVEYING - 7
GRINDING - 8
SHAKEOUT - 9
FURNACE FUME - 10
REACTION FUME - 11
DUMPING - 12
INTAKE CLEANING - 13
PROCESS - 14
BLASTING - 15

B — FINENESS FACTOR

Micron Size	Factor
> 100	1.2
50 - 100	1.1
10 - 50	1.0
3 - 10	0.9
1 - 3	0.8
< 1	0.7

C — DUST LOAD FACTOR

Loading gr/cu ft	Factor
1 - 3	1.2
4 - 8	1.0
9 - 17	0.95
18 - 40	0.90
> 40	0.85

This information constitutes a guide for commonly encountered situations and should not be considered a "hard-and-fast" rule. Air-to-cloth ratios are dependent on dust loading, size distribution, particle shape and "cohesiveness" of the deposited dust. These conditions must be evaluated for each application. The larger the interval between bag cleaning the lower the air-to-cloth ratio must be. Finely-divided, uniformly sized particles generally form more dense filter cakes and require lower air-to-cloth ratios than when larger particles are interspersed with the fines. Sticky, oily particles, regardless of shape and size, form dense filter cakes and require lower air-to-cloth ratios.

Example: Foundry shakeout unit handling 26,000 CFM and collecting 3,500 lb/hr of sand. The particle distribution shows 90% greater than 10 microns. The air is to exhaust to room in winter, to atmosphere in summer.

$$3{,}500\frac{lb}{hr} + 60\frac{\text{min}}{hr} + 26{,}000\frac{ft^3}{\text{min}} \times 7{,}000\frac{gr}{lb} = 15.7\frac{gr}{ft^3}$$

*Chart A = 3/1 ratio, Chart B = Factor 1.0, Chart C = 0.95; 3 x 1 x 0.95 = 2.9 air-to-cloth ratio. 26,000 / 2.9 = 9,000 sq. ft.

Reprinted with permission from Buffalo Forge Company Bulletin AHD-29

Table 1.4: Factors for Pulse-Jet Gas-to-Cloth Ratios[a]

A. Material Factor

15[b]	12	10	9.0	6.0[c]
Cake mix	Asbestos	Alumina	Ammonium phosphate-fertilizer	Activated carbon
Cardboard dust	Buffing dust	Aspirin	Cake	Carbon black (molecular)
Cocoa	Fiborous and cellulosic material	Carbon black (finished)	Diatomaceous earth	Detergents
Feeds	Foundary shakeout	Cement	Dry petro-chemicals	Fumes and other dispersed products direct from reactions
Flour	Gypsum	Ceramic pigments	Dyes	Powdered milk
Grain	Lime (hydrated)	Clay and brick dusts	Fly ash	Soap
Leather dust	Perlite	Coal	Metal powder	
Sawdust	Rubber chemicals	Fluorspar	Metal oxides	
Tobacco	Salt	Gum, natural	Pigments metallic end synthetic	
	Sand	Kaolin	Plastics	
	Sandblast dust	Limestone	Resins	
	Soda ash	Perchlorates	Silicates	
	Talc	Rock dust, ores and minerals	Starch	
		Silica	Stearates	
		Sorbic acid	Tannic acid	
		Sugar		

B. Application Factor

Nuisance Venting Relief of transfer points, conveyors, packing stations, etc.	1.0
Product Collection Air conveying-venting, mills, flash driers, classifiers, etc.	0.9
Process Gas Filtration Spray driers, kilns, reactors, etc.	0.8

[a]Reference [20]

[b]In general, physically and chemically stable material.

[c]Also includes those solids that are unstable in their physical or chemical state due to hygroscopic nature, sublimation, and/or polymerization.

1.3.1.3 Gas-to-Cloth Ratio From Theoretical/Empirical Equations

Shaker and reverse-air baghouses The system described by Equations 1.1 through 1.6 is complicated; however, numerical methods can be used to obtain an accurate solution. A critical weakness in baghouse modeling that has yet to be overcome is the lack of a fundamental description of the bag cleaning process. That is, to solve Equations 1.1 through 1.6, the value of W_r (the dust load after cleaning) must be known. Clearly, there must be a relationship between the amount and type of cleaning energy and the degree of dust removal from a bag. Dennis et al.[13] have developed correlations for the removal of coal fly ash from woven fiberglass bags by shaker cleaning and by reverse-air cleaning. These correlations have been incorporated into a computer program that generates the solution to the above system of equations.[14],[21],[22] If one were to apply the correlations developed with coal ash and woven glass fabrics to other dust/fabric combinations, the accuracy of the results would depend on how closely that dust/fabric combination mimicked the coal ash/woven glass fabric system.

Physical factors that affect the correlation include the particle size distribution, adhesion and electrostatic properties of the dust and fabric, and fabric weave, as well as cleaning energy. More research is needed in this area of fabric filtration.

The rigorous design of a baghouse thus involves several steps. First, the design goal for average pressure drop (and maximum pressure drop, if necessary) must be specified along with total gas flow rate and other parameters, such as S_e and K_2 (obtained either from field or laboratory measurements). Second, a face velocity is assumed and the number of compartments in the baghouse is computed based on the total gas flow, face velocity, bag size, and number of bags per compartment. (Typical compartments in the U.S. electric utility industry use bags 1 ft in diameter by 30 ft in length with 400 bags per compartment.) Standard practice is to design a baghouse to meet the specified pressure drop when one compartment is off-line for maintenance. The third step is to specify the operating characteristics of the baghouse (i.e., filtration period, cleaning period, and cleaning mechanism). Fourth, the designer must specify the cleaning efficiency so that the residual dust load can be estimated. Finally, the specified baghouse design is used to establish the details for Equations 1.1 through 1.6, which are then solved numerically to establish the pressure drop as a function of time. The average pressure drop is then computed by integrating the instantaneous pressure drop over the filtration cycle and dividing by the cycle time. If the computed average is higher than the design specification, the face velocity must be reduced and the procedure repeated. If the computed average pressure drop is significantly lower than the design specification, the proposed baghouse was oversized and should be made smaller by increasing the face velocity and repeating the procedure. When the computed average pressure drop comes sufficiently close to the assumed specified value, the design has been determined. A complete description of the modeling process can be found in the reports by Dennis et al.[13,22] A critique on the accuracy of the model is presented by Viner et al.[23]

Pulse-jet baghouses The overall process of designing a pulse jet baghouse is actually simpler than that required for a reverse-air or shaker baghouse if the baghouse remains on-line for cleaning. The first step is to specify the desired average tube-sheet pressure drop. Second, the operating characteristics of the baghouse must be established (e.g., on-line time, cleaning energy). Third, the designer must obtain values for the coefficients in either Equation 1.9 or Equation 1.10 from field, pilot plant, or laboratory measurements. Fourth, a value is estimated for the face velocity and the appropriate equation (Equation 1.8 or 1.10) is solved for the pressure drop as a function of time for the duration of the filtration cycle. This information is used to calculate the cycle average pressure drop. If the calculated pressure drop matches the specified pressure drop, the procedure is finished. If not, the designer must adjust the face velocity and repeat the procedure.

Table 1.5: Manufacturer's Factor Method for Estimating Gas-to-Cloth Ratio for Horizontal Cartridge Baghouses
Factor A Table for Selected Materials

	2.5	2.1	1.9	1.3	Dust Sample Required
M A T E R I A L S	Rock dust and ores Salt, Mineral[a] Sand (Not foundry)	Activated carbon Alumina (transfer) Cake Mix[a] Carbon black (finished) Ceramic pigment Coal Coke Diatomaceous earth Flour Fluorspar Fly ash Foundry shakeout Gypsum Lime, hydrated Limestone Paint, electrstatic spray (powder coating) Petrochemicals (dry) Pigments, metallic, synthetic Plaster Rubber additives Silicates Soda ash Starch Sugar[a] Welding fumes	Fertilizers[a] Talc	Alumina (air lift) Dyes Fumes, metallurgical Pigments, paint Stearates	Detergents Feeds Grains Perlite Pharmaceuticals Powdered milk Resins Soap Tobacco
			1.7	**0.7**	**Excluded dusts**
			Aspirin Cement Clay & brick dust Cocoa[a] Coffee[a] Graphite Kaolin Metal oxides Metal powder Perchlorates Selenium Silica (flour)	Silica (fume)	Asbestos Arc washing Fiberglass Fibrous and cellulosic materials Leather Metallizing Mineral Wool P.C. board grinding Paper dust Particle board Sawdust

[a] Under controlled humidity (40 %R.H.) And room temperature only.

The approximate gas-to-cloth (G/C) ratio for a Mikropul horizontal cartridge collector in acfm per square foot of filter area is obtained by multiplying the following five factors: G/C = A x B x C x D x E

For example, G/C for process gas filtration of 10 μm rock dust at 250 °F and 2 gr/acf = 2.5 x 0.8 x 0.75 x 0.9 x 1.1 = 1.49.

Courtesy of Hosokawa Mikropul

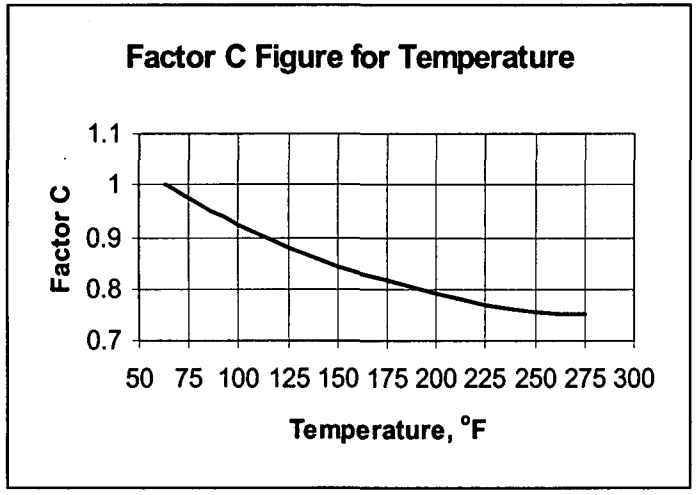
Table 1.5: (Cont.)

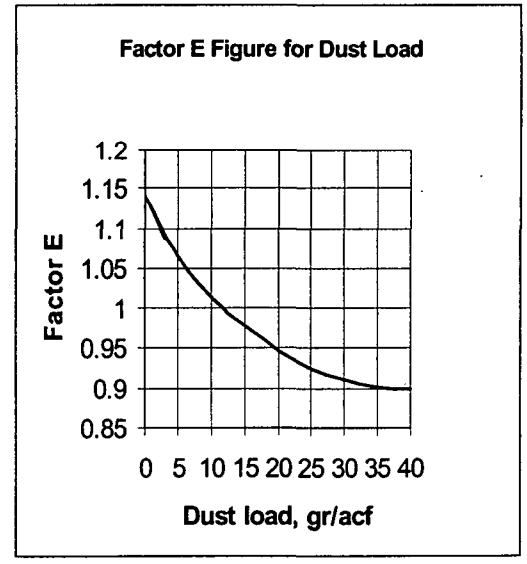
Factor B Table for Applications

Application	Factor B
Nuisance Venting Relief of transfer points, conveyors, packing stations, etc.	1.0
Product Collection Air conveying-venting, mills, flash driers, classifiers, etc.	0.9
Process Gas Filtration Spray driers, kilns, reactors, etc	0.8

Factor D Table for Dust Fineness

Fineness	Factor D
Over 50 μm	1.1
20 - 50 μm	1.0
2-20 μm	0.9
Under 2 μm	0.85





Courtesy of Hosokawa Mikropul

1.3.2 Pressure Drop

Pressure drop for the bags can be calculated from the equations given in the preceding section if values for the various parameters are known. Frequently they are not known, but a maximum pressure drop of 5 to 10 in. H_2O across the baghouse and 10 to 20 in. H_2O across the entire system can be assumed if it contains much ductwork.

A comparable form of Equations 1.1 and 1.3 that may be used for estimating the maximum pressure drop across the fabric in a shaker or reverse-air baghouse is:

$$\Delta P = S_e V + K_2 C_i V^2 \theta \qquad (1.12)$$

where

ΔP	=	pressure drop (in. H_2O)
S_e	=	effective residual drag of the fabric [in. H_2O/(ft/min)]
V	=	superficial face velocity or gas-to-cloth ratio (ft/min)
K_2	=	specific resistance coefficient of the dust {[in. H_2O/(ft/min)]/(lb /ft^2)}
C_i	=	inlet dust concentration (lb/ft^3)
θ	=	filtration time (min)

Although there is much variability, values for S_e may range from about 0.2 to 2 in. H_2O/(ft/min) and for K_2 from 1.2 to 30–40 in. H_2O/(ft/min)]/(lb/ft^2). Typical values for coal fly ash are about 1 to 4. Inlet concentrations vary from less than 0.05 gr/ft^3 to more than 100 gr/ft^3, but a more nearly typical range is from about 0.5 to 10 gr/ft^3. Filtration times may range from about 20 to 90 minutes for continuous duty baghouses, but 30 to 60 minutes is more frequently found. For pulse-jet baghouses, use Equations 1.8 and 1.9 to estimate ΔP, after substituting $C_i V \theta$ for W_o and $(PE)_{\Delta w}$ for $S_e V$.

1.3.3 Particle Characteristics

Particle size distribution and adhesiveness are the most important particle properties that affect design procedures. Smaller particle sizes can form a denser cake, which increases pressure drop. As shown in Tables 1.3 and 1.5 and Equation 1.11, the effect of decreasing average particle size is a lower applicable gas-to-cloth ratio.

Adhering particles, such as oily residues or electrostatically active plastics, may require installing equipment that injects a precoating material onto the bag surface, which acts as a buffer that traps the particles and prevents them from blinding or permanently plugging the fabric pores. Informed fabric selection may eliminate electrostatic problems.

1.3.4 Gas Stream Characteristics

Moisture and corrosives content are the major gas stream characteristics requiring design consideration. The baghouse and associated ductwork should be insulated and possibly heated if condensation may occur. Both the structural and fabric components must be considered, as either may be damaged. Where structural corrosion is likely, stainless steel substitution for mild steel may be required, provided that chlorides are not present when using 300 series stainless. (Most austenitic stainless steels are susceptible to chloride corrosion.)

1.3.4.1 Temperature

The temperature of the pollutant stream must remain above the dew point of any condensables in the stream. If the temperature can be lowered without approaching the dew point, spray coolers or dilution air can be used to drop the temperature so that the temperature limits of the fabric will not be exceeded. However, the additional cost of a precooler will have to be weighed against the higher cost of bags with greater temperature resistance. The use of dilution air to cool the stream also involves a tradeoff between a less expensive fabric and a larger filter to accommodate the additional volume of the dilution air. Generally, precooling is not necessary if temperature and chemical resistant fabrics are available. (Costs for spray chambers, quenchers, and other precoolers are found in the "Wet Scrubbers" section of the Manual) Table 1.6 lists several of the fabrics in current use and provides information on temperature limits and chemical resistance. The column labeled "Flex Abrasion" indicates the fabric's suitability for cleaning by mechanical shakers.

1.3.4.2 Pressure

Standard fabric filters can be used in pressure or vacuum service but only within the range of about ± 25 inches of water. Because of the sheet metal construction of the house, they are not generally suited for more severe service. However, for special applications, high-pressure shells can be built.

Table 1.6: Properties of Leading Fabric Materials[a]

Fabric	Temp °F[b]	Acid Resistance	Alkali Resistance	Flex Abrasion
Cotton	180	Poor	Very good	Very good
Creslan[c]	250	Good in mineral	Good in weak acids	Good to very good alkali
Dacron[d]	275	Good in most mineral acids; dissolves partially in concentrated H_2SO_4	Good in weak alkali; fair in strong alkali	Very good
Dynel[e]	160	Little effect even in high concentration	Little effect even in high concentration	Fair to good
Fiberglas[f]	500	Fair to good	Fair to good	Fair
Filtron[e]	270	Good to excellent	Good	Good to very good
PTFE membrane	Depends on backing	Depends on backing	Depends on backing	Fair
Nextel[g]	1,400	Very good	Good	Good
Nomex[d]	375	Fair	Excellent at low temperature	Excellent
Nylon[d]	200	Fair	Excellent	Excellent
Orlon[d]	260	Good to excellent in mineral acids	Fair to good in weak alkali	Good
P84[h]	475	Good	Good	Good
Polypropylene	200	Excellent	Excellent	Excellent
Ryton[i]	375	Excellent	Excellent	Good
Teflon[d]	450	Inert except to fluorine	Inert except to trifluoride, chlorine, and molten alkaline metals	Fair
Wool	200	Very good	Poor	Fair to good

[a]Reference [24]
[b]Maximum continuous operating temperatures recommended by the Institute of Clean Air Companies.
[c]American Cyanamid registered trademark.
[d]Du Pont registered trademark.
[e]W. W. Criswell Div. of Wheelabrator-Fry, Inc. trade name.
[f]Owens-Corning Fiberglas registered trademark.
[g]3M Company registered trademark
[h]Inspec Fibres registered trademark
[i]Phillips Petroleum Company registered trademark

1.3.5 Equipment Design Considerations

1.3.5.1 Pressure or Suction Housings

The location of the baghouse with respect to the fan in the gas stream affects the capital cost. A suction-type baghouse, with the fan located on the downstream side of the unit, must withstand high negative pressures and therefore must be more heavily constructed and reinforced than a baghouse located downstream of the fan (pressure baghouse). The negative pressure in the suction baghouse can result in outside air infiltration, which can result in condensation, corrosion, or even explosions if combustible gases are being handled. In the case of toxic gases, this inward leakage can have an advantage over the pressure-type baghouse, where leakage is outward. The main advantage of the suction baghouse is that the fan handling the process stream is located at the clean-gas side of the baghouse. This reduces the wear and abrasion on the fan and permits the use of more efficient fans (backward-curved blade design). However, because for some designs the exhaust gases from each compartment are combined in the outlet manifold to the fan, locating compartments with leaking bags may be difficult and adds to maintenance costs. Pressure-type baghouses are generally less expensive because the housing must only withstand the differential pressure across the fabric. In some designs the baghouse has no external housing. Maintenance also is reduced because the compartments can be entered and leaking bags can be observed while the compartment is in service. With a pressure baghouse, the housing acts as the stack to contain the fumes with subsequent discharge through long ridge vents (monitors) at the roof of the structure. This configuration makes leaking bags easier to locate when the plume exits the monitor above the bag. The main disadvantage of the pressure-type baghouse in that the fan is exposed to the dirty gases where abrasion and wear on the fan blades may become a problem.

1.3.5.2 Standard or Custom Construction

The design and construction of baghouses are separated into two groups, standard and custom.[19] Standard baghouses are further separated into low, medium, and high capacity size categories. Standard baghouses are predesigned and factory built as complete off-the-shelf units that are shop-assembled and bagged for low-capacity units (hundreds to thousands of acfm throughput). Medium-capacity units (thousands to less than 100,000 acfm) have standard designs, are shop-assembled, may or may not be bagged, and have separate bag compartment and hopper sections. One form of high-capacity baghouses is the shippable module (50,000 to 100,000 acfm), which requires only moderate field assembly. These modules may have bags installed and can be shipped by truck or rail. Upon arrival, they can be operated singly or combined to form units for larger-capacity applications. Because they are preassembled, they require less field labor.

Custom baghouses, also considered high capacity, but generally 100,000 acfm or larger, are designed for specific applications and are usually built to specifications prescribed by the customer. Generally, these units are much larger than standard baghouses. For example, many are used on power plants. The cost of the custom baghouse is much higher per square foot of fabric because it is not an off-the-shelf item and requires special setups for manufacture and expensive field labor for assembly upon arrival. The advantages of the custom baghouse are many and are usually directed towards ease of maintenance, accessibility, and other customer preferences. In some standard baghouses, a complete set of bags must be replaced in a compartment at one time because of the difficulty in locating and replacing single leaking bags, whereas in custom baghouses, single bags are accessible and can be replaced one at a time as leaks develop.

1.3.5.3 Filter Media

The type of filter material used in baghouses depends on the specific application and the associated chemical composition of the gas, operating temperature, dust loading, and the physical and chemical characteristics of the particulate. Selection of a specific material, weave, finish, or weight is based primarily on past experience. For woven fabrics, the type of yarn (filament, spun, or staple), the yarn diameter, and twist are also factors in the selection of suitable fabrics for a specific application. Some applications are difficult, i.e., they have small or smooth particles that readily penetrate the cake and fabric, or have particles that adhere strongly to the fabric and are difficult to remove, or have some other characteristic that degrades particle collection or cleaning. For some of these applications Gore-Tex, a polytetrafluoroethylene (PTFE) membrane laminated to a fabric backing (felt or woven) may be used. Backing materials are chosen to be compatible with the application for which they are used. Other PTFE membrane laminated fabrics are supplied by Tetratec (Tetratex) and BHA (BHA-Tex). These membranes, because of their small pores (1 or 2 μm to less than 1 μm) are advantageous in being able to collect small particles almost immediately after filtration begins. In contrast, woven fabrics and nonwovens, (with pores about 10 μm to 100 μm) allow particles to penetrate the filter for a short time before the cake covering the fabric is reconstituted. Overall mass collection efficiency for a baghouse with membrane bags may not appear to be greater than a baghouse with other fabrics, but the efficiency may be greater for fine particles. For applications able to use paper media, cartridge filters can be particularly effective for particles in the submicron range.

Because of the violent agitation of mechanical shakers, spun or heavy weight staple yarn fabrics are commonly used with this type of cleaning, while lighter weight filament yarn fabrics are used with the gentler reverse air cleaning. Needlepunched felts are typically used for pulse-jet baghouses. These heavier fabrics are more durable than wovens when subjected to cleaning pulses. Woven fiberglass bags are an exception for high-temperature application, where they compete successfully, on a cost basis, against felted glass and other high temperature felts.

The type of material limits the maximum operating gas temperature for the baghouse. Cotton fabric has the least resistance to high temperatures (about 180°F), while of the commonly used fabrics, Fiberglas has the most (about 500°F).[2] If condensibles are contained in the gas stream, its temperature must be well above the dew point because liquid particles will usually plug the fabric pores within minutes or hours. However, the temperature must be below the maximum limit of the fabric in the bags. These maximum limits are given in Table 1.6.

1.4 Estimating Total Capital Investment

Total capital investment includes costs for the baghouse structure, the initial complement of bags, auxiliary equipment, and the usual direct and indirect costs associated with installing or erecting new structures. These costs are described below. (Costs for improving baghouse performance with electrical enhancement are not discussed in this section, but are mentioned in the example problem.)

1.4.1 Equipment Cost

1.4.1.1 Bare Baghouse Costs

Correlations of cost with fabric area for seven types of baghouses are presented. These seven types, six of which are preassembled and one, field-assembled, are listed in Table 1.7.

Table 1.7: List of cost curves for seven baghouse types

	Baghouse Type	Figure No.
	Preassembled Units	
Intermittent	Shaker (intermittent)	1.6
Continuous	Shaker (modular)	1.7
Continuous	Pulse-jet (common housing)	1.8
Continuous	Pulse-jet (modular)	1.9
Continuous	Pulse-jet (cartridge)	1.10
Continuous	Reverse-air	1.11
	Field-assembled Units	
Continuous	Any method	1.12

Each figure displays costs for a baghouse type and for additional cost items.[3] All curves are based on vendor quotes. A regression line has been fitted to the quotes and its equation is

[2]Technically, Nextel can withstand even higher temperatures—up to 1400°F. However, at approximately $15 to $20/ft^2, its price reserves its use for the relatively small number of cases in which filtration is required at temperatures above 550°F. A less expensive version of the fabric, with temperature capability to about 900EF, may be available.

given. In most cases these lines should not be extrapolated beyond the limits shown. If the reader obtains vendor quotes, they may differ from these curves by as much as ± 25%. All estimates include inlet and exhaust manifold supports, platforms, handrails, and hopper discharge devices. The indicated prices are flange to flange. The reader should note that the scale of each figure changes to accommodate the different gas flow ranges over which the various types of baghouses operate.

The 304 stainless steel add-on cost is used when such construction is necessary to prevent the exhaust gas stream from corroding the interior of the baghouse. Stainless steel is substituted for all metal surfaces that are in contact with the exhaust gas stream.

Insulation costs represent 3 inches of shop-installed glass fiber encased in a metal skin, except for custom baghouses, which have field-installed insulation. Costs for insulation include only the flange-to-flange baghouse structure on the outside of all areas in contact with the exhaust gas stream. Insulation for ductwork, fan casings, and stacks must be calculated separately as discussed later.

Figure 1.6 represents an intermittent service baghouse cleaned by a mechanical shaker.[24] This baghouse is suitable for operations that require infrequent cleaning. It can be shut down and cleaned at convenient times, such as the end of the shift or end of the day. Figure 1.6 presents the baghouse cost as a function of required fabric area. Because intermittent service baghouses do not require an extra compartment for cleaning, gross and net fabric areas are the same. The plot is linear because baghouses are made up of modular compartments and thus have little economy of scale.

Figure 1.7 presents costs for a continuously operated modular baghouse cleaned by mechanical shaker.[24] Again, price is plotted against the gross cloth area in square feet. Costs for these units, on a square foot basis, are higher than for intermittent shaker baghouses because of increased complexity and generally heavier construction.

Figures 1.8 and 1.9 show [24] common-housing and modular pulse-jet baghouses, respectively. Common housing units have all bags within one housing; modular units are constructed of separate modules that may be arranged for off-line cleaning. Note that in the single-unit (common-housing) pulse jet, for the range shown, the height and width of the unit are constant and the length increases; thus, for a different reason than that for the modular units discussed above, the cost increases linearly with size. Because the common housing is relatively inexpensive, the stainless steel add-on is proportionately higher than for modular units. Added material costs and setup and labor charges associated with the less workable stainless steel

[3]Costs in Figures 1.6 to 1.12 are in second quarter 1998 dollars. For information on escalating these prices to more current dollars, refer to the EPA report Escalation Indexes for Air Pollution Control Costs and updates thereto, all of which are installed on the OAQPS Technology Transfer Network at http://www.epa.gov/ttn/catc.

account for most of the added expense. Figure 1.10 shows costs for cartridge baghouses cleaned by pulse.

Figures 1.11 and 1.12 show costs for modular and custom-built reverse-air baghouses, respectively.[24] The latter units, because of their large size, must be field assembled. They are often used on power plants, steel mills, or other applications too large for the factory-

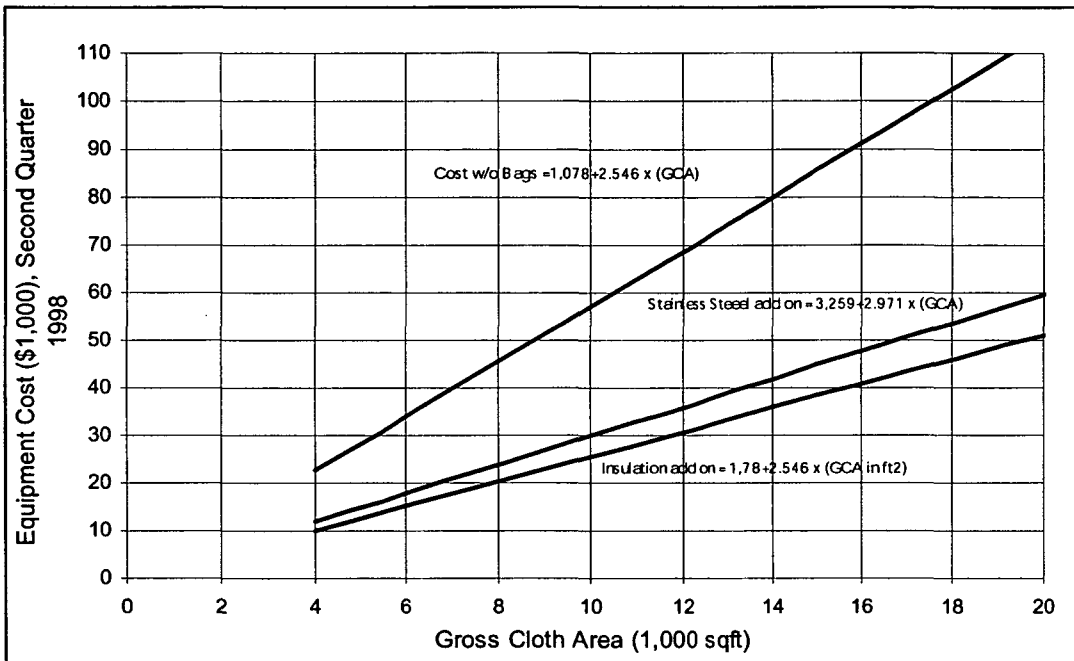


Figure 1.6: Equipment Costs for Shaker Filters (Intermittent)
Note: This graph should not be extrapolated.
Note: GCA = Gross Cloth Area in sqft
Source: ETS Inc.

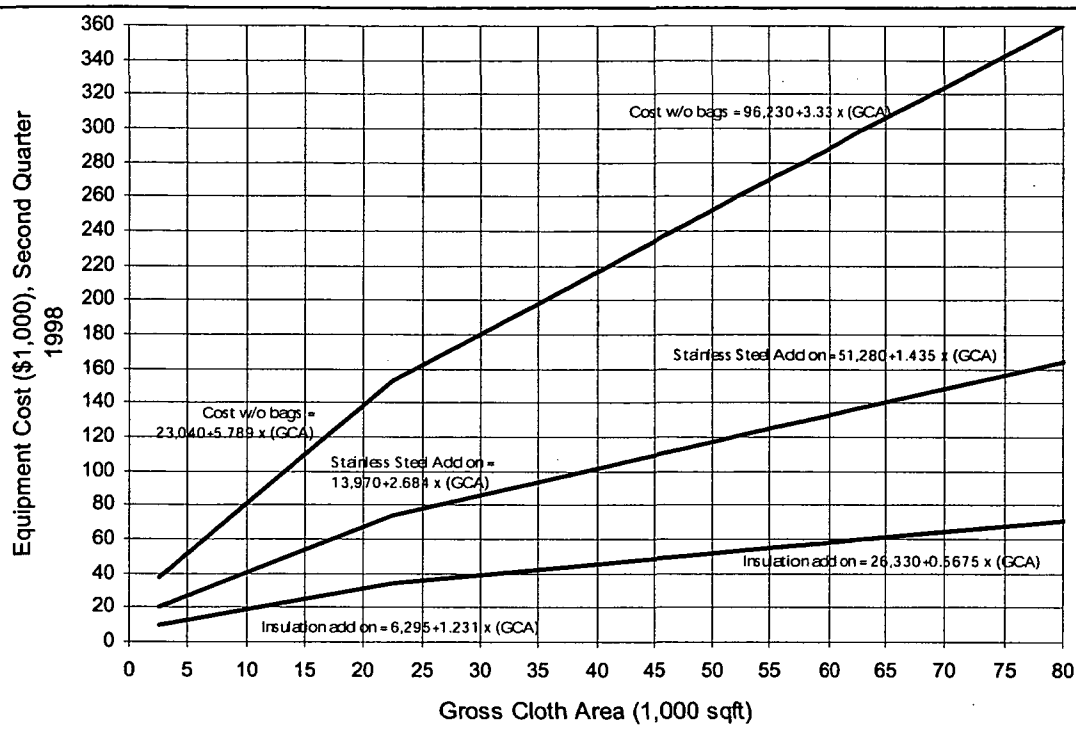


Figure 1.7: Equipment Costs for Shaker Filters (Continuous)

Note: this graph should not be extrapolated

Note: GCA = Gross Cloth Area in sqft

Source: ETS Inc.

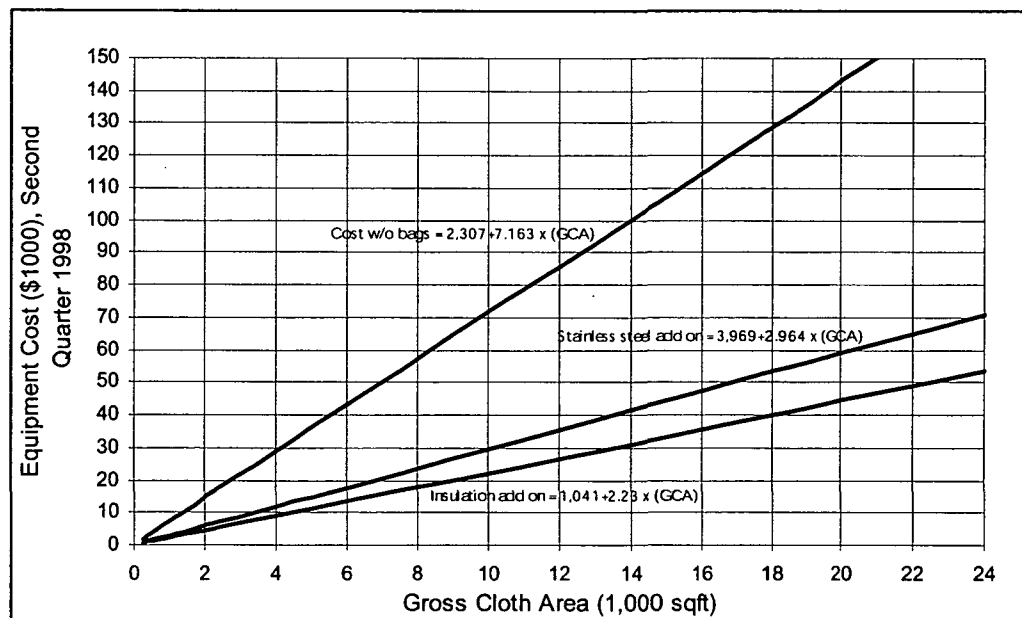


Figure 1.8: Equipment Costs for Pulse-Jet Filters (Common Housing)

Note: this graph should not be extrapolated

Note: GCA= Gross Cloth Area in sqft

Source: ETS Inc.

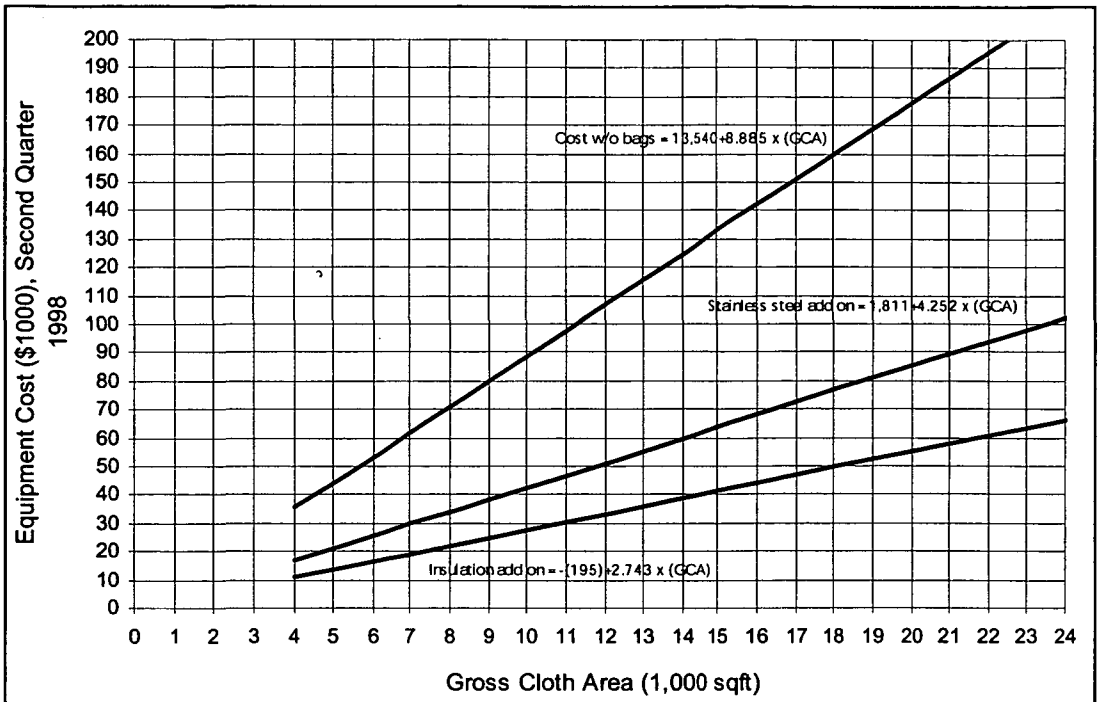


Figure 1.9: Equipment Costs for Pulse-Jet Filters (modular)
Note: this chart should not be extrapolated
Note: GCA= Gross Cloth Area in sqft
Source: ETS Inc.

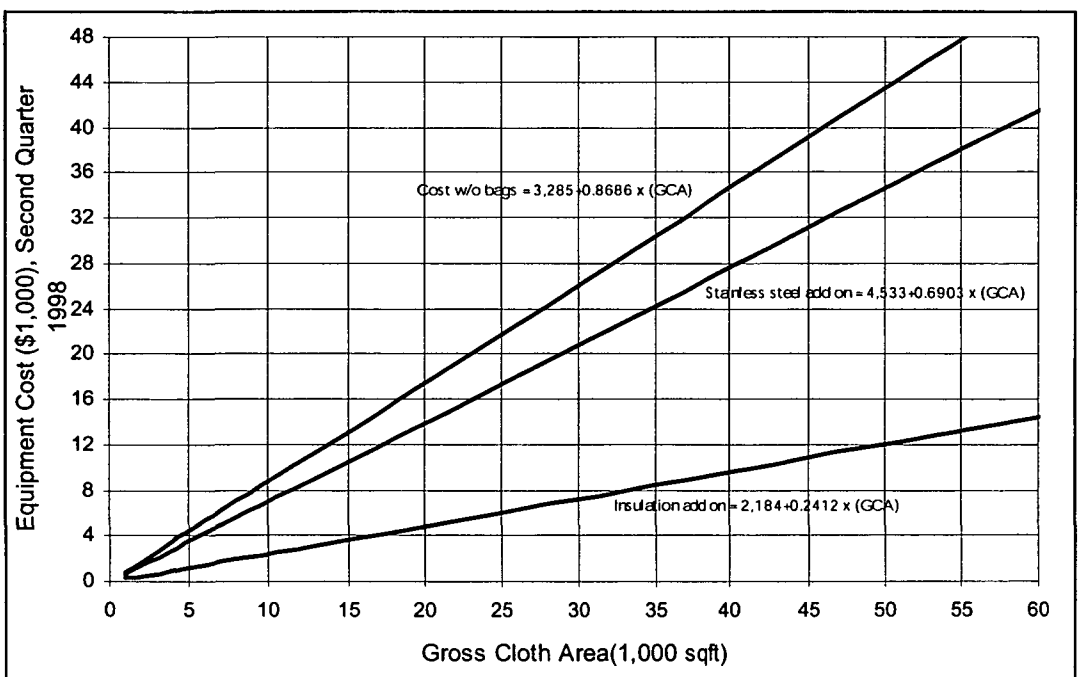


Figure 1.10: Equipment Costs for Cartirdge Filters
Note: this graph should not be extrapolated
Note: GCA= Gross Cloth Area in sqft
Source: ETS Inc.

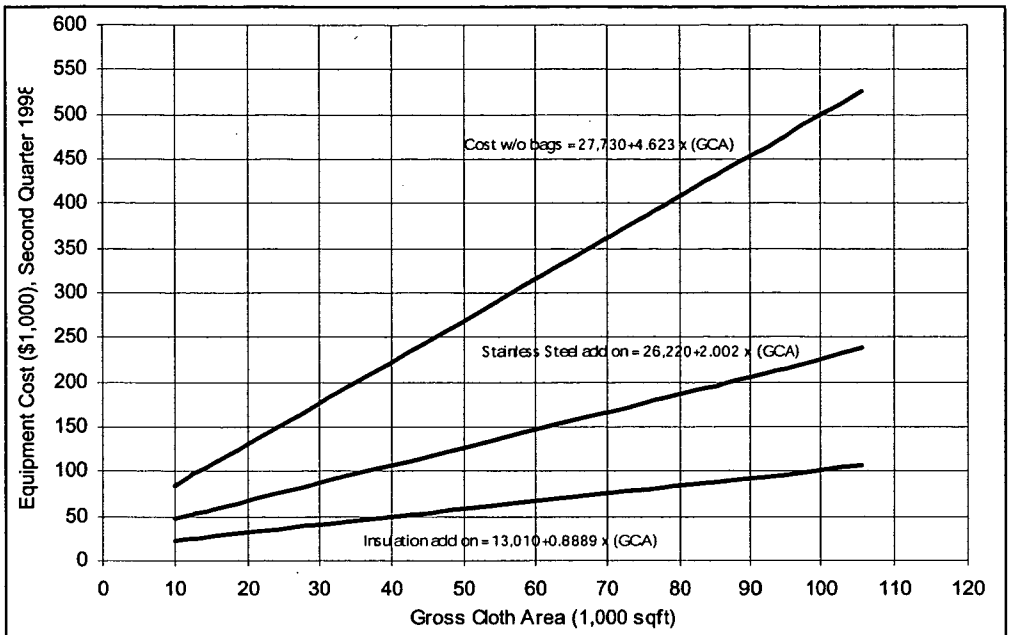


Figure 1.11: Equipment Costs for Reverse-Air Filters (Modular)
Note: this graph should not be extrapolated
Note GCA= Gross Cloth Area in sqft
Source: ETS Inc.

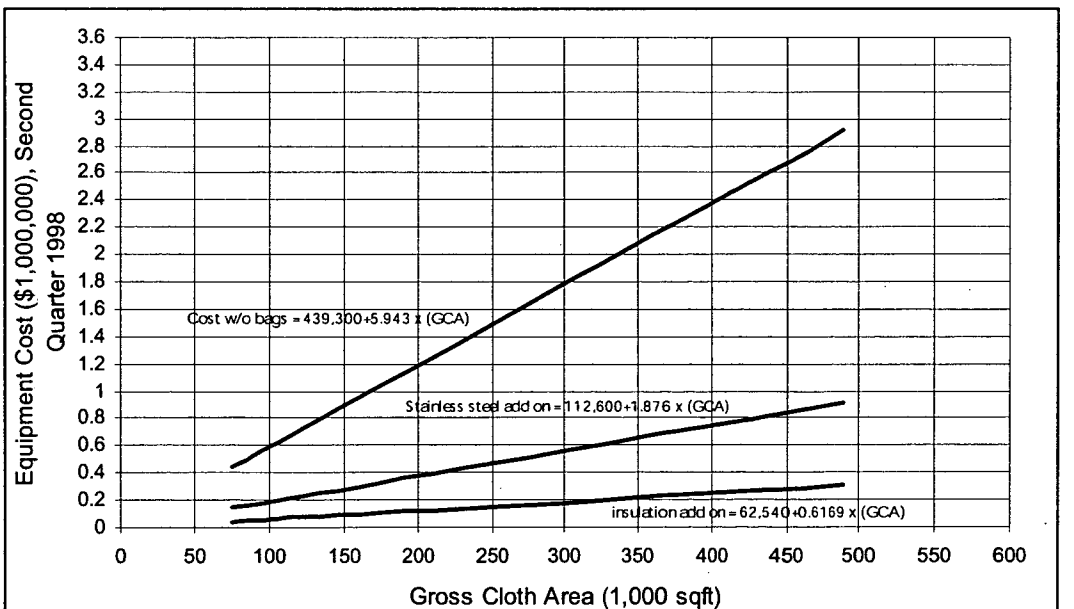


Figure 1.12: Equipment Costs for Reverse -Air filters (Custom Built)
Note: this graph should not be extrapolated
Note GCA= Gross Cloth Area in sqft
Source: ETS Inc.

assembled baghouses. Prices for custom-built shaker units are not shown, but are expected to be similar to custom-built reverse-air units.

1.4.1.2 Bag Costs

Table 1.8 gives the 1998 price per square foot of bags by type of fabric and by type of cleaning system used. Actual quoted prices may vary by ± 10 % from the values in the table. When estimating bag costs for an entire baghouse, gross cloth area as determined from Table 1.2 should be used. Membrane PTFE fabric costs are a combination of the base fabric cost and a premium for the PTFE laminate and its application. As fiber market conditions change, the costs of fabrics relative to each other also change. Prices are based on typical fabric weights in ounces/ square yard. Sewn-in snap rings are included in the price, but other mounting hardware, such as clamps or cages, must be added, based on the type of baghouse.

1.4.1.3 Auxiliary Equipment

Figure 1.1 shows auxiliary equipment, which is discussed elsewhere in the Manual. Because hoods, ductwork, precoolers, cyclones, fans, motors, dust removal equipment and stacks are common to many pollution control systems, they are (or will be) given extended treatment in separate chapters. For instance, Section 2 provides sizing and costing procedures and data for hoods, ductwork, and stacks.

1.4.2 Total Purchased Cost

The total purchased cost of the fabric filter system is the sum of the costs of the baghouse, bags, and auxiliary equipment; instruments and controls, taxes, and freight. Instruments and controls, taxes, and freight are generally taken as percentages of the estimated total cost of the first three items. Typical values, from Section 1, are 10% for instruments and controls, 3% for taxes, and 5% for freight.

Bag costs vary from less than 15% to more than 100% of the cost of the bare baghouse (baghouse without bags or auxiliaries), depending on the type of fabric required. This situation makes it inadvisable to estimate total purchased cost without separately estimating baghouse and bag costs, and discourages the use of a single factor to estimate a cost for the combined baghouse and bags.

Table 1.8: Bag Prices
(2^{nd} quarter 1998 $/ft^2$)

Type of Cleaning	Bag Diameter (inches)	Type of Material[a] PE	PP	NO	HA	FG	CO	TF	P8	RT	NX
Pulse jet, TR[b]	4-1/2 to 5-1/8	0.75	0.81	2.17	1.24	1.92	NA	12.21	4.06	2.87	20.66
	6 to 8	0.67	0.72	1.95	1.15	1.60	NA	9.70	3.85	2.62	NA
Pulse jet, BBR	4-1/2 to 5-1/8	0.53	0.53	1.84	0.95	1.69	NA	12.92	3.60	2.42	16.67
	6 to 8	0.50	0.60	1.77	0.98	1.55	NA	9.00	3.51	2.30	NA
Pulse jet, Cartridge[c]	4-7/8	2.95	NA	6.12	NA	NA	NA	NA	NA	NA	NA
	6-1/ 8	1.53	NA	4.67	NA	NA	NA	NA	NA	NA	NA
Shaker, Strap top	5	0.63	0.88	1.61	1.03	NA	0.70	NA	NA	NA	NA
Shaker, Loop top	5	0.61	1.01	1.53	1.04	NA	0.59	NA	NA	NA	NA
Reverse air with rings	8	0.63	1.52	1.35	NA	1.14	NA	NA	NA	NA	NA
	11-1/2	0.62	NA	1.43	NA	1.01	NA	NA	NA	NA	NA
Reverse air w/o rings	8	0.44	NA	1.39	NA	0.95	NA	NA	NA	NA	NA
	11-1/2	0.44	NA	1.17	NA	0.75	NA	NA	NA	NA	NA

NA = Not applicable.

[a]Materials:

PE = 16-oz polyester
PP = 16-oz polypropylene
NO = 14-oz Nomex
HA = 16-oz homopolymer acrylic
FG = 16-oz fiberglass with 10% Teflon
CO = 9-oz cotton
TF = 22-oz Teflon felt
P8 = 16-oz P84
RT = 16-oz Ryton
NX = 16-oz Nextel

[b]Bag removal methods:

TR = Top bag removal (snap in)
BBR = Bottom bag removal

[c]Costs for 12.75-in. diameter by 26-in. length cartridges are $59.72 for a polyester/cellulose blend ($0.26/ft^2 for 226 ft^2) and $126.00 for spunbonded polyester ($1.26/ft^2 for 100 ft^2).

NOTE: For pulse-jet baghouses, all bags are felts except for the fiberglass, which is woven. For bottom access pulse jets, the mild steel cage price for one 4 1/2-in. diameter cage or one 5 5/8-in. diameter cage can be calculated from the single-bag fabric area using the following two sets of equations, respectively.

Table 1.8: (Cont.)

4-1/2 in. x 8 ft cages:	5-5/8 in x 10 ft cages:
$\$ = 7.8444\ exp(0.0355\ ft^2)$ in 25 cage lots	$\$ = 5.6542\ ft^{2\ (0.4018)}$ in 25 cage lots
$\$ = 6.0211\ exp(0.0423\ ft2)$ in 50 cage lots	$\$ = 4.3080\ ft^{2\ (0.4552)}$ in 50 cage lots
$\$ = 4.2635\ exp(0.0522\ ft^2)$ in 100 cage lots	$\$ = 3.0807\ ft^{2\ (0.5249)}$ in 100 cage lots
$\$ = 3.4217\ exp(0.0593\ ft^2)$ in 500 cage lots	$\$ = 2.5212\ ft^{2\ (0.5686)}$ in 500 cage lots

These costs apply to 8-foot and 10-foot cages made of 11 gauge mild steel and having 10 vertical wires and "Roll Band" tops. For snap-band collar with built-in venturi, add $6.00 per cage for mild steel and $13.00 per cage for stainless steel. For stainless steel cages use:

$\$ = 8.8486 + 1.5734\ ft^2$ in 25 cage lots	$\$ = 21.851 + 1.2284\ ft^2$ in 25 cage lots
$\$ = 6.8486 + 1.5734\ ft^2$ in 50 cage lots	$\$ = 8.8486 + 1.2284\ ft^2$ in 50 cage lots
$\$ = 4.8466 + 1.5734\ ft^2$ in 100 cage lots	$\$ = 8.8486 + 1.2284\ ft^2$ in 100 cage lots
$\$ = 3.8486 + 1.5734\ ft^2$ in 500 cage lots	$\$ = 8.8486 + 1.2284\ ft^2$ in 500 cage lots

For shakers and reverse air baghouses, all bags are woven. All prices are for finished bags, and prices can vary from one supplier to another. For membrane bag prices, multiply base fabric price by factors of 3 to 4.5.

Sources: ETS Inc.[24]

1.4.3 Total Capital Investment

The total capital investment (TCI) is the sum of three costs, purchased equipment cost, direct installation costs, and indirect installation costs. The factors needed to estimate the TCI are given in Table 1.9. The Table 1.9 factors may be too large for "packaged" fabric filters—those pre-assembled baghouses that consist of the compartments, bags, waste gas fan and motor, and instruments and controls. Because these packaged units require very little installation, their installation costs would be lower (20–25% of the purchased equipment cost). Because bag costs affect total purchased equipment cost, the cost factors in Table 1.9 may cause overestimation of total capital investment when expensive bags are used. Using stainless steel components can also cause overestimation. Because baghouses range in size, specific factors for site preparation or for buildings are not given. Costs for buildings may be obtained from such references as Means Construction Cost Data 1998.[25] Land, working capital, and off-site facilities are not normally required and have been excluded from the table. When necessary, these costs can be estimated.

1.5 Estimating Total Annual Costs

1.5.1 Direct Annual Cost

Direct annual costs include operating and supervisory labor, operating materials, replacement bags, maintenance (labor and materials), utilities, and dust disposal. Most of these costs are discussed individually below. They vary with location and time, and, for this reason, should be obtained to suit the specific baghouse system being costed. For example, current labor rates may be found in such publications as the Monthly Labor Review, published by the U.S. Department of Labor, Bureau of Labor Statistics (BLS), or obtained from the BLS web site at: *http://stats.bls.gov.*

1.5.1.1 Operating and Supervisory Labor

Typical operating labor requirements are 2 to 4 hours per shift for a wide range of filter sizes.[26] When fabric filters are operated to meet Maximum Achievable Control Technology (MACT) regulations, it is likely that the upper end of the range is appropriate. Small or well-performing units may require less time, while very large or troublesome units may require more. Supervisory labor is taken as 15% of operating labor.

1.5.1.2 Operating Materials

Operating materials are generally not required for baghouses. An exception is the use of precoat materials injected on the inlet side of the baghouse to provide a protective dust layer on the bags when sticky or corrosive particles might harm them. Adsorbents may be similarly injected when the baghouse is used for simultaneous particle and gas removal. Costs for these materials should be included on a dollars-per-mass basis (e.g., dollars per ton).

1.5.1.3 Maintenance

Maintenance labor varies from 1 to 2 hours per shift.[26] As with operating labor, these values may be reduced or exceeded depending on the size and operating difficulty of a particular unit. The upper end of the range may be required for operation to meet MACT regulations. Maintenance materials costs are assumed to be equal to maintenance labor costs.[26]

Table 1.9 Capital Cost Factors for Fabric Filters[a]

Cost Item	Factor
Direct costs	
Purchased equipment costs	
Fabric filter (EC) + bags + auxiliary equipment	As estimated, A
Instrumentation	0.10 A
Sales taxes	0.03 A
Freight	0.05 A
Purchased Equipment Cost, PEC	B = 1.18 A
Direct installation costs	
Foundations & supports	0.04 B
Handling & erection	0.50 B
Electrical	0.08 B
Piping	0.01 B
Insulation for ductwork[b]	0.07 B
Painting[c]	0.04 B
Direct installation cost	0.74 B
Site preparation	As required, SP
Buildings	As required, Bldg.
Total Direct Cost	**1.74 B + SP + Bldg.**
Indirect Costs (installation)	
Engineering	0.10 B
Construction and field expense	0.20 B
Contractor fees	0.10 B
Start-up	0.01 B
Performance test	0.01 B
Contingencies	0.03 B
Total Indirect Cost, IC	**0.45 B**
Total Capital Investment = DC + IC	**2.19 B + SP + Bldg.**

[a]Reference [29], revised

[b]Ductwork and stack costs, including insulation costs, may be obtained from Chapter 10 of the manual. This installation factor pertains solely to insulation for fan housings and other auxiliaries, except for ductwork and stacks.

[c]The increased use of special coatings may increase this factor to 0.06B or higher. [The factors given in Table 1.8 are for average installation conditions. Considerable variation may be seen with other-than-average installation circumstances.]

1.5.1.4 Replacement Parts

Replacement parts consist of filter bags, which have a typical operating life of about 2 to 4 years. The following formula is used for computing the bag replacement cost:

$$CRC_B = (C_B + C_L) \times CRF_B \quad (1.13)$$

where

CRC_B = bag capital recovery cost ($/year)
C_B = initial bag cost including taxes and freight ($)
C_L = bag replacement labor ($)
CRF_B = capital recovery factor (defined in Chapter 2) whose value is a function of the annual interest rate and the useful life of the bags (For instance, for a 7% interest rate and a 2-year life, $CRF_B = 0.5531$.)

Bag replacement labor cost (C_L) depends on the number, size, and type of bags; their accessibility; how they are connected to the baghouse tube-sheet; and other site-specific factors that increase or decrease the quantity of labor required. For example, a reverse-air baghouse probably requires from 10 to 20 person-minutes to change an 8-inch by 24-foot bag that is clamped in place. Based on a filtering surface area of approximately 50 ft^2 and a labor rate of $29.15/h (including overhead), C_L would be $0.10 to $0.19/ft^2 of bag area. As Table 1.8 shows, for some bags (e.g., polyester), this range of C_L would constitute a significant fraction of the purchased cost. For pulse jets, replacement time would be about 5 to 10 person-minutes for a 5-inch by 10-foot bag in a top-access baghouse, or $0.19 to $0.37/ft^2 of bag area. This greater cost is partially offset by having less cloth in the baghouse, but there may be more of the smaller bags. These bag replacement times are based on changing a minimum of an entire module and on having typical baghouse designs. Times would be significantly longer if only a few bags were being replaced or if the design for bag attachment or access were atypical. Cartridge baghouses with horizontal mounting take about 4 minutes to change one cartridge. Older style baghouses with vertical mounting and blow pipes across the cartridges take about 20 min/cartridge.

The Manual methodology treats bags and bag replacement labor as an investment amortized over the useful life of the bags, while the rest of the control system is amortized over its useful life, typically 20 years (see Subsection 1.5.2). Capital recovery factor values for bags with different useful lives can be calculated based on the method presented in Section 1.

1.5.1.5 Electricity

Electricity is required to operate system fans and cleaning equipment. Primary gas fan power can be calculated as described in Chapter 2 of Section 2 and assuming a combined fan-motor efficiency of 0.65 and a specific gravity of 1.000. We obtain:[27]

$$Power_{fan} = 0.000181\ Q(\Delta P)\theta \qquad (1.14)$$

where

$Power_{fan}$	=	fan power requirement (kWh/yr)
Q	=	system flow rate (acfm)
ΔP	=	system pressure drop (in. H_2O)
θ	=	operating time (h/yr)

Cleaning energy for reverse-air systems can be calculated (using equation 1.14) from the number of compartments to be cleaned at one time (usually one, sometimes two), and the reverse gas-to-cloth ratio (from about one to two times the forward gas-to-cloth ratio). Reverse-air pressure drop varies up to 6 or 7 in. H_2O depending on location of the fan pickup (before or after the main system fan).[28] The reverse-air fan generally runs continuously.

Typical energy consumption in kWh/yr for a shaker system operated 8,760 h/yr can be calculated from:[5]

$$P = 0.053\ A \qquad (1.15)$$

where

A	=	gross fabric area (ft^2)

1.5.1.6 Fuel

Fuel costs must be calculated if the baghouse or associated ductwork is heated to prevent condensation. These costs can be significant, but may be difficult to predict. For methods of calculating heat transfer requirements, see Perry.[29]

1.5.1.7 Water

Cooling process gases to acceptable temperatures for fabrics being used can be done by dilution with air, evaporation with water, or heat exchange with normal equipment. Evaporation and normal heat exchange equipment require consumption of plant water, although costs are not usually significant. Chapter 1 of Section 3.1, Adsorbers, provides information on estimating cooling-water costs.

1.5.1.8 Compressed Air

Pulse-jet filters use compressed air at pressures from about 60 to 100 psig. Typical consumption is about 2 scfm/1,000 cfm of gas filtered.[5] For example, a unit filtering 20,000 cfm of gas uses about 40 scfm of compressed air for each minute the filter is operated. For each pulse, cartridge filters with nonwoven fabrics use 10 scfm/1,000 ft^2 or 14 scfm/1,000 ft^2 at 60 psig or 90 psig pulse pressure, respectively, in one manufacturer's design.[30] When using paper media, the air quantities are 1.7 scfm/1,000 ft^2 and 2.2 scfm/1,000 ft^2 at the respective pressures. Pulse frequency ranges from about 5 min. to 15 min. A typical cost for compressed air is $0.25/1,000 scf in 1998 dollars.

1.5.1.9 Dust Disposal

If collected dust cannot be recycled or sold, it must be landfilled or disposed of in some other manner. Disposal costs are site-specific, but typically run $35 to $55 per ton at municipal waste sites in Pennsylvania, exclusive of transportation (see Section 1). Lower costs may be available for industrial operations with long-term disposal contracts. Hazardous waste disposal can cost $150 per ton or more.

1.5.2 Indirect Annual Cost

Indirect annual costs include capital recovery, property taxes, insurance, administrative costs ("G&A"), and overhead. The capital recovery cost is based on the equipment lifetime and the annual interest rate employed. (See Section 1 for a discussion of the capital recovery cost and the variables that determine it.) For fabric filters, the system lifetime varies from 5 to 40 years, with 20 years being typical.[26] However, this does not apply to the bags, which usually have much shorter lives. Therefore, one should base system capital recovery cost estimates on the installed capital cost, less the cost of replacing the bags (*i.e.*, the purchased cost of the bags plus the cost of labor necessary to replace them). Algebraically:

$$CRC_s = [TCI - C_B - C_L]CRF_s \qquad (1.16)$$

where

CRC_s	=	capital recovery cost for fabric filter system ($/yr)
TCI	=	total capital investment ($)
C_B	=	initial cost of bags *including* taxes and freight ($)[4]
C_L	=	labor cost for replacing bags ($)
CRF_s	=	capital recovery factor for fabric filter system (defined in Chapter 2).

For example, for a 20-year system life and a 7% annual interest rate, the CRF_s would be 0.09439.

The suggested factor to use for property taxes, insurance, and administrative charges is 4% of the *TCI* (see Section 1). Finally, overhead is calculated as 60% of the total labor (operating, supervisory, and maintenance) and maintenance materials.

1.5.3 Recovery Credits

For processes that can reuse the dust collected in the baghouse or that can sell the dust (e.g., fly ash sold as an extender for paving mixes), a recovery credit (*RC*) should be taken. As used in equation 1.17, this credit (*RC*) is subtracted from the *TAC*.

1.5.4 Total Annual Cost

Total annual cost for owning and operating a fabric filter system is the sum of the components listed in Sections 1.5.1 through 1.5.3:

$$TAC = DC + IC - RC \tag{1.17}$$

where

TAC	=	total annual cost ($)
DC	=	direct annual cost ($)
IC	=	indirect annual cost ($)
RC	=	recovery credits (annual) ($)

1.6 Example Problem

Assume a baghouse is required for controlling fly ash emissions from a coal-fired boiler. The flue gas stream is 50,000 acfm at 325 °F and has an ash loading of 4 gr/acf. Analysis of the ash shows a mass median diameter of 7 μm. Assume the baghouse operates for 8,640 h/yr (360 d).

[4]Typically, 8% of the bag initial cost.

The gas-to-cloth ratio (G/C) can be taken from Table 1.1 as 2.5, for woven fabrics in shaker or reverse-air baghouses, or 5, for felts used in pulse-jet baghouses. If a factor method were used for estimating G/C, Table 1.3 for shakers would yield the following values: $A = 2$, $B = 0.9$, and $C = 1.0$. The gas-to-cloth ratio would be:

$$2 \times 0.9 \times 1.0 = 1.8.$$

This value could also be used for reverse-air cleaning. For a pulse-jet unit, Table 1.4 gives a value of 9.0 for factor A and 0.8 for factor B. Equation 1.11 becomes:

$$V = 2.878 \times 9.0 \times 0.8(275)^{-0.2335}(4)^{-0.06021}(0.7471 + 0.0853 \ln 7)$$
$$= 4.69$$

Because this value is so much greater than the shaker/reverse-air G/C, we conclude that the pulse-jet baghouse would be the least costly design. This conclusion is based on the inference that a much bigger G/C would yield lower capital and, in turn, annual costs. However, to make a more rigorous selection, we would need to calculate and compare the total annual costs of all three baghouse designs (assuming all three are technically acceptable). The reader is invited to make this comparison. Further discussion of the effects of G/C increases, and accompanying pressure drop increases, on overall annual costs will be found in Reference 30.[5] Assume the use of on-line cleaning in a common housing structure and, due to the high operating temperature, the use of glass filter bags (see Table 1.6).[6] At a gas-to-cloth ratio of 4.69, the fabric required is[7]

$$50{,}000 \text{ acfm}/4.69 \text{ fpm} = 10{,}661 \text{ ft}^2.$$

From Figure 1.8, the cost of the baghouse ("common housing" design) is:

$$Cost = 2{,}307 + 7.163(10{,}661) = \$78{,}672$$

[5]In addition, the CO$T-AIR control cost spreadsheet for fabric filters computes capital and annual costs for all three designs. Download CO$T-AIR at: *http://www.epa.gov/ttn/catc/products.html#ccc.info.*

[6]As Table 1.6 shows, other bag materials (e.g., Nomex) also could withstand this operating temperature. But Fiberglas is the least expensive on a purchased cost basis. For harsh environments, a more expensive, but more durable bag might cost less on a total annual cost basis.

[7]This is the total (gross) bag area required. No bag adjustment factor has been applied here, because this is a common housing pulse jet unit that is cleaned continuously during operation. Thus, no extra bag compartment is needed, and the gross and net bag areas are equal.

Insulation is required. The insulation add-on cost from Figure 1.4 is:

$$Cost = 1{,}041 + 2.23(10{,}661) = \$24{,}815$$

From Table 1.8, bag costs are \$1.69/ft² for 5-1/8-inch diameter glass fiber, bottom removal bags. Total bag cost is

$$10{,}661 \text{ ft}^2 \text{ x } \$1.69/\text{ft}^2 = \$18{,}017.$$

For 10 ft long cages,

$$fabric\ area\ per\ cage = \frac{\left(5\frac{1}{8}\,in\right)}{\left(12\,\frac{in}{ft}\right)} \times \pi \times 10\,ft = 13.42\,ft^2$$

$$the\ number\ of\ cages = \frac{\left(10{,}661\,ft^2\right)}{\left(13.42\,ft^2\right)}$$

$$= 795\ \ cages\ \ \left(rounded\ up\ to\ the\ next\ integer\right)$$

From Table 1.7, individual cage cost is

$$2.5212 \text{ x } 13.42 \text{ ft}^{2(0.5686)} = \$11.037.$$

Total cage cost is

$$795 \text{ cages x } \$11.037/\text{cage} = \$8{,}774.$$

Assume the following auxiliary costs have been estimated from data in other parts of the Manual:

Ductwork	$19,000
Fan	19,000
Motor	12,000
Starter	4,700
Dampers	9,800
Compressor	8,000
Screw conveyor	5,000
Stack	12,000
Total	**$89,500**

Direct costs for the fabric filter system, based on the factors in Table 1.9, are given in Table 1.10. (Again, we assume site preparation and buildings costs to be negligible.) Total capital investment is $569,000. Table 1.11 gives the direct and indirect annual costs, as calculated from the factors given in Section 1.5.1. For bag replacement labor, assume 10 min per bag for each of the 795 bags. At a maintenance labor rate of $29.65 (including overhead), the labor cost is $3,943 for 133 h. The bags and cages are assumed to be replaced every 2 yr. The replacement cost is calculated using Equation 1.13.

Pressure drop (for energy costs) can be calculated from Equations 1.8 and 1.9, with the following assumed values:

$$K_2 = 15 \frac{\frac{in\ H_2O}{1(ft/min)}}{\frac{lb}{ft^2}}$$

$$P_j = 100\ \ psig$$

$$cleaning\ interval = 10\ min$$

We further assume that a *G/C* of 4.69 ft/min is a good estimate of the mean face velocity over the duration of the filtering cycle.

$$W_o = C_i V \theta$$

$$= 4 \frac{gr}{ft^3} \times \frac{1lb}{7{,}000gr} \times 4.69 \frac{ft}{min} \times 10min$$

$$= 0.0268 \frac{lb}{ft^2}$$

$$\Delta P = 6.08 \times 4.69 \frac{ft}{min} \times (100\ psig)^{-0.65}$$

$$+ 15 \frac{\frac{in H_2O}{ft/min}}{lb/ft^2} \times 0.0268 \frac{lb}{ft^2} \times 4.69 \frac{ft}{min}$$

$$= 3.32\ in\ H_2O \text{ across the fabric (when fully loaded).}$$

Assume that the baghouse structure and the ductwork contribute an additional 3 in. H_2O and 4 in. H_2O, respectively. The total pressure drop is, therefore, 10.3 inches.

The total annual cost is $474,000, 39 percent of which is for ash disposal. If a market for the fly ash could be found, the total annual cost would be greatly reduced. For example, if $2/ton were received for the ash, the total annual cost would drop to $274,000 ($474,000 – $185,000 – $14,800), or 58% of the cost when no market exists. Clearly, the total annual cost is extremely sensitive to the value chosen for the dust disposal cost in this case. In this and in similar cases, this value should be selected with care.

Table 1.10 Capital Costs for Fabric Filter System
Example Problem (2nd quarter 1998 $)

Cost Item	Cost
Direct Costs	
Purchased equipment costs	
Fabric filter (with insulation)(EC)	$103,847
Bags and cages	26,791
Auxiliary equipment	89,500
Sum = A	$220,138
Instrumentation, 0.1A	22,014
Sales taxes, 0.03A	6,604
Freight, 0.05A	11,007
Purchased equipment cost, B	$259,763
Direct installation costs	
Foundation and supports, 0.04B	10,391
Handling and erection, 0.50B	129,882
Electrical, 0.08B	20,781
Piping, 0.01B	2,598
Insulation for ductwork, 0.07B	18,183
Painting, 0.04B	10,391
Direct installation cost	192,226
Site preparation	-
Facilities and buildings	-
Total Direct Cost	**$451,989**
Indirect Costs (installation)	
Engineering, 0.10B	25,976
Construction and field expenses, 0.20B	51,953
Contractor fees, 0.10B	25,976
Start-up, 0.01B	2,598
Performance test, 0.01B	2,598
Contingencies, 0.03B	7,793
Total Indirect Cost	**$116,894**
Total Capital Investment (rounded)	**$569,000**

Table 1.11 Annual Costs for Fabric Filter System
Example Problem (2nd quarter 1998 $)

Cost Item	Calculations	Cost
Direct Annual Costs, DC		
Operating labor		
Operator	$\frac{2\ h}{shift} \times \frac{3\ shifts}{day} \times \frac{360\ days}{yr} \times \frac{\$17.26}{h}$	$37,282
Supervisor	15% of operator = 0.15 x 37,282	5,592
Operating materials		—
Maintenance		
Labor	$\frac{1\ h}{shift} \times \frac{3\ shifts}{day} \times \frac{360\ days}{yr} \times \frac{\$17.74}{h}$	19,159
Material	100% of maintenance labor	19,159
Replacement parts, bags	[3,943 + (26,791 x 1.08[a])] x 0.5531	18,184
Utilities		
Electricity	$0.000181 \times 50{,}000\ acfm \times 10.3\ in\ H_2O \times \frac{8{,}640\ h}{yr} \times \frac{\$0.0671}{kWh}$	54,041
Compressed air (dried and filtered)	$\frac{2\ scfm}{1{,}000\ acfm} \times 50{,}000\ acfm \times \frac{\$0.25}{1{,}000\ scf} \times \frac{60\ min}{h} \times \frac{\$8{,}640\ h}{yr}$	12,960
Waste disposal	at $25/ton on-site for essentially 100% collection $\frac{4\ gr}{ft^3} \times \frac{1\ lb}{7{,}000\ gr} \times 50{,}000\ ft^3 \times \frac{60\ min}{h} \times \frac{8{,}640\ h}{yr} \times \frac{1\ ton}{2{,}000\ lb} \times \frac{\$25}{ton}$	185,134
Total DC	(rounded)	**351,500**
Indirect Annual Costs, IC		
Overhead	60% of sum of operating, supv., & maint. labor & maint. materials = 0.6(37,282+5,592+19,159+19,159)	48,715
Administrative charges	2% of Total Capital Investment = 0.02 ($568,883)	11,378
Property Tax	1% of Total Capital Investment = 0.01 ($568,883)	5,689
Insurance	1% of Total Capital Investment = 0.01 ($568,883)	5,689
Capital recovery[b]	0.09439 (568,883- 3,943 - 28,934 x 1.08)	50,594
Total IC (rounded)		**122,100**
Total Annual Cost	**(rounded)**	**$474,000**

[a]The 1.08 factor is for freight and sales taxes.

[b]The capital recovery cost factor, CRF, is a function of the fabric filter or equipment life and the opportunity cost of the capital (*i.e.*, interest rate). For example, for a 20-year equipment life and a 7% interest rate, CRF = 0.09439.

1.7 Acknowledgments

We gratefully acknowledge the following companies and individuals for contributing data to this section:

- Aget Manufacturing Company (Adrian, Michigan)
- Andrew S. Viner (formerly of RTI, North Carolina)
- BACT Engineering, Inc. (Arlington Heights, Illinois)
- The BHA Group (Kansas City, Missouri)
- Dustex Corporation (Charlotte, North Carolina)
- EV Air Systems, Inc. (Charlotte, North Carolina)
- Fuller Company (Bethlehem, Pennsylvania)
- W. L. Gore and Associates, Inc. (Elkton, Maryland)
- Griffin Environmental Company, Inc. (Syracuse, New York)
- Hosokawa Mikropul (Summit, New Jersey)
- Richard E. Jenkins (formerly of EPA)
- W. W. Sly Manufacturing Company (Cleveland, Ohio)
- Zurn Industries, Inc. (Birmingham, Alabama)

References

[1] Van Osdell, D. W., M. B. Ranade, G. P. Greiner, and D. F. Furlong, *Electrostatic Augmentation of Fabric Filtration: Pulse-jet Pilot Unit Experience*, November 1982 (EPA-600/7-82-062).

[2] Viner, A. S., G. P. Greiner, D. F. Furlong, and R. G. Hurst, *Pilot-Scale Evaluation of Top-Inlet and Advanced Electrostatic Filtration*, October 1986 (EPA-600/7-86-042).

[3] Donovan, R. P., *Fabric Filtration For Combustion Sources*, Marcel Dekker, Inc., New York, 1985.

[4] Turner, J. H., "Bag Filtration," in *Handbook of Multiphase Systems*, ed. by G. Hetsroni, Hemisphere, New York, 1982.

[5] Turner, J. H., and J. D. McKenna, "Control of Particles by Filters," in *Handbook of Air Pollution Technology*, ed. by S. Calvert and E. Englund, John Wiley & Sons, New York, 1984.

[6] Palazzo, L., J. Woolston, and P. Ristevski, *Retrofitting Shaker Baghouses to Cartridge Pulse Jet Technology* in *The User and Fabric Filtration Equipment - VII*, Proceedings, Toronto, September 12 - 14, 1994, Air and Waste Management Association, Pittsburgh.

[7] Klimczak, William, personal communication and information from Dustex publication P6100, *Principles of Pulse Jet Filter Performance.*

[8] Broadway, R. M. and R. W. Cass, *Fractional Efficiency of a Utility Boiler Baghouse: Nucla Generating Plant*, August 1975 (EPA-600/2-75-013-a [NITS PB 246 641]).

[9] Cass, R. W. and R. M. Bradway, *Fractional Efficiency of a Utility Boiler Baghouse: Sunbury Steam Electric Station*, March 1976 (EPA-600/2-76-077a [NTIS PB253 943]).

[10] Penny, C. W., *Electrostatic Effects in Fabric Filtration: Volume 1. Fields, Fabrics, and Particles (Annotated Data)*, September 1978 (EPA-600/7-78-142A [NTIS PB 288576]).

[11] Frederick, E. R., *Electrostatic Effects in Fabric Filtration: Volume 11. Triboelectric Measurements and Bag Performance*, July 1978 (EPA600/7-78.1A2B [NTIS PB 287207]).

[12] Frederick, E. R., *Electrical Effects in Particulate Matter Processes*, Filter Media Specification, Pittsburgh, 1987.

[13] Dennis, R., *et al..*, *Filtration Model for Coal Fly Ash with Glass Fabrics*, August 1977 (EPA-600/7-77-084 [NITS PB 276489]).

[14] Dennis, R., and H. A. Klemm, "Modeling Concepts for Pulse Jet Filtration." *JAPCA*, 30(l), January 1980.

[15] Leith, D. and M. J. Ellenbecker, "Theory for Pressure Drop in a Pulse-Jet Cleaned Fabric Filter." *Atm. Environment*, 14, 1980, pp. 845-852.

[16] Koehler, J. L. and D. Leith, "Model Calibration for Pressure Drop in a Pulse-Jet Cleaned Fabric Filter," *Atm. Environment*, 17(10), 1983, pp. 1909-1913.

[17] Viner, A. S., and B. R. Locke, *Cost and Performance Models for Electrostatically Stimulated Fabric Filters*, April 1984 (EPA 600/8-84-016).

[18] Northrop Services, Inc. *Fabric Filter Workshop Reference Materials, 1977 Workshop*, Air Pollution Training Institute.

[19] Vatavuk, W. M., and R. B. Neveril, "Estimating Costs of Air-Pollution Control Systems, Part XI: Estimate the Size and Cost of Baghouses," *Chemical Engineering*, March 22, 1982, pp. 153-158.

[20] Frey, R. F., and T. V. Reinauer, "New Filter Rate Guide," *Air Engineering*, 30 April 1964.

[21] Owen, M. K. and A. S. Viner, *Microcomputer Programs for Particulate Control,* June 1985 (EPA-600/8-85-025a).

[22] Dennis, R. and H. A. Klemm, Fabric Filter Model Change: Vol. I, Detailed Technical Report, February 1979 (EPA-600/7-79-043a) [NTIS PB 293551]).

[23] Viner, A. S., *et al.*, "Comparison of Baghouse Test Results with the GCA/EPA Design Model," *JAPCA*, 34(8), August 1984.

[24] ETS, Inc., Roanoke, VA.

[25] R. S. Means Company, Inc., *Means Construction Cost Data, 1998*, Kingston, MA.

[26] Vatavuk, W. M., and R. B. Neveril, "Estimating Costs of Air-Pollution Control Systems, Part II: Factors for Estimating Capital and Operating Costs," *Chemical Engineering*, November 3, 1980, pp. 157-162.

[27] Perry, Robert H., *et al.*, *Perry's Chemical Engineers' Handbook* (Fourth Edition), McGraw-Hill, New York, 1963, p. 6-20.

[28] Personal communication from Gary Greiner, ETS, Inc., to Jim Turner, Research Triangle Institute, October 24, 1986.

[29] Perry, Robert H., *et al.*, *Perry's Chemical Engineers' Handbook* (Sixth Edition), McGraw-Hill, New York, 1984.

[30] Hosokawa Mikropul Environmental System, *Cartridge Dust Collectors*, product bulletin, undated (about 1997).

EPA/452/B-02-001

Section 6

Particulate Matter Controls

EPA/452/B-02-001

Section 6

Particulate Matter Controls

EPA/452/B-02-001

Chapter 2

Wet Scrubbers for Particulate Matter

Daniel Mussatti
Innovative Strategies and Economics Group
Air Quality Strategies and Standards Division
Office of Air Quality Planning and Standards
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

Paula Hemmer
E.H. Pechan & Associates, Inc.
3622 Lyckan Parkway, Suite 2002
Durham, NC 27707

July, 15 2002

Contents

2.1 Introduction 2-4

2.2 Process Description 2-5
2.2.1 Capture Mechanisms 2-5
2.2.2 Scrubber Types 2-5
2.2.2.1 Spray Tower 2-5
2.2.2.2 Cyclonic Spray Tower 2-7
2.2.2.3 Dynamic Scrubber 2-7
2.2.2.4 Tray Towers 2-9
2.2.2.5 Venturi Scrubbers 2-9
2.2.2.6 Orifice Scrubber 2-10
2.2.2.7 Other Designs 2-11

2.3 Configuration and Operation of Venturi Scrubbers 2-12
2.3.1 Liquid Storage and Delivery System 2-12
2.3.2 Liquid Injection System 2-14
2.3.3 Venturi Throat Section 2-15
2.3.4 Collection Chamber and Mist Eliminator 2-16
2.3.5 Waste Liquid Collection and Disposal 2-17
2.3.7 Auxiliary Equipment 2-17

2.4 Design Parameters 2-18
2.4.1 System Performance 2-18
2.4.1.1 Liquid-to-Gas Ratio 2-18
2.4.1.2 Velocity and Pressure Drop 2-19
2.4.1.3 Particle Size Distribution and Loading 2-19
2.4.1.4 Waste Gas Flow Rate, Temperature, and Humidity 2-20
2.4.1.5 Residence Time 2-21
2.4.1.6 Droplet Size 2-22

2.5 System Design 2-22
2.5.1 PM Distribution and Loading 2-22
2.5.2 Collection Efficiency 2-23
2.5.3 Waste Gas Properties 2-25
2.5.3.1 Psychrometric Chart 2-27

2.5.4 Scrubber Design 2-29
2.5.4.1 Estimating Pressure Drop 2-30
2.5.4.2 Scrubber Velocity and Cross-Sectional Area 2-39
2.5.5 Consumables 2-40
2.5.5.1 Water Usage 2-40
2.5.5.2 Electrical Power Usage 2-42

2.6 Cost Analysis 2-43
2.6.1 Total Capital Investment 2-43
2.6.1.1 Purchased Equipment Cost 2-43
2.6.1.2 Direct and Indirect Installation Costs 2-48
2.6.2 Annual Costs 2-49
2.6.2.1 Direct Annual Costs 2-49
2.6.2.2 Indirect Annual Costs 2-50

2.7 Example Problem 2-52

2.8 Acknowledgments 2-58

References 2-59

2.1 Introduction

Particulate matter (PM) is the general term used for a mixture of solid particles and liquid droplets suspended in air. U.S. EPA defines PM_{10} as particle matter having a nominal aerodynamic diameter of 10 micrometer (μm) or less. $PM_{2.5}$ is defined as PM less than or equal to 2.5μm in aerodynamic diameter. In general, "coarse PM" refers to PM_{10} while "fine" PM refers to $PM_{2.5}$.

A wet scrubber is an air pollution control device that removes PM and acid gases from waste gas streams of stationary point sources. The pollutants are removed primarily through the impaction, diffusion, interception and/or absorption of the pollutant onto droplets of liquid. The liquid containing the pollutant is then collected for disposal. There are numerous types of wet scrubbers which remove both acid gas and PM. This chapter addresses the design and cost of wet scrubbers for control of PM_{10} and $PM_{2.5}$. (See Section 5.2 Chapter 1 for information regarding wet scrubbers for acid gas control.)

Collection efficiencies for wet scrubbers vary with the particle size distribution of the waste gas stream. In general, collection efficiency decreases as the PM size decreases. Collection efficiencies also vary with scrubber type. Collection efficiencies range from greater than 99% for venturi scrubbers to 40-60% (or lower) for simple spray towers [1]. Improvements in wet scrubber designs have increased collection efficiencies in the sub-micron range.

Wet scrubber systems have some advantages over electrostatic precipitators (ESPs) and baghouses. Wet scrubbers are smaller and more compact than baghouses or ESPs. They have lower capital cost and comparable operation and maintenance (O&M) costs. Wet scrubbers are particularly useful in the removal of PM with the following characteristics:

- Sticky and/or hygroscopic materials (materials that readily absorb water);
- Combustible, corrosive and explosive materials;
- Particles which are difficult to remove in their dry form;
- PM in the presence of soluble gases; and
- PM in waste gas streams with high moisture content.

Wet scrubbers have numerous industrial applications including industrial boilers, incinerators, metals processing, chemical production, and asphalt production, and fertilizer production.

The primary disadvantage of wet scrubbers is that increased collection efficiency comes at the cost of increased pressure drop across the control system [2]. Another disadvantage is that they are limited to lower waste gas flow rates and temperatures than ESPs or baghouses. Current wet scrubber designs accommodate air flow rates over 47 actual cubic meters per second (m^3/s) (100,000 actual cubic feet per minute (acfm)) and temperatures of up to 400°C (750°F). Another disadvantage is that they generate waste in the form of a sludge which requires treatment and/or

disposal. Lastly, downstream corrosion or plume visibility problems can result unless the added moisture is removed from the gas stream.

2.2 Process Description

This section presents wet scrubber designs used for the control of PM_{10} and $PM_{2.5}$ emitted from stationary point sources. Only commonly applied scrubber designs are addressed.

2.2.1 Capture Mechanisms

Particulates contact liquid droplets in wet scrubbers through several mechanisms. Impaction is the primary capture mechanism. When waste gas approaches a water droplet, it flows along streamlines around the droplet. Particles with sufficient inertial force maintain their forward trajectory and impact the droplet. Due to their mass, particles with diameters greater than 10 μm are generally collected using impaction [3]. Turbulent flow enhances capture by impaction.

Particles dominated by fluid drag forces follow the streamlines of the waste gas. However, particles that pass sufficiently close to a water droplet are captured by interception, capture due to the surface tension of the water droplet. Particles of of roughly 1.0 to 0.1 μm in diameter are subject to interception [21]. Increasing the density of droplets in a spray increases interception [1].

Very small-sized particles are subject to Brownian motion, irregular motion caused by random collisions with gas molecules. These particles are captured by the water droplet as they diffuse through the waste gas. Collection due to diffusion is most significant for particles less than 0.5 μm in diameter [1].

Capture mechaisms that are used less frequently include condensation and electrostatics. In condensation scrubbing, a gas stream is saturated with water vapor and the particle is captured when the water condenses on the particle [3]. In electrostatic scrubbing, contact is enhanced by placing an electrostatic charge on the particle, droplet, or both [2].

2.2.2 Scrubber Types

2.2.2.1 Spray Tower

The simplest type of scrubber is the spray tower. In a spray tower, particulate-laden air passes into a chamber where it contacts a liquid spray produced by spray nozzles. Towers can be placed in either vertical or horizontal waste gas flow paths. The liquid spray can be directed counter to the gas flow, in the same direction as the gas flow, or perpendicular to the gas flow. Figure 2.1 shows an example of a vertical countercurrent spray chamber. The gas flow enters at the bottom of the tower and flows upward. Water sprays downward from nozzles mounted on the

walls of the tower or mounted on an array at the tower center. Water droplets capture particles suspended in the gas flow through impaction, interception and diffusion. Droplets large enough to settle by gravity collect at the bottom of the chamber. Droplets that remain entrained in the gas stream are collected on a mist eliminator upstream of the nozzles [2]. (Section 2.3.4 discusses mist eliminators in more detail)

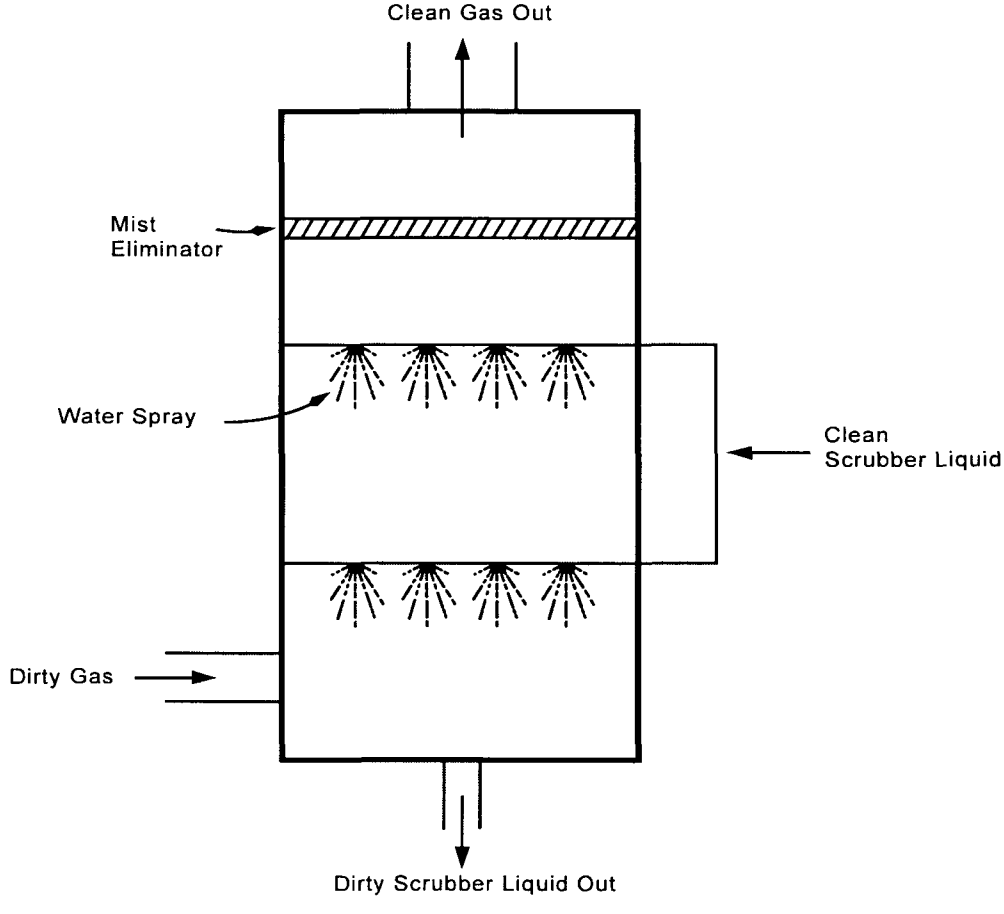


Figure 2.1: Spray Tower [4]

Spray towers rely primarily on particle collection by impaction; therefore, they have high collection efficiencies for coarse PM. Typical removal efficiencies for a spray tower can be as great as 90% for particles larger than 5 μm. Removal efficiencies for particles from 3 to 5 μm in diameter range from 60 to 80%. Below 3 μm, removal efficiencies decline to less than 50%. Spray tower applications include control of PM emissions from grinding operations, pigment

operations, and dust control in fertilizer plants. Spray towers can also be applied to control PM from asphalt plant aggregate dryers [1].

Spray towers have lower capital costs than other wet scrubbers. Also, spray towers generally have lower power consumption and are not prone to fouling, so operating costs are also lower [5]. Operating costs of spray towers increase for fine PM applications, because such systems require high liquid to gas ratios (over 20 gallons per 1000 cubic feet (gal/1000 ft^3)). Typical gas flow rates for spray towers are 1 to 47 standard m^3/s (1,500 to 100,000 standard cubic feet per minute (scfm)) [6].

2.2.2.2 Cyclonic Spray Tower

Cyclonic spray towers differ from spray tower designs in that the waste gas stream flows through the chamber in a cyclonic motion. The cyclonic motion is produced by postitioning the gas inlet tangential to the wall of the scrubbing chamber or by placing turning vanes within the scrubbing chamber. The gas inlet is tapered so that the gas velocity increases as it enters the tower. The scrubbing liquid is sprayed from nozzles in a central pipe (tangential inlet) or from the top of the tower (turning vanes) [1]. Figure 2.2 shows a diagram of a cyclonic spray tower with a tangential inlet [4]. Liquid droplets entrained in the gas stream experience a centrifugal force resulting from the rotating motion of the gas stream, causing them to migrate toward to the tower walls [2]. The droplets impact on the tower walls and fall to the bottom of the tower. Droplets that remain entrained in the waste gas can be removed with a mist eliminator.

Cyclonic spray towers have greater collection efficiencies than simple spray towers due to the greater relative velocity between the droplets and the waste gas in a cyclonic tower. Collection efficiencies for this type of scrubber are as high as 95% for particles greater than 5 μm, and from 60% to 75% for submicron particles. Typical applications are for dust control in fertilizer plants, grinding operations, and foundries [1]. Gas flow rates range from 1 to 47 m^3/s (1,500 to 100,000 scfm), and power input for a cyclonic scrubber is generally 1 to 3.5 horsepower per 1000 cubic feet per minute (hp/1000 cfm) [2, 6]. Capital costs and operation and maintenance costs are slightly higher for cycolonic spray towers due to their more complex design.

2.2.2.3 Dynamic Scrubber

Dynamic scrubbers are also known as mechanically-aided scrubbers or disintegrator scrubbers. This type of scrubber is similar to spray towers, but with the addition of a power-driven rotor that shears the scrubbing liquid into finely dispersed droplets. The rotor can be located inside the tower or outside the tower, connected by a duct. A mist eliminator or cyclonic separator removes the liquid and captured PM. Most dynamic scrubber systems humidify the waste gas upstream of the rotor to reduce evaporation and particle deposition in the rotar area [1].

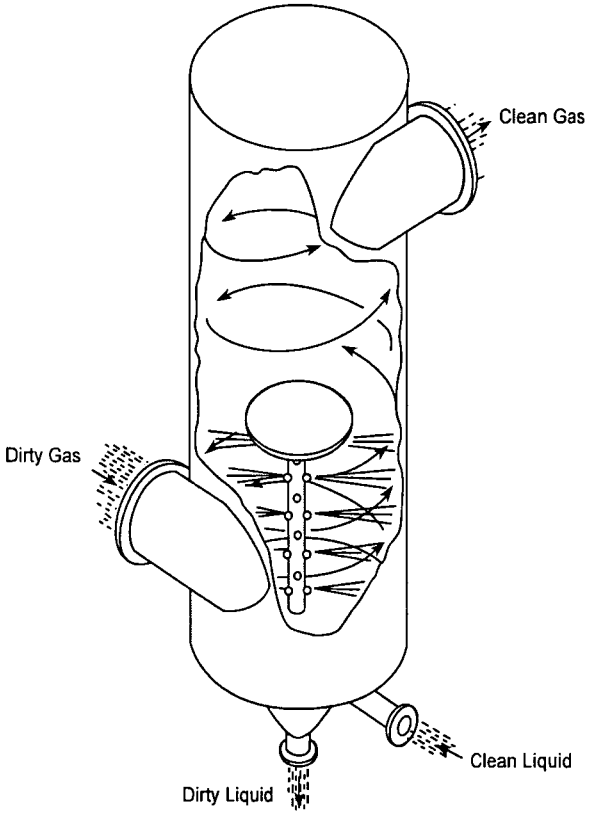


Figure 2.2: Cyclonic Spray Scrubber [3]

Dynamic scrubbers efficiently remove fine PM, but the addition of a rotar to the scrubber system increases the maintence costs. Large PM abrades the rotars and the humid gas stream corrodes them. A pretreatment device, such as a cyclone, often precedes a dynamic scrubber to remove large PM from the waste gas stream [6]. Power consumption is also high for this type of scrubber, between 4 to 10 kilowatts (kW) per 1000 acfm [6, 7]. Dynamic scrubbers generally can treat gas flow rates between 1,000 and 50,000 scfm. Collection efficiencies for dynamic scrubbers are similar to those for cyclonic spray towers [1]. Capital and O&M costs are moderately higher than costs for simple spray towers due to the rotar.

2.2.2.4 Tray Towers

Tray tower scrubbers consist of a vertical tower with several perforated trays mounted horizontally in the tower. Gas enters the tower at the bottom and travels upward through openings in the trays, while the scrubbing liquid flows from the top and across each tray. The gas mixes with the liquid flowing over the tray, providing more gas-liquid contact than in spray tower designs. The gas velocity prevents liquid from flowing down through the perforations in the tray. The impingement plates are continuously washed clean of collected particles by the flowing liquid. Tray towers are designed to provide access to each tray for cleaning and maintenance [6]. Large PM can clog the perforations, therefore, some designs place impingement baffles upstream of each perforation to remove large PM prior to the waste gas entering the opening [6]. This type of tray tower is referred to as an impingement-plate or impactor scrubber.

Tray towers do not effectively remove submicron particles, however, collection efficiencies of 97% are possible for particles larger than 5 μm [5]. Tray towers also effectively remove soluble gases; therefore they are useful when both particulate and gaseous pollutants must be removed. Typical applications include lime kilns, bagasse and bark boilers, and secondary metals industries [1, 6]. Gas flow rates for tray tower designs are generally between 1,000 to 75,000 scfm. Liquid to gas ratios are low compared to spray towers and venturi scrubbers because the scrubbing liquid is essentially static [1]. Capital and O&M costs of tray and impingement towers are moderately higher than simple spray towers.

2.2.2.5 Venturi Scrubbers

A venturi scrubber has a "converging-diverging" flow channel. In this type of system the cross-sectional area of the channel decreases then increases along the length of the channel. Figure 2.3 presents a venturi scrubber. The narrowest area is referred to as the "throat". In the converging section, the decrease in area causes the waste gas velocity and turbulence to increase. The scrubbing liquid is injected into the scrubber slightly upstream of the throat or directly into the throat section. The scrubbing liquid is atomized by the turbulence in the throat, improving gas-liquid contact. The gas-liquid mixture then decelerates as it moves through the diverging section, causing additional particle-droplet impacts and agglomeration of the droplets. The liquid droplets are then separated from the gas stream in an entrainment section, usually consisting of a cyclonic separator and mist eliminator [3]. Design, operation and cost of venturi scrubbers are the focus of this chapter and are discussed in greater detail in Section 2.3.

Venturi scrubbers are more expensive than spray tower, cyclonic, or tray tower scrubbers, but collection efficiencies for fine PM are higher. High gas velocities and turbulence in the venturi throat result in high collection efficiencies, ranging from 70% to 99% for particles larger than 1 μm in diameter and greater than 50% for submicron particles [1,6]. Increasing the pressure drop in a venturi scrubber increases the efficiency, but the system's energy demand also increases leading to greater operational costs. Capital and O&M costs are moderately higher than costs for simple spray towers.

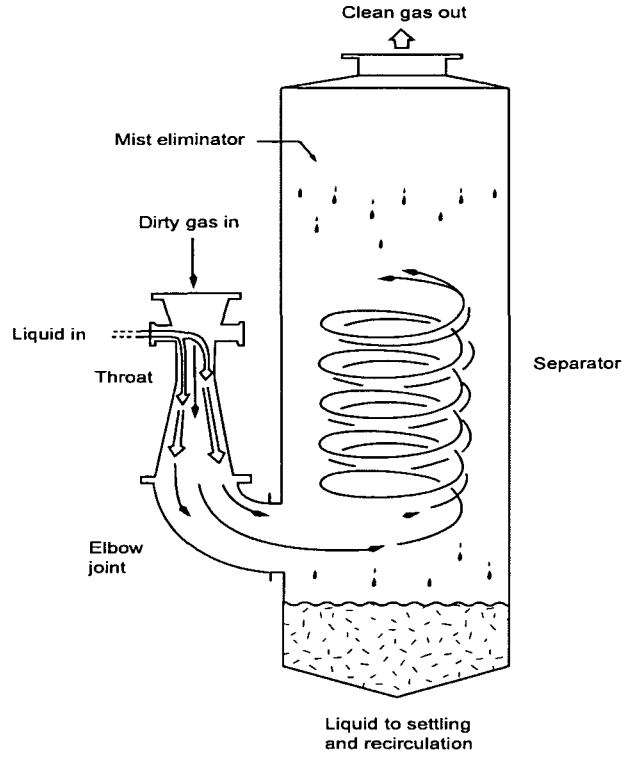


Figure 2.3: Venturi Scrubber with Cyclone Separator and Mist Eliminator [3]

2.2.2.6 Orifice Scrubber

In an orifice scrubber, also referred to as an impaction scrubber, the gas stream flows over the surface of a pool of scrubbing liquid. As the gas impinges on the water surface, it entrains droplets of the liquid. The waste gas then flows upward and enters an orifice with a narrower opening than the duct. The orifice induces turbulence in the flow which atomizes the entrained droplets. The atomized droplets capture the PM in the gas stream. A series of baffles then removes the droplets, which fall into the liquid pool below. Some orifice scrubbers have adjustable orifices to control the gas velocity. Orifice scrubbers accomodate gas flow rates up to 50,000 scfm and particle loadings up to 23 g/m^3 (10 grains per scf). The primary advantage of this type of scrubber is the elimination of a recirculation pump for the scrubbing liquid, which is a major contributor to operating costs for most scrubber designs [6]. The primary disadvantage is the difficulty of removing waste sludge. In most scrubber designs, waste continually drains from the bottom. Orifice scrubbers employ a static pool of scrubbing liquid, so waste sludge is removed with a sludge ejector, which operates like a conveyor belt. The sludge settles onto the ejector, which conveys it out of the scrubber [8].

Orifice scrubbers are not widely used, but have been applied to dryers, cookers, crushing and grinding operations, spray operations(pill coating, ceramic glazing), ventilation (bin vents, dumping operations), and material handling (transfer stations, mixing, dumping, packaging). This type of scrubber can effectively remove PM over 2 μm in diameter, with control efficiencies ranging from 80-99%. Though orifice scrubbers can be designed as high-energy units, most are built for low-energy service. Capital and O&M costs are significantly higher than costs for simple spray towers.

2.2.2.7 Other Designs

Packed tower scrubbers are towers containing a bed of packing material. The packing material provides a large wetted surface for gas-liquid contact. Scrubbing liquid is introduced at the top of the tower and flows down through the packing, coating the packing and forming a thin film. Packing materials are available in a variety of forms, each having specific characteristics with respect to surface area, pressure drop, weight, corrosion resistance, and cost. Packed towers are most often used for gas adsorption rather than PM removal, because high particle concentrations can build up on the packing and clog the tower [6]. Packed-bed scrubbers are discussed in detail in Section 5.2, Chapter 1 of the Manual, "Wet Scrubbers for Acid Gas Control".

In a condensation scrubber, the particles act as condensation nuclei for the formation of water droplets. First, the gas stream is saturated with water vapor. Steam may also be injected to further increase the humidity ratio. The injection of water vapor and/or steam creates a condition of super-saturation leading to the condensation of water on particles in the gas stream. The droplets are then removed by a conventional device, such as a mist eliminator. Condensation scrubbers can effectively remove fine PM and have collection efficiencies of greater than 99%. However, the scrubber can only remove relatively small amounts of dust due to the amount of saturation and condensation that are capable of being maintained in the gas stream. Condensation scrubbers are generally intended to be used downstream of another scrubber that has already removed particles larger than 1 μm in diameter. Condensation scrubbing is a relatively new technology and has limited commercial availability [6]. Its most frequent application is to hazardous waste or medical waste incinerators.

Charged scrubbers enhance removal by placing an electrostatic charge on the water droplets, particles, or both prior to entering the scrubber. These scrubbers usually employ a conventional scrubber design, such as a spray tower. The particulates can be negatively or positively charged, with droplets given the opposite charge. Wet ESPs are similar devices which combine an ESP with flowing liquid to continuously clean electrostatic plates [1].

Commercially available wet scrubbers employ a wide range of design variations, including several hybrids of technologies. For example, a few manufacturers offer venturi scrubbers with multiple throats. Other manufacturers combine wet scrubber devices with other types of particulate removal, such as a baghouse or ESP.

2.3 Configuration and Operation of Venturi Scrubbers

There are three basic types of venturi scrubbers. The primary difference between the configurations is the energy required for scrubbing the waste gas and moving it through the unit. In a conventional venturi, an external device, typically an induced draft (ID) fan, transfers energy to the liquid-gas stream. The fan can be located either upstream or downstram of the venturi unit. The basic venturi scrubber design is very efficient at removing PM_{10} from both a cost and performance perspective. In a "jet" or "eductor" venturi, pressurized scrubbing liquid is injected into the throat. This type of venturi operates at at low pressure drops, generally a few inches of water column. A jet venturi has a lower collection efficiency for fine PM than a conventional venturi. A "high energy" venturi provides increased collection efficiency for fine and submicron PM. A high energy system utilizes a large ID fan to create a high gas side pressure drop, 30 inches of water column or greater. This greatly increases the waste gas velocity prior to entering the throat and results in high collection efficiency. However, capital costs and electrical power requirements for high energy systems are much higher than a conventional venturi.

Venturi systems can be installed on either horizontal or vertical waste gas flow paths. They can be purchased as packaged, skid mounted units or as field errected units. Materials of construction for system components include carbon steel, stainless steel, duplex alloys, FRP or lined steel. The waste gas properties determine which material is most appropriate for a given application. More than one type of material can be incorporated into a venturi system if necessary.

Figure 2.4 presents the schematic of a venturi system. The basic system components of a venturi scrubber are:

- liquid storage system and delivery system;
- liquid injection system;
- venturi throat section;
- collection chamber with a mist eliminator;
- waste liquid collection system and disposal;
- instrumentation and controls; and
- auxiliary equipment.

Each of these components are discussed in the following sections.

2.3.1 Liquid Storage and Delivery System

The liquid storage and delivery system consists of a recirculation tank, pump, filters, valves, piping, pressure gauges, and flow meters. Most systems are designed as recycle systems, meaning the spent scrubbing liquid is recirculated through the scrubber system. Since the scrubbing liquid is recycled, the solids content of the liquid increases as PM is collected. The concentration of solids in a recycle system must be maintained below a design limit or the spray characteristics of the system cannot be maintained. To reduce the solids concentration, a

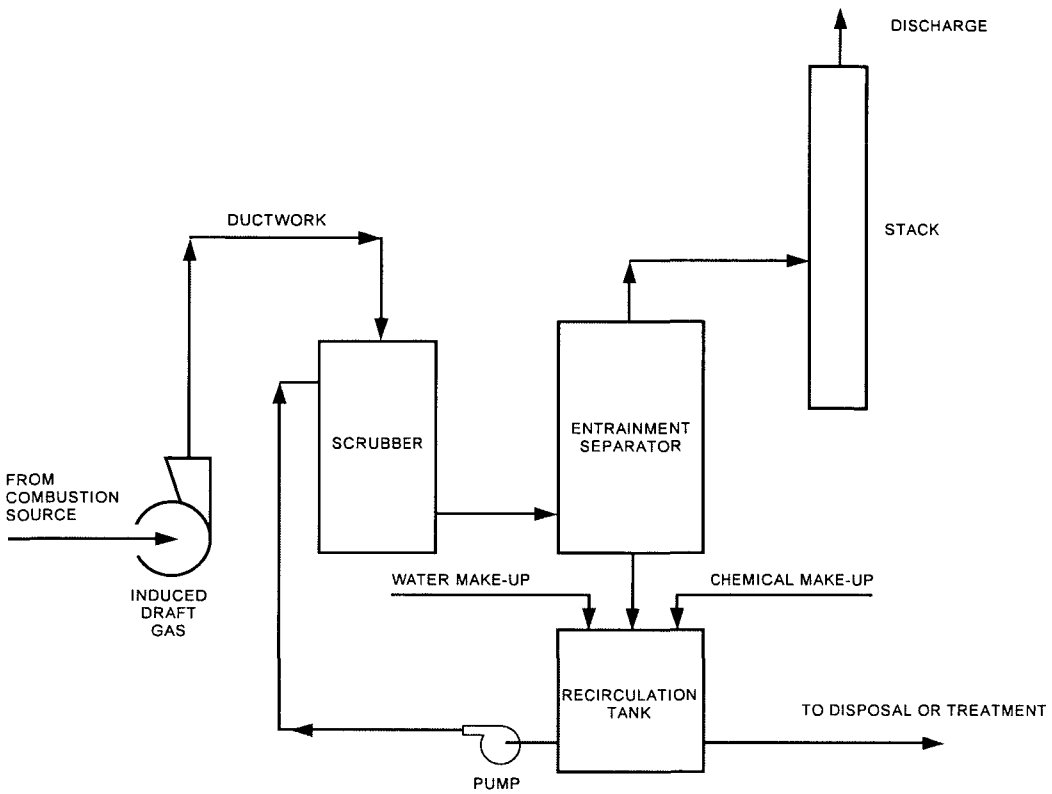


Figure 2.4: Schematic of Venturi Scrubber System

portion of the liquid is bled from the system and fresh scrubbing liquid is added. The volume of "make up" liquid also includes the volume of water that is lost due to evaportaion by hot waste gas.

The tank must be sized to provide continuous operation and minimize frequent changing of the liquid. A sensor in the tank monitors the level of liquid. An automated system for adding liquid can be incorporated into the scrubber design. However, oversizing the tank and and automating the addition of make up water increases the capital cost and complexity of the system. The increase in capital cost must be weighed with the O&M cost for operating the liquid storage and delivery system manually.

Solids can be removed from the scrubbing liquid using several different methods. In one method, the scrubbing liquid is gravity fed to a set of filters located upstream of the pump to remove solids. The filter is generally constructed out of stainless steel and is removable for cleaning and replacement. A backup filter and set of isolation valves are often included in the system to

facilitate cleaning of the filters while operating the system. Other solids removal methods include liquid cyclones or settling tanks.

After leaving the tank, the scrubbing liquid flows to a pump to increase the pressure and flow rate to the values required for proper operation of the venturi system. Pressure gauges and flow meters downstream of the pump monitor the scrubbing liquid flow and pump operation. A feedback control system can be added to the system to automate control of the pump.

2.3.2 Liquid Injection System

The injection system design promotes mixing of the waste gas and scrubbing liquid in the venturi. There are two basic systems for injecting scrubbing liquid into a venturi system: open pipe (also referred to as "wet approach") and spray nozzles. The injection systems are generally located in the waste gas duct, directly upstream of the venturi throat section. In both systems, the liquid is injected in the same direction as the waste gas stream. Most injection systems are constructed out of stainless steel or other non-corrosive material.

In an open pipe system, several small diameter pipes feed the scrubbing liquid into the duct section. The pipes inject the liquid tangentially, along the duct walls or radially against baffle plates. The water flows downward, covering the walls of the duct. The piping system is designed so that the entire surface area of the section is flooded with the scrubbing liquid. This ensures that there is no dry/wet transitional zone. Dry/wet areas lead to a build up of solids on the duct wall that interferes with the operation of the scrubber. The dust laden gas enters the scrubber vertically from the top and immediately hits the film of scrubbing water. Some separation of the PM from the waste gas takes place in this area. In the throat section, the waste gas stream becomes very turbulent and the scrubbing liquid is sheared to form a dispersion of droplets. Open pipe systems have lower capital and O&M costs than spray nozzles due to their simpler design.

Spray nozzles systems are sometimes referred to as "jet venturis"or "eductor venturis". These systems inject liquid through nozzles to create a fine droplet spray pattern. The droplets can be produced either pneumatically or hydraulically using specially designed nozzles heads. While spray nozzles improve mixing between the scrubbing liquid and the waste gas, they generally have higher capital and operating costs than open pipe systems due to the higher pump horsepower required for this type of system.

In jet venturis, the nozzles can be attached to the wall of the duct or can be located in the duct cross section. For throat areas greater than 1 foot in width, a spray nozzle must be located in the center of the duct to ensure adequate liquid-PM contact [1]. Nozzles can be constructed out of stainless steel or more specialized materials such as stellite and ceramic [9]. Because nozzles are prone to plugging and abrasion in high PM load conditions, this type of system requires clean liquid feed to avoid clogging [6]. High temperatures and gas velocities can damage the nozzles, consequently, they should be designed to be removable for cleaning and replacement. See Hueman for examples of nozzles and spray patterns [10].

2.3.3 Venturi Throat Section

The throat consists of the narrowest portion of the converging-diverging venturi section. This is where the velocity and turbulence of the waste gas is greatest. In the throat, the waste gas shears the scrubbing liquid into a high density distribution of fine droplets. These droplets collect PM primarily through impaction. The waste gas and scrubbing liquid then pass into the diverging section where the velocity decreases, causing more impaction and liquid agglomeration. Some of the energy imparted to the liquid-gas system is recovered in the diverging section as gas pressure. However, the overall energy of the system decreases due to friction and other mechanical losses. This loss is measured as the decrease in pressure across the venturi converging-diveging section, referred to as the pressure drop, ΔP.

There are a number of different throat configurations that are commercially available, including fixed throat, variable throat, variable annual throat multiple throat, and multiple stage. Figure 2.5 presents schematics of a fixed throat, a variable throat with a damper, an annuar variable throat, and a multiple throat.

A fixed throat venturi is the simplest type of venturi. The throat section can be circular or rectangular, depending on the duct shape of the current waste gas system. Rectangular throats are generally limited to a width of approximately 10 in. due to mixing considerations. Circular fixed throats are typically used in high-pressure applications. [1]

A variable throat venturi changes the cross-sectional area of the throat through the use of an adjustable damper. There are many different damper designs including conical plugs, discs, and blades. The venturi throat area is increased or decreased by the dampers when the waste gas inlet conditions change. This allows the venturi to maintain the same throat velocity and, therefore, the same collection efficiency even with fluctuations in the waste gas flow conditions. A control system can be incorporated into the variable throat device to automatically adjust to changes in the gas flow. Automatic throat adjustement is typically used where flow conditions vary widely and frequent adjustments are required. The complexity of a variable throat increases the capital and O&M costs of the venturi.

A multiple throat venturi is a set of parallel venturis in one duct section. The throats are created by fixed length rods or flat plates located across the cross-section of the duct. The number of throats and width of the throats vary between designs. Most systems have throats on the order of 1 to 2 inches in width [1]. The throats can be designed as fixed or variable. This type of design not only decreases the throat area but increases the wetted area of the venturi, resulting in higher collection efficiency. Multiple throat venturi systems work efficiently in low pressure applications.

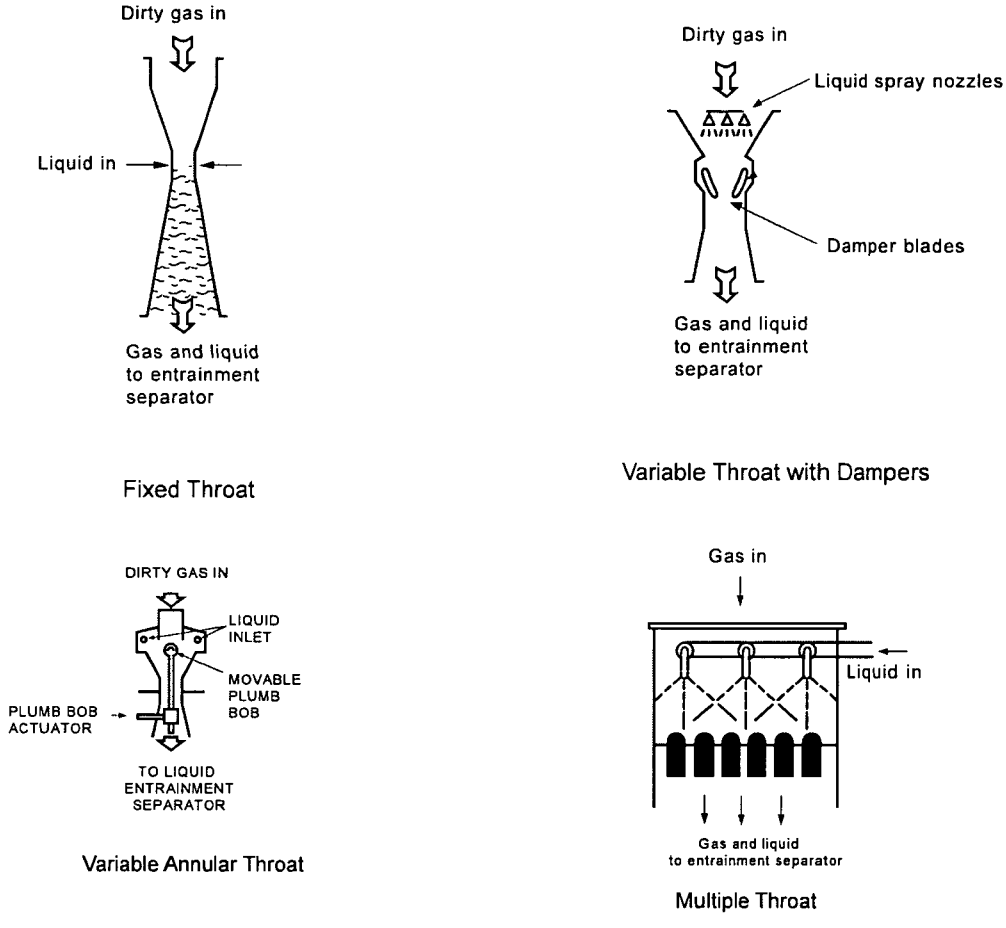


Figure 2.5: Venturi Throat Configurations [9, 11]

A multiple stage venturi is simply a series of venturi scrubbers. This type of scrubber system can have two basic forms. The first type is a set of venturi throats in series which share a common collection chamber and liquid injection system. The throats are comprised of sets of vanes or baffles within a duct section. The second type of multiple stage venturi places a set of stand-alone venturi scrubber systems in series, each with its own collection chamber and liquid injection system.

2.3.4 Collection Chamber and Mist Eliminator

After passing through the venturi section, the scrubbing liquid and waste gas enter a collection chamber that separtes the the entrained liquid-PM droplets from the waste gas. A portion of the droplets settle via gravity to the bottom of the chamber. The droplets which remain entrained in the

waste gas are generally removed with a mist eliminator. An outlet is located at the bottom of the collection chamber to drain the liquid-PM waste from the chamber. The collection chamber can be a simple tower design, a tower with interior baffles, or a cyclone. Baffle and cyclone designs enhance the removal of liquid and PM from the waste gas stream using impaction as well as gravity. However, they result in larger waste gas pressure drops and cost more than a simple tower.

Mist eliminators remove between 90% and 99% of the liquid droplets from the waste gas stream [1]. There are two basic designs, chevron and mesh pad mist eliminators. The droplets collect and coalasce on the chervon blades or mesh. When the droplets become large enough, they fall by gravity or capillary action. Mesh pad mist eliminators can be clogged by the particluate, therefore, chevron designs are more frequently applied. Pressure drop across a mist eliminator is low, 0.5 to 1.0 inches of water column (in w.c.) All mist eliminators require periodic washing to remove buildup of PM [1].

In many venturi systems, the direction of flow through the venturi section is downward. After passing through the venturi, the flow turns horizontal prior to entering the collection chamber. Turning the high velocity flow results in a decrease of the waste gas pressure and abrasion of the elbow joint due to friction. To minimize these effects, many venturi designs utilize a "flooded elbow", an elbow duct with a pool of water or scrubbing liquid at the bottom. The liquid in the elbow duct decreases friction and collects heavy droplets in the waste gas stream.

2.3.5 Waste Liquid Collection and Disposal

Spent scrubber liquid drains from the bottom of the chamber to the recirculation tank. A portion of the liquid is bled from the system to limit the solids concentration to 20% to 30% by weight [12]. The effluent is in the form of a slurry with high solids content and, in some applications, may contain hazardous material. Nonhazardous effluent can be treated in an existing wastewater system or by an off-site contractor. The liquid is separated from the solid waste, then the wastewater is reused or discharged. The remaining solid or sludge is landfilled if non-toxic and inert. Waste gas containing hazardous PM requires treatment and/or hazardous waste disposal of the sludge. The annual cost for hazardous waste disposal is a direct function of the wastewater flow rate, suspended solids content, and the hazardous nature of the waste (i.e. flammable, toxic, corrosive, etc.). Disposal costs include the cost of laboratory analysis, transportation costs, and the cost of treatment, destruction, landfill, or other disposal method. Due to the high variablity of disposal costs, theses costs are not included as part of annual costs in Section 2.6.2.

2.3.7 Auxiliary Equipment

An ID fan is generally required to make up for pressure lost in a low energy venturi system. Fans marginally increase the capital cost of the venturi system but greatly increase O&M costs due to the electrical power usage and maintenance requirements of the fan. Power input for the fan ranges from 3 to 12 hp/1000 cfm [2]. The ID fan can be placed either upstream or downstream

of the venturi wet scrubber. The fan placement is dependant on the waste gas characteristics. For instance, applications with high particulate loading place the fan downstream to avoid pitting of the fan blades. To reduce corrosion and pitting, the fan can be manufactured out of stainless steel or coated with special materials.

High temperature gas streams evaporate large amounts of scrubbing liquid, causing a decrease in the number of droplets generated by the venturi. Therefore, high temperature applications must either cool the gas before entering the venturi, or spray a greater volume of water into the venturi. The gas can be cooled with a quencher, which sprays water into the gas stream. When the water evaporates, the temperature of the gas stream decreases. Quenchers marginally increase the capital and operating costs of the system.

Venturi systems may require additional equipment to a fan and quencher. A PM collection system such as a hood may be required to collect the waste gas from the source. An upstream PM collection device such as a cyclone may be required to remove large PM and prevent abrasion of the venturi components. A stack may be required after the venturi to release the waste gas into the atmosphere at the a specified height. Lastly, a reheating device may be required to increase plume buoyancy (height) for better dispersion and to decreases plume visibility caused by condensing water.

2.4 Design Parameters

2.4.1 System Performance

The parameters affecting the overall performance of a wet scrubber are:

- Particle size distribution and loading;
- Waste gas flow rate, temperature and humidity;
- Gas velocity and pressure drop;
- Liquid-to-gas (L/G) ratio;
- Droplet size; and
- Residence time.

Each of these parameters are briefly discussed below.

2.4.1.1 Liquid-to-Gas Ratio

The liquid-to-gas ratio (L/G) is the volume of liquid injected per volume of waste gas treated. In general, a higher L/G ratio increases collection efficiency since the density of droplets across a given cross-section of the venturi is higher. Liquid flow rates between 7 and 10 gal/1000 ft^3 give optimum performance. L/G ratios in this range produce fairly constant collection efficiencies given a constant pressure drop [13]. L/G ratios of greater than 10 gal/1000 ft^3 do not improve the

scrubber performance significantly. While increasing the L/G ratio increases collection effciency, operating costs are increased as well due to greater scrubbing liquid and pump usage.

2.4.1.2 Velocity and Pressure Drop

Increasing the relative velocity between the gas and the liquid droplets increases the momentum of the particulate, allowing smaller particles to be collected by impaction. The relative velocity can be increased by narrowing the throat, injecting the scrubbing liquid counter-current to the flow, or spraying the liquid into the throat. However, increasing relative velocity generally increases the pressure drop, energy demand, and operating costs for the scrubber [5]. High energy venturis increase the gas velocity using an induced draft fan upstream of the venturi. These systems have much higher operating costs than low energy venturis due to the higher fan power.

The smaller the cross-sectional area of the throat, the greater the increase in the gas velocity obtained. The highest gas velocity occurs at the center point of the narrowest cross-section, generally ranging from 45 to 150 meters per second (m/s) (150 to 500 feet per second (ft/s)). The resulting pressure loss of the gas stream across the venturi is in the range of 10 to 80 in w.c. In general, increasing the pressure drop above 45 in w.c. does not significantly increase the removal efficiency for conventional venturi designs [1]. Venturi designs optimize the cross sectional area of the throat to provide high gas velocities while minimizing pressure drop. In addition, the diverging section of the venturi is designed to recover the most pressure. Diverging sections are designed to decrease the waste gas velocity to between 30 and 15 m/s (100 and 50 f/s) [13, 24]. At this speed, turbulent losses are minimized and the greatest amount of energy recovery is achieved.

2.4.1.3 Particle Size Distribution and Loading

The performance of a given scrubber type is highly dependant on the size distribution of the PM in the waste gas stream. The size distribution determines the capture mechanism, impaction, interception or diffusion, that dominates. Most wet scrubber designs rely almost exclusively on inertial impaction for particulate collection. Particles smaller than 0.1 μm are captured primarily through diffusion mechanisms [5]. Figure 2.6 presents the approximate collection efficiency of a venturi wet scrubber as a function of particle size. Note that the efficiency decreases exponentially with particle size.

PM loading, also called dust loading, is the mass of PM per unit volume in the waste gas at the inlet of the scrubber. As PM loading increases, the L/G ratio must increase to maintain the same collection efficiency. Figure 2.7 presents the L/G as a function of particulate loading. Higher PM loading also results in higher solids content of the recycled scrubbing liquid. In order to maintain the solids content, a greater volume of scrubbing liquid must be bled from the system as waste and a greater volume of clean scubbing liquid must be added to the system. Higher PM loadings increase the operating costs of the system due to increased pump usage, scrubbing liquid usage, and waste liquid disposal. Applications with high PM loadings also require more maintenance, as particles can cause plugging of orifices and wear to parts such as nozzles and fans.

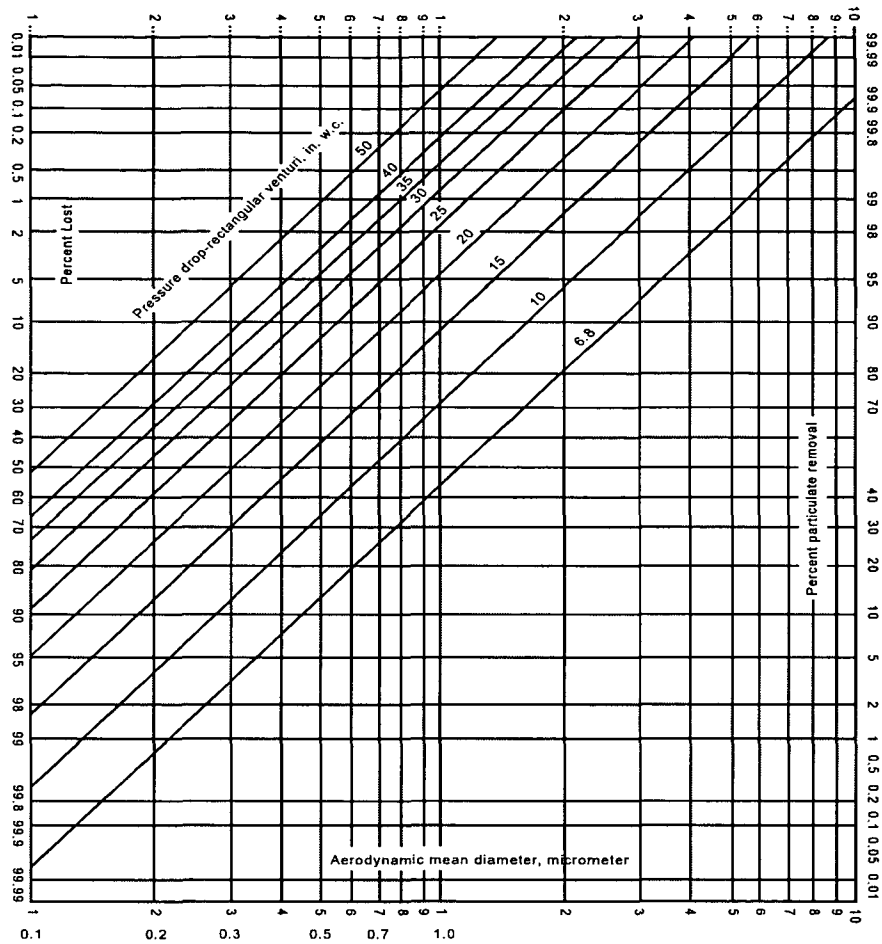


Figure 2.6: Approximate Collection Efficiencies of Particles in the 0.1-10 μm Diameter Range at Venturi Pressure Drops in the 6.8-50 in. H_2O Range [2]

2.4.1.4 Waste Gas Flow Rate, Temperature, and Humidity

The waste gas flow rate is the most important sizing parameter in a wet scrubber. The higher the waste gas flow rate, the larger the venturi system and volume of scrubbing liquid required to treat the waste gas. Wet scrubbers operate at lower gas flow rates than baghouses or ESPs because of the liquid injection. New low energy venturis can accomodate air flow rates of up to 95 m/s (300,000 acfm). Jet venturi systems are generally limited to approximately 3 m/s (10,000 acfm) and multi-throat and high energy venturi systems are limited to approximately 47 m^3/s (150,000 acfm).

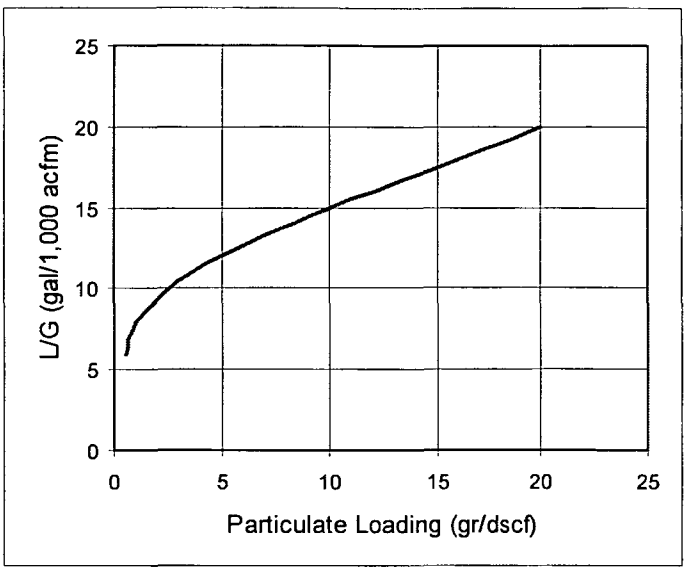


Figure 2.7: Liquid to Gas Ratio as a Function of Particle Loading [16] [22]

The waste gas temperature and humidity also impacts the venturi design. When air passes through a wet scrubber, water evaporates, which increases humidity and cools the gas stream. The amount of evaporation is determined by the inlet temperature and humidity. High evaporation rates will increase the L/G ratio required by the system. For PM applications, wet scrubbers are generally limited to a temperature range of (50°F to 700°F) due to evaporation. A quencher may be needed for higher temperature applications. High temperature affects the material used to manufacture the scrubber components.

2.4.1.5 Residence Time

Increasing the length of the throat and the diverging section, increases the contact time between the liquid and the PM suspended in the waste gas. For example, a venturi with a throat length of 1 foot and a velocity of approximately 450 ft/sec has a contact time of 1/450 of a second. This is minimal time for mixing and contact between the liquid and waste gas. For a cylindrical throat, a throat length to throat diameter ratio of 3:1 is the minimum recommended [13]. For high energy systems, it is recommended that the length of the diverging section of the throat to be at least 4 times the width of the throat in order to have sufficient contact time. [1]

2.4.1.6 Droplet Size

There is an optimum droplet size for maximizing collection of PM. Smaller droplets have a larger surface area to volume ratio, therfore, they capture more particles per volume of liquid injected. However, if the droplet size becomes too small, the momentum of the waste gas can be imparted to the droplets which decreases the relative velocity between the droplet and particles. Lower relative velocity results in lower collection efficiency. Wet scrubbers control the size of droplets using several techniques. In scrubbers using preformed droplets, such as spray towers, the droplet size is determined by the type of nozzle and the system operating conditions. In dynamic scrubbers, the speed of the rotor and L/G controls the droplet size. In venturi scrubbers, the droplet size is controlled by the L/G and the gas velocity in the throat.

2.5 System Design

2.5.1 PM Distribution and Loading

The design and performance of a given scrubber type is highly dependant on the properties of the particulate matter in the waste gas stream. As discussed in section 2.4, the most critical properties are particle size distribution and PM loading. There is a wide distribution of both particle sizes and loading across industrial sources. Source-specific PM distribution and loading determine the most efficient PM collection device on a case-by-case basis.

Because particles have various shapes and densities, particle size is usually expressed as the aerodynamic diameter. The aerodynamic diameter of a particle is the diameter of a sphere with the density of water that settles in still air at the same rate as the particle in question. The size distribution is usually measured using a cascade impactor, which separates particles by their aerodynamic diameter onto plates. The mass of particles on each impaction plate is measured [5]. Figure 2.8 presents a typical PM particle size distribution, the cumulative mass verses the particle size. Notice it is a log-normal distribution. PM from industrial sources, generally have a log-normal distribution.

The geometric mean diameter is the aerodynamic diameter of the 50th percentile of PM on a mass basis (also referred to as mass median particle diameter). By definition, the standard deviation of a log-normal distribution is the ratio of the 84th percentile to 50th percentile particle sizes on a mass basis:

$$\sigma = \frac{d_{84}}{d_{50}} \tag{2.1}$$

where σ = standard deviation,
d_{50} = mass fraction of the 50th percentile particle size, and
d_{84} = mass fraction of the 84th percentile particle size

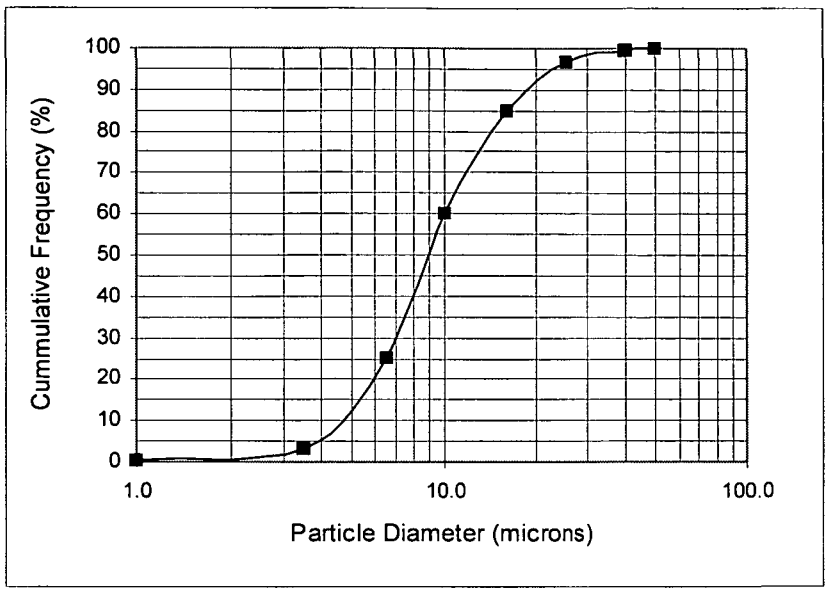


Figure 2.8: Particle Size Distribution for a PM Source

2.5.2 Collection Efficiency

Collection efficiency is the amount of PM removed from the gas stream by the wet scrubber. This efficiency can be expressed a number of ways including the efficiency of a single water droplet, the efficiency of the scrubber on a mass basis, or the efficiency of the scrubber on a particle size basis. Each of these efficiencies is defined below.

The collection efficiency of a single droplet η_{drop} is defined as the area swept free of aerosol particles divided by the projected cross-sectional area of the droplet.

$$\eta_{drop} = \frac{area\ swept\ free\ of\ particles}{droplet\ cross\text{-}sectional\ area} \qquad (2.2)$$

The overall efficiency of the scrubber is usually related to η_{drop} by an empirical exponential equation. However, most of the parameters in the equation are generally not available at the study level. Therefore, this chapter does not present this method of determing efficiency.

Collection efficiency on a mass basis is given by:

$$\eta_m = \frac{\dot{m}_i - \dot{m}_o}{\dot{m}_i} = \frac{l_i - l_o}{l_i} \tag{2.3}$$

where η_m = overall collection efficiency on a mass basis,
$\dot{m}_{i,o}$ = total mass flow rate at inlet, outlet, and
$l_{i,o}$ = particle loading at inlet, outlet.

Collection efficiency as a function of particle size distribution is the cumulative collection efficiency for each particle size range given by:

$$\eta_d = \sum_0^j \eta_j m_j \tag{2.4}$$

where η_d = overall collection efficiency,
η_j = fractional efficiency for jth particle diameter range,
m_j = mass fraction of jth particle diameter range, and
j = number of particle diameter ranges.

The mass fraction is defined by:

$$m_j = \frac{\text{mass of particles in range of intrest}}{\text{total mass}}$$

Collection efficiency on a mass basis is generally higher than the collection efficiency on a particle basis. This is because the larger size particles, which are generally more massive, tend to be collected at higher efficiencies than smaller diameter particles. Therefore, it is more common to express efficiency on a particle size basis than a mass basis.

Penetration is defined as the fraction of particles that pass through the collection device. Penetration is directly related to collection efficiency by:

$$Pt_d = 1 - \eta_d \tag{2.5}$$

where Pt_d = overall penetration of collection device,
η_d = overall efficiency on a particle size basis.

Design equations for PM removal devices often utilize the cut diameter, the diameter at which the collection efficiency of the scrubber is 50%.Cut diameter is a characteristic of the control device and operating conditions, not the particle size range. It is determined experimentally using the particle collection efficiency and particles size distribution data collected for a given device and set of operating conditions.

2.5.3 Waste Gas Properties

The physical and chemical properties of the waste gas are generally given to the vendor to properly size the scrubber system and choose appropriate materials for fabrication. These properties have a direct impact on capital and annual costs associated with the scrubber as well as impacting design. The designer needs information on conditions at both the inlet and outlet including:

Q	=	volumetric flow rate;
V	=	volume;
T	=	temperature;
P	=	pressure; and
θ_{H2O}	=	fractional moisture content.

Waste gas properties are measured at the scrubber inlet to the scrubber. Outlet waste gas properties change as a function of the evaporation rate of the scrubbing liquid. This discussion uses the following subscripts:

m	=	dry air and water vapor mixture;
a	=	dry air;
wv	=	water vapor;

As hot flue gas passes through the scrubber, a portion of the water in the scrubbing liquid evaporates. The temperature of the flue gas decreases, the moisture content and humidity increases, and the volume decreases. For design purposes, the conditions of the waste gas at the scrubber outlet are assumed to be at the saturation point.

Evaporation through the scrubber is generally modeled as a direct evaporative cooling process (also referred to as an adiabatic saturation process). In this process, the non-saturated (dry) air is cooled by transferring the air's sensible heat to the water vapor as latent heat. The total quantity of heat energy contained in the air, the enthalpy, remains constant. In addition, the system is assumed to conserve mass and the waste gas is assumed to behave as an ideal gas. Under these assumptions, the mass flow rate of dry air through the system remains constant and the difference between the inlet and outlet mass water vapor is the mass evaporated in the scrubber.

This subsection presents a procedure for estimating the waste gas outlet conditions and the water evaporated in the scrubber. The procedure uses a psychrometric chart. The thermodynamic variables on the chart are defined in the following paragraph. Commercial computer programs are now available that directly calculate the variables on the chart.

Moisture content refers to the volume of water vapor in a given volume of gas. Moisture content is given by the equation:

$$\theta_{H2O} = \frac{V_{wv}}{(V_a + V_{wv})} \tag{2.6}$$

The humidity ratio, or absolute humidity is the ratio of the mass of dry air to water vapor:

$$\omega = \frac{m_v}{m_a} \tag{2.7}$$

where ω = humidity ratio.

This differs from relative humidity, which is the ratio of the water vapor partial pressures at the current conditions and at saturation for a given temperature. The relationship between moisture content, θ_{H2O}, and the humidity ratio, ω, for an ideal gas is given as:

$$\theta_{H2O} = \frac{\frac{\omega}{MWa}}{\left(\frac{1}{MWwv} + \frac{\omega}{MWa}\right)} \approx \omega \frac{MWwv}{MWa} \tag{2.8}$$

where MW_{wv}, MW_a = molecular weight of water vapor and air, respectively.

In order to use the psychometric chart, the waste gas properties must be known at standard conditions (or the conditions of the chart, if different from standard). The properties are determined for standard conditions using the Ideal Gas Law:

$$\frac{P_1 V_1}{T_1} = \frac{P_2 V_2}{T_2} \tag{2.9}$$

where temperature is in units of degrees Kelvin. Assuming constant pressure, the volume of the waste gas at standard temperature can be calculated as:

$$V_2 = V_1 \frac{T_2}{T_1} \quad \text{or} \quad Q_2 = Q_1 \frac{T_2}{T_1} \tag{2.10}$$

where Q_2 = waste gas outlet volume flow rate (air and water vapor, Q_m) at standard conditions.

The relationship between the mass and volume of an ideal gas can be calculated from it's volume at standard temperature and pressure using the equation:

$$\dot{m} = Q\frac{MW}{V_{mole}} \tag{2.11}$$

where $\dot{m}$ = mass flow rate,
MW = molecular weight, and
V_{mole} = volume of one mole of air
= 22.4 liters per gmole (385 ft^3 per lbmole).

Moisture content (or humidity ratio) is used to calculated the mass flow rate of dry air and water vapor at the inlet using Equation 2.11:

$$\dot{m}_{wv(in)} = \left[Q_{m(in)} \times \theta_{H2O(in)}\right]\frac{MW_{wv}}{V_{mole}}$$
$$\dot{m}_{a(in)} = \left[Q_{m(in)}\left(1 - \theta_{H2O(in)}\right)\right]\frac{MW_{wv}}{V_{mole}} \tag{2.12}$$

Note this is the mass flow rate at standard conditions. The inlet humidity ratio, $\omega_{(in)}$ at standard conditions can now be calculated as well using Equation 2.7.

2.5.3.1 Psychrometric Chart

Figure 2.9 is an example of a psychrometric chart at standard atmospheric conditions, 14.7 pounds per square inch (psi). The dry bulb temperature is typically the x-axis of the chart. The dry bulb temperature is the temperature of the waste gas mixture of water vapor and air, or the temperature measured by an ordinary thermometer. The chart's y-axis has a scale for both the humidity ratio and relative humidity. The y-axis to the far right on the chart is the saturation curve, representing a relative humidity of 100%. Lines that slant upward to the left are generally lines of constant wet bulb temperature, specific volume, and enthalpy. The specific volume of the air-water vapor mixture, or humid volume, is the volume of the mixture per unit mass of dry air.

Use the chart to determine outlet conditions at saturation. First, fix a point on the chart based on the dry bulb temperature and humidity ratio. Follow the corresponding line of constant enthalpy (adiabatic saturation line) to the saturation curve on the left. This is the location for reading the outlet conditions of the waste gas. This is the point at which the maximum water from the scrubber has evaporated into the waste gas. Note that the mass of the dry air is conserved during the process, $m_{a(in)} = m_{a(out)}$.

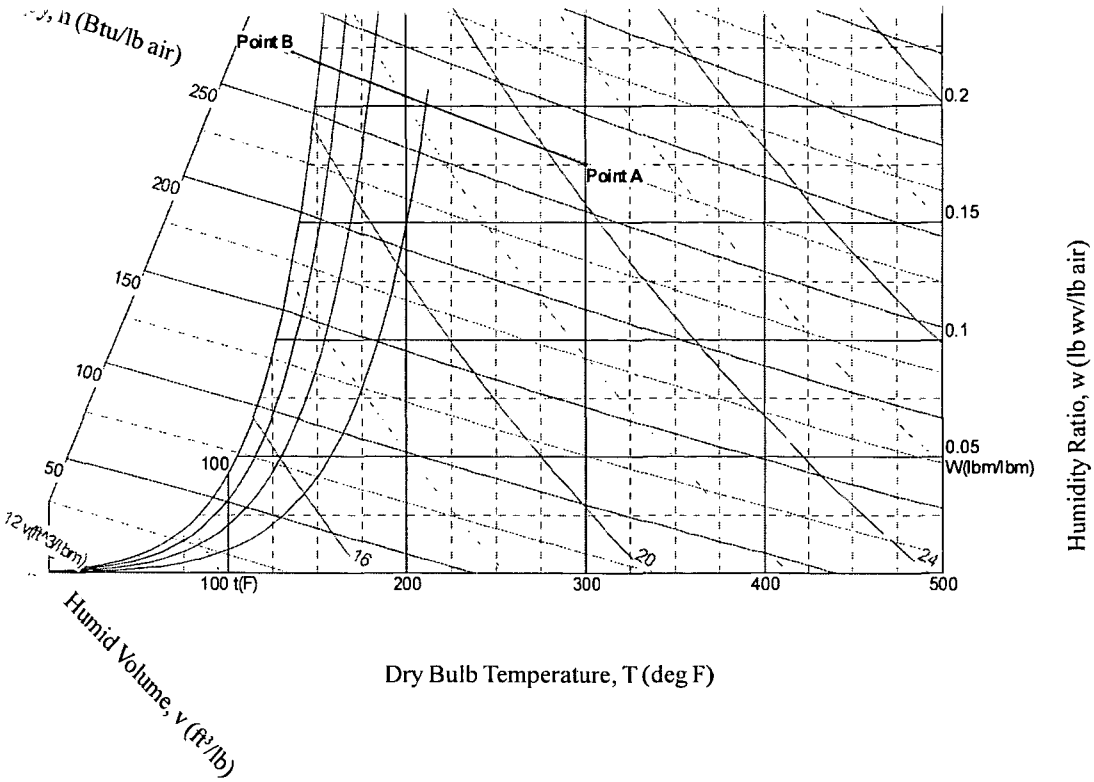


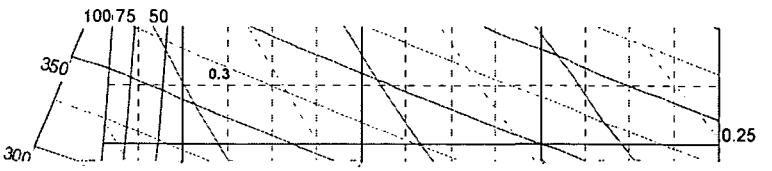
Figure 2.9: Psychometric Chart at Standard Pressure, 14.7 psi
(Chart developed using Akton Psychrometrics, Akton Associates, Inc.)

The volumetric flow rate of the waste gas at the outlet of the scrubber can now be determined. Using the dry air mass flow rate and the humid volume, v, read from the chart, the volume of saturated air per unit mass of dry air, obtained from the chart, the volumetric flow rate is:

$$Q_{m(out)} = \nu \dot{m}_a \tag{2.13}$$

where ν = humid volume of saturated air

Relative Humidity (%)



Enthal...

The outlet humidity ratio can be used to calculate the mass of water vapor in the waste gas at the outlet of the scrubber:

$$\dot{m}_{wv(out)} = \omega_{(out)} \ \dot{m}_a \tag{2.14}$$

Additional water must be added to the scrubber system to make up for the water lost to evaporation. By determining the difference between the inlet and outlet humidity of the waste gas stream, the volume of water evaporated can be calculated. The mass of water evaporated through the scrubber is:

$$\dot{m}_{wv(evap)} = \dot{m}_{wv(out)} - \dot{m}_{wv(in)} \tag{2.15}$$

The volume flow rate of make up water is given by:

$$Q_{wv(evap)} = \frac{\dot{m}_{wv(evap)}}{\rho_{H2O}} \tag{2.16}$$

2.5.4 Scrubber Design

In general, wet scrubber manufacturers guarantee a collection efficiency for a specific design. This collection efficiency is dependent on both the liquid to gas ratio, gas velocity in the venturi and pressure drop. The L/G ratio and pressure drop impact the O&M cost while the venturi gas velocity impacts the capital cost. (See Section 1 Chapter 2 for a discussion of capital and annual costs.) Therefore, there is no straightforward design approach. An iterative approach must be utilized which looks at both capital and annual costs. The relationship between L/G ratio, gas velocity in the venturi, and pressure drop is shown in Figure 2.10 for a venturi scrubber.

A number of methods for estimating venturi design parameters have been developed by various researchers. This chapter presents four of the most common methods used by designers. For more theoretical discussions of scrubber design equations, such as calculating pressure drop for a specific particle diameter, refer to References [1, 2, and 4]

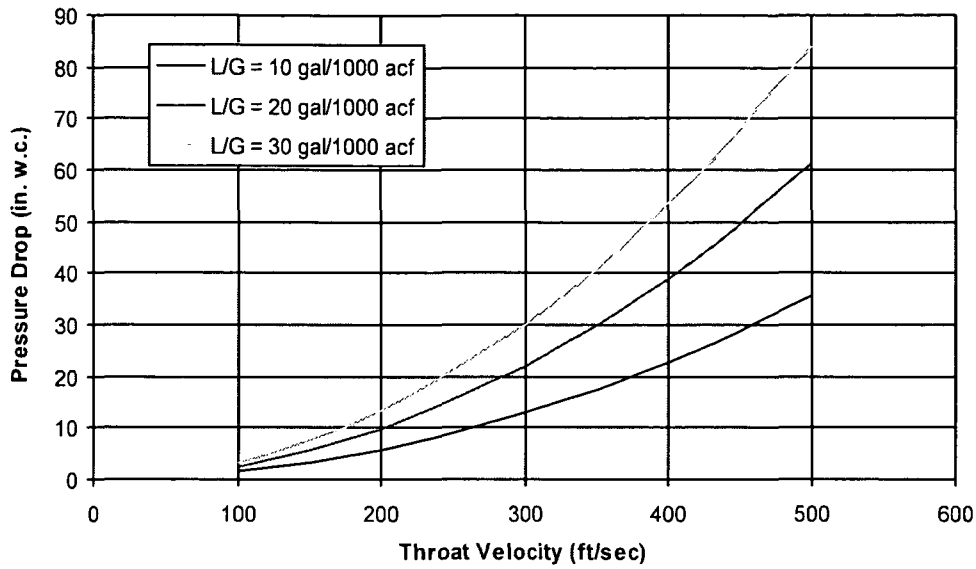


Figure 2.10: Relationship between Pressure Drop, Throat Velocity and L/G

2.5.4.1 Estimating Pressure Drop

2.5.4.1.1 Pressure Drop Equations

Most pressure drop equations for venturi scrubbers are of the form:

$$\Delta P = k\, v^2 \rho_g \left(L/G\right) \tag{2.17}$$

where ΔP = pressure drop across venturi,
v = throat velocity,
ρ_g = gas density,
L/G = liquid to gas ratio, and
k = correlation factor for a specific scrubber design.

One of the more widely accepted equations for estimating pressure drop across a venturi scrubber was published by Calvert [24]. The pressure drop is given as:

$$\Delta P = 5.4\times10^{-4}\, v^2 \rho_g \left(L/G\right) \tag{2.18}$$

where ΔP = pressure drop across venturi in inches of water (in H_2O),
v = throat velocity in feet per second (ft/s),
ρ_g = saturated gas stream density in pounds per cubic feet (lb/ft^3), and
L/G = liquid to gas ratio in gallons per 1000 cubic feet (gal/1000 ft^3).

The Calvert equation predicts pressure drop reasonably well at moderate liquid to gas ratios. At ratios between 3 gal/1000 ft^3 and 10 gal/1000 ft^3 the equation was found to perform well, but at or above 12 gal/1000 ft^3 the equation over predicts the pressure drop by 80% or more [13].

A model for pressure drop published by Hesketh is also widely used [16]. The model is based upon a correlation of experimental data obtained from many different venturi scrubbers. Hesketh's equation for pressure drop is given by:

$$\Delta P = \frac{v^2 \rho_g A^{0.133}}{507}\left(0.56 + 0.125\left(L/G\right) + 0.0023\left(L/G\right)^2\right) \tag{2.19}$$

where ΔP = pressure drop across venturi in inches of water (in H_2O),
v = throat velocity in feet per second (ft/s),
ρ_g = saturated gas stream density in pounds per cubic feet (lb/ft^3),
A = the cross-sectional area of the throat in sqiuare feet (ft^2), and
L/G = liquid to gas ratio in gallons per 1000 cubic feet (gal/1000 ft^3).

This equation is often simplified to:

$$\Delta P = \frac{v^2 \rho_g A^{0.133}\left(L/G\right)^{0.78}}{1270} \tag{2.20}$$

Hesketh experimentally determined a relationship between pressure drop and collection efficiency [2]. Based on the collected data, Hesketh concluded that the venturi is essentially 100% efficient for particles greater than 5 μm. He developed a correlation between pressure drop and penetration for particles sizes below this value, given by:

$$Pt = \frac{C_i}{C_o} = 3.47\,\Delta P^{-1.43} \tag{2.21}$$

where C_i and C_o = concentration of particles <5 μm at inlet and outlet of the venturi on a mass basis, and
ΔP = pressure drop, in H_2O.

Figure 2.6 presents pressure drop for a rectangular venturi scrubber. [2]

2.5.4.1.2 Contact Power Theory

Contacting power is defined as the energy dissipated per unit volume of gas treated. The contact power theory was developed by Lapple and Kamack [17] and extended by Semrau [18,19]. It states that all scrubbers give the same degree of particle collection at the same level of power consumption regardless of how the power is obtained, either from the gas phase pressure drop, the liquid phase atomization, or mechanical means. It is often referred to as the "Equivalent Energy Theory"

The total contact power, P_T, given by:

$$P_T = P_G + P_L + P_{mech} \tag{2.22}$$

where P_T = total contact power,
P_G = power due to pressure drop of gas passing through the scrubber,
P_L = power due to the scrubber liquid atomization, and
P_{mech} = power due to mechanical devices to increase contact, i.e., a rotor.

Contacting power is determined from the friction loss across the wetted portion of the scrubber [5]. Pressure loss due to gas stream kinetics is assumed to be negligible.

P_G is the contacting power from the gas stream energy input, generally expressed as horsepower per 1000 cubic feet per minute under actual conditions, (hp/1000 acfm). For most wet scrubbers, P_G dominates the total contact power equation. It is estimated from the measured pressure drop across the scrubber as:

$$P_G = 0.157\Delta P \tag{2.23}$$

where ΔP = pressure drop across venturi, in H_2O.

The contacting power from the liquid stream energy input, P_L, is also expressed as hp/1000 acfm. It is based on the liquid to gas ratio and given by:

$$P_L = 0.583 p_L \left(L/G\right) \tag{2.24}$$

where p_L = liquid inlet pressure in pounds per square inch (psi), and
L/G = liquid to gas ratio in gallons per cubic feet (gal/ft^3).

Table 2.1 presents the operating parameters for several scrubber types.

Table 2.1: Operating Parameters for Particulate Wet Scrubbers, [2, 12]

Scrubber Type	Pressure Drop (in. H_2O)	L/G Ratio (gal/1000 acf)	Liquid Pressure (psig)	Gas Velocities (ft/sec)	Cut Diameter (μm)
Spray Tower	0.5-3	0.5-20	10-400	10	2-8
Cyclonic	2-10	2-10	10-400	105-140b	2-3
Venturi	10-150	2-20	0.5-2	90-400c	0.2

Contacting power is correlated to the scrubber collection efficiency, η. This correlation is often expressed as the number of "transfer units", a dimensionless number defined by the equation:

$$N_t = \ln\left(\frac{1}{1-\eta}\right) \tag{2.25}$$

The number of transfer units for a given contacting power depends on the scrubber type and the characteristics of the particulate matter. For a given scrubber and particulate type, the relationship between transfer units and contact power is:

$$N_t = \alpha P_T^{\beta} \tag{2.26}$$

where α and β are emperical coefficients which are characteristic of the scrubber type and particulate being collected. Table 2.2 presents the coefficients for various scrubber types. [2, 14] The pressure drop associated with that collection efficiency can then be calculated.

Table 2.2: Contact Power Scrubber Correlation Parameters, [2, 14]

Aerosol	Scrubber	α	β
Lime kiln dust	Venturi and cyclonic spray	1.47	1.05
Prewashed lime kiln dust	Venturi, pipe line, and cyclonic spray	0.915	1.05
Talc dust	Venturi	2.97	0.362
Talc dust	Orifice and pipe line	2.7	0.362
Phospheric acid mist	Venturi	1.33	0.647
Foundry cupola dust	Venturi	1.35	0.621
Open hearth steel furnace	Venturi	1.26	0.569
Talc dust	Cyclone	1.16	0.655
Ferrosilicon furnace	Venturi and cyclonic spray	0.870	0.459
Odorous Mist	Venturi	0.363	1.41

2.5.4.1.3 Performance Curves

A performance curve is the scrubber's collection efficiency as a function of the particle diameter at a specified pressure drop. The curves are specific to a given venturi design. In general, multiple curves are presented for various pressure drops. Performance curves must be obtained from vendors. Figure 2.6 is an example of a performance curve for a rectangular venturi [2].

The overall collection efficiency must be calculated to obtain the total penetration of all PM. The overall collection efficiency is the sum of the fractional collection efficiencies in each particle size range. This calculation is presented in Section 2.5.2, Equation 2.4. The collection efficiency for each size range is read off a performance curve. The fractional collection efficiency is the mass fraction of the size range multiplied by the collection efficiency for that range. Summing these values gives the cumulative collection efficiency, η_d, at a specific pressure drop. The total penetration, P_t, can be calculted from η_d using quation 2.5.

Since each pressure drop has it's own curve, the total penetration at several pressure drops must be calculated using the above procedure. Then, the total penetration, Pt_d, is plotted for each pressure drop as shown in Figure 2.11. The design pressure drop across the scrubber is read from the graph based on the required total penetration.

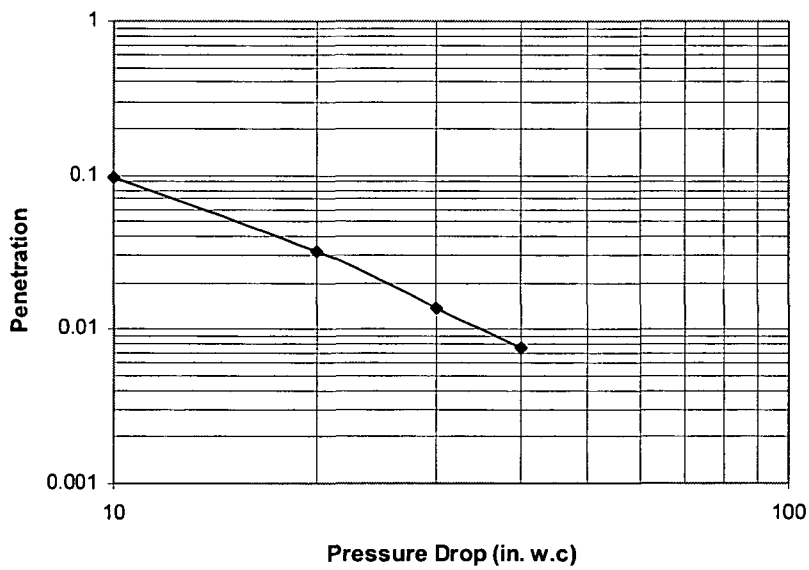


Figure 2.11: Penetration vs. Pressure Drop as Determined from Performance Curves

2.5.4.1.4 Calvert Cut Diameter Method

The method assumes that the most significant design parameter for a wet scrubber is the particle diameter that is collected at 50% efficiency or the "cut diameter" [20]. Utilizing this approaches requires a log-normal particle size distribution. It relates the cut diameter to the overall collection efficiency and the size distribution parameters.

The scrubber penetration is modeled as an exponential function of the aerodynamic diameter given by:

$$Pt_j = e^{\left(-A\, d_j^{\,B}\right)} \tag{2.27}$$

where Pt_j = penetration of the jth particle diameter,
d_j = diameter of the jth particle size, and
A, B = emperical constants.

B is a constant that is scrubber specific. B is equal to 2.0 for venturi, packed bed, and tray wet scrubbers and B is equal to 0.67 for cyclonic wet scrubbers.

The overall penetration is given by:

$$Pt_d = \int Pt_j\, m_j \tag{2.28}$$

where Pt_d = overall penetration,
Pt_j = penetration of the jth particle diameter size, and
m_j = mass fraction of the jth particle diameter size.

The Calvert Cut Diameter approach uses the cut ratio defined as:

$$Cut\ Ratio = \frac{d_{cut}}{d_{50}} \tag{2.29}$$

where d_{cut} = the required cut diameter, and
d_{50} = the mass median aerodynamic diameter.

Section 2.5.1 PM Distribution and Loading presents calculations for the mass median particle diameter and standard deviation of a log-normal distribution.

The design requirement for overall penetration and the standard deviation of the distribution define the cut diameter. Cut diameter can be obtained graphically using vendor-specific cut diameter performance curves. Figure 2.12 and Figure 2.13 from the equipment guide published by Calvert Environmental Equipment, Inc., presents penetration as a function of the standard deviation of the distribution and cut ratio. Note that Figure 2.13 is the same as Figure 2.12 but it assumes B =2.0. The required cut ratio is read from the curve and used to calculate cut diameter.

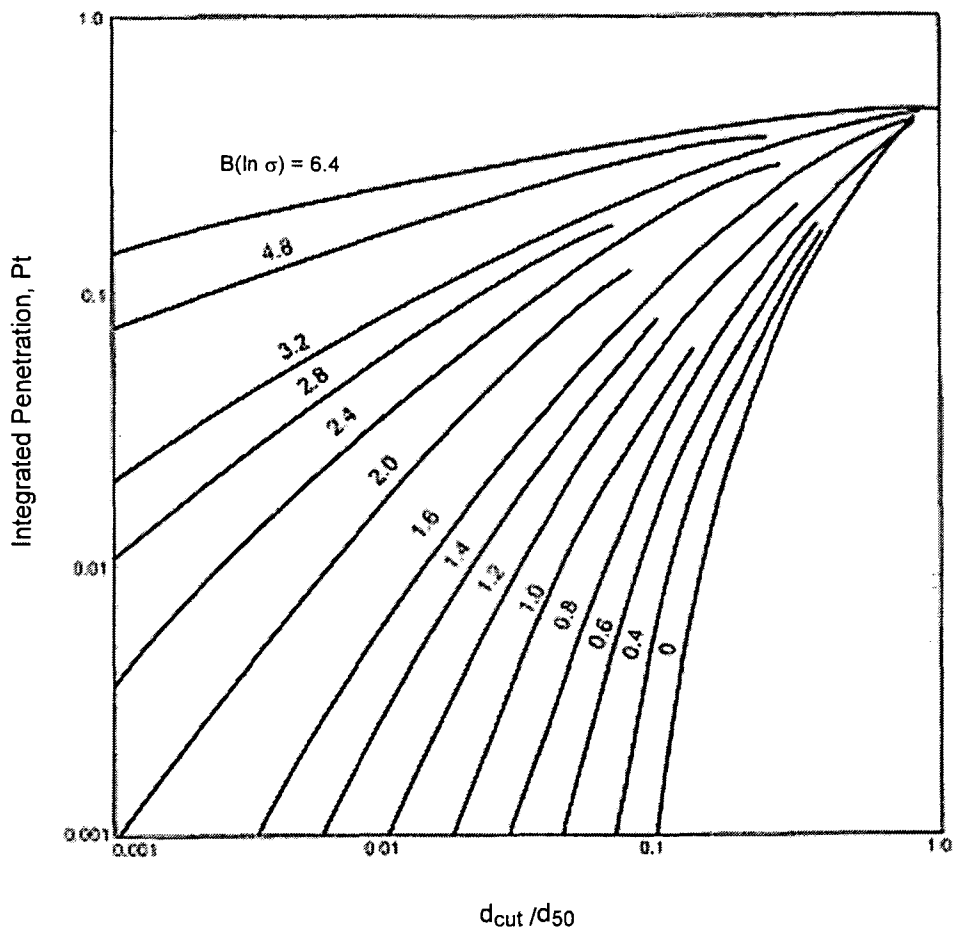


Figure 2.12: Cut Diameter as a Function of Cut Ratio and Standard Deviation of the Particle Size Distribution [20]

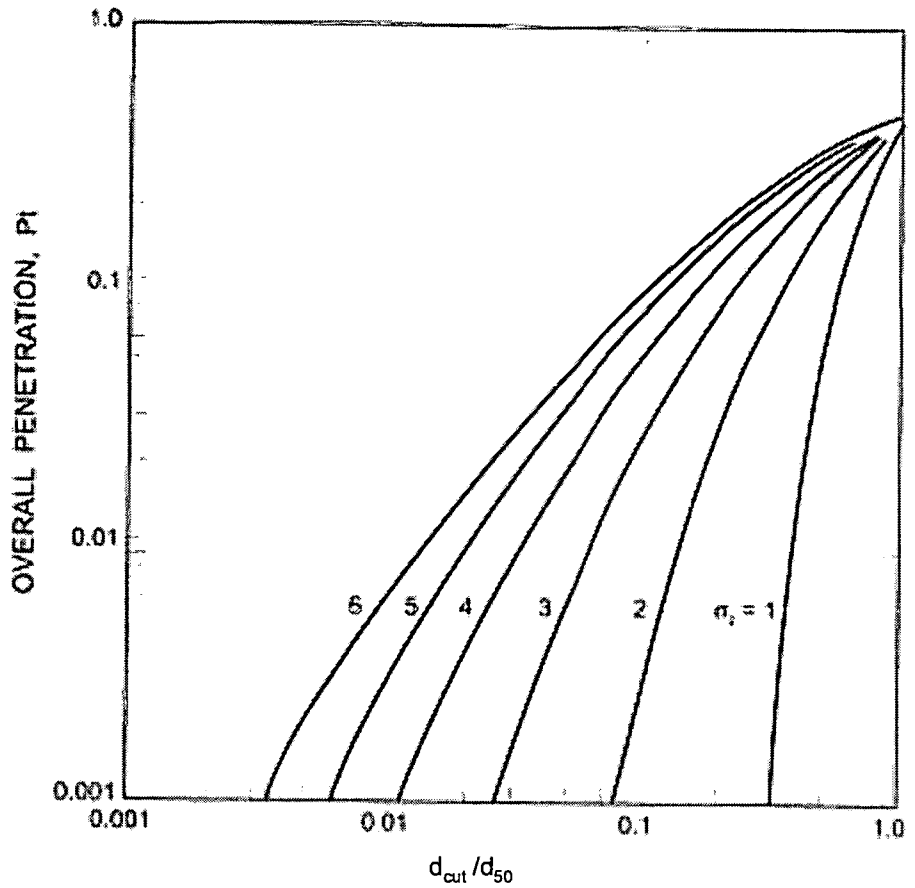
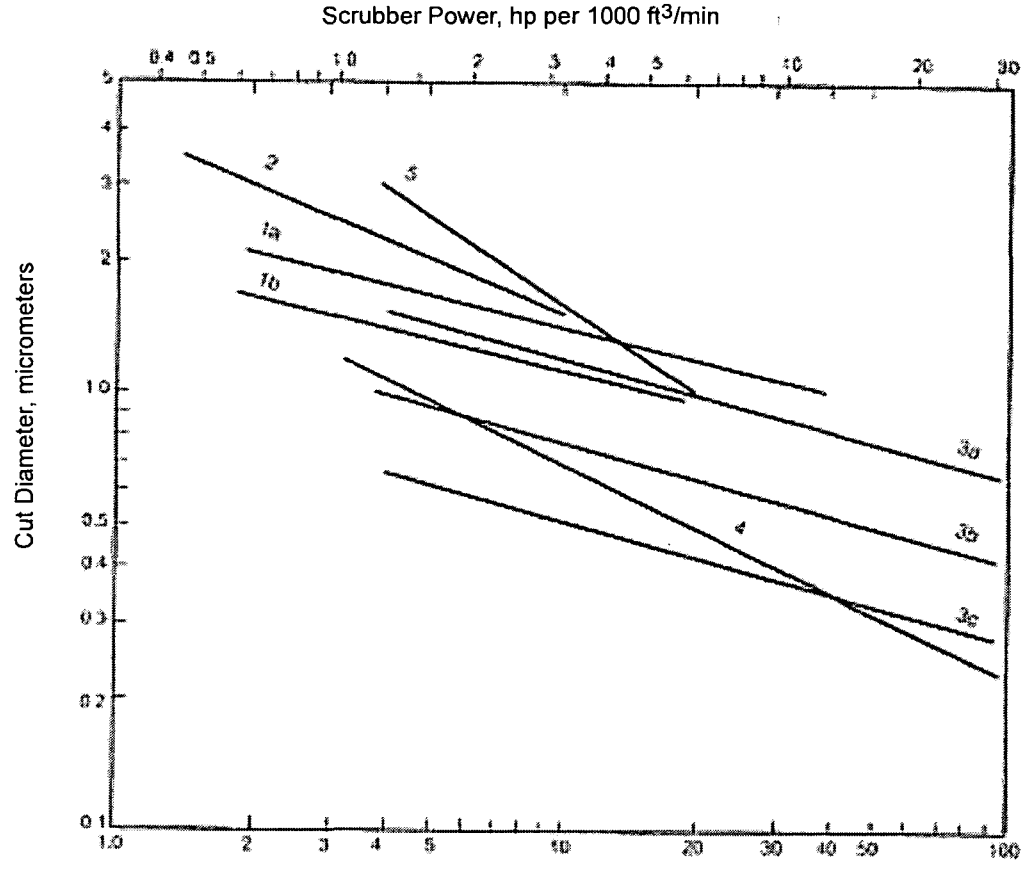


Figure 2.13: Cut Diameter as a Function of Cut Ratio and Standard Deviation of the Particle Size Distribution for B = 2.0, [20]

Once cut diameter is known, the pressure drop and scrubber power, horsepower per 1,000 acfm of gas scrubbed, can be read from another vendor-specific performance curve. An example of this type of performance curve is given by Figure 2.14. It presents the pressure drop and scrubber power for several types of scrubbers. The power requirement of this figure assumes a fan efficiency of 50%. To obtain power requirements at other efficiencies, use the following relationship:

$$Power_{\eta_2} = Power_{\eta_1} \frac{\eta_1}{\eta_2} \tag{2.30}$$



1a. Sieve-plate column with foam density of 0.4 g/cm3 and 0.2-in hole dia. The number of plates does not affect the relationship much. (Experimental data and mathematical model.)
1b. Same as 1a except 0.125-in hole-dia.
2. Packed column with 1-in rings or saddles. Packing depth does not affect relationship much. (Experimental data and mathematical model.)
3a. Fiberous packed bed with 0.012-in dia. Fiber – any depth. (Experimental data and mathematical model.)
3b. Same as 3a except 0.004-in. dia. Fibers
3c. Same as 3a except 0.002-in. dia. Fibers
4. Gas atomized spray.
5. Mobile bed with 1-3 stages of fluidized hollow plastic spheres. (Experimental data from pilot plant and large-scale power plant scrubbers)

Figure2.14: Scrubber Power and Pressure Drop as a Function of Cut Diameter, [23]

2.5.4.2 Scrubber Velocity and Cross-Sectional Area

Any of the four approaches discussed previously can be used to estimate the pressure drop that is required to attain a given collection efficiency. This pressure drop determines the required gas velocity in the venturi throat. Throat velocity depends on:

- mixture of gas and scrubbing liquid and L/G ratio,
- turbulence,
- distribution of liquid-gas mixture and PM across the throat, and
- hydraulic losses through the throat.

Throat velocity is limited by the acceptable gas velocities in downstream scrubber equipment such as the mist eliminator.

There are theoretical equations available for calculating throat velocity, however, most manufacturers determine throat velocity experimentally. The throat velocity and cross-sectional area can be estimated using a modified Bernoulli equation: [12]

$$v_t = \frac{Q_m}{A_t} = C\sqrt{\frac{\Delta P}{\rho_{sat}}} \tag{2.31}$$

where v_t = velocity at the throat,
A_t = cross-sectional area of the throat,
Q_m = maximum actual volumetric air flow rate,
ρ_{sat} = density of the gas at saturation, and
C = constant.

C is a function of the L/G ratio. A relationship between C and L/G was developed for venturis with a 30° converging section, a 10° to12° diverging section, and a gas density of 0.06 lb/ft^3 [12]:

$$C = 1060\, e^{(-0.279\, L/G)} \tag{2.32}$$

Increasing the gas density above 0.075 causes the value of C, and hence v_t, to increase rapidly due to increased resistance of the gas.

The throat cross-sectional area is calculated from the scrubber inlet and throat velocities:

$$A_t = A_i \frac{v_t}{v_i} \tag{2.33}$$

where A_t and A_i = area of the throat and inlet, respectively.

From the cross-sectional areaof the throat, the dimensions of the throat can be estimated. The diameter for a circular throat and the width of a rectuangular throat can be calculated as:

$$d_t = \sqrt{\frac{4A_t}{\pi}} \quad \textit{for circular throat}$$
$$d_t = \sqrt{A_t} \quad \textit{for rectangular throat} \tag{2.34}$$

The length of the throat and diverging section of the venturi is optimized for pressure recovery. For optimal pressure recovery, the length of the throat is generally on the order of 3 times the throat diameter (or width) and the length of the diverging section is genrally 4 times the throat diameter (or width).

$$l_t = 3\, d_t \tag{2.35}$$
$$l_{div} = 3\, d_t$$

The flow path of the gas exiting the venturi is often turned 90° prior to entering the particle separator. Due to the high velocity of the gas, the pressure drop from turning the flow can be high. The radius of the elbow duct section must be sufficiently large to minimize the pressure drop across the joint. See Section 2, Chapter 1 Hoods Ducts and Stacks for sizing and costing of duct sections. Design of cyclones and mist eliminators is beyond the scope of this document. Several of the references discuss design of these components including [2, 3, and 4].

2.5.5 Consumables

2.5.5.1 Water Usage

Most wet scrubbers systems recirculate the scrubbing liquid. In order to decrease the solids content of the scrubbing liquid, part of the liquid is bled from the system and fresh water is added. Venturi scrubbers typically have peak solids concentrations of 20 to 30% [1]. A higher PM loading of the gas stream requires a higher bleed rate resulting in a greater volume of liquid waste and higher disposal costs.

The mass flow of particulate matter into the scrubber liquid is:

$$\dot{m}_{PM} = \eta\, L_{PM}\, Q_i \tag{2.36}$$

where $\dot{m}_{PM}$ = mass flowrate of PM,
η = overall collection efficiency of the scrubber,
L_{PM} = PM loading at the inlet, and
Q_i = waste gas flow rate at the inlet.

Using the density of water, 1.0 kg/l (8.3 lb/gal), and the design solids concentration, a bleed rate for the scrubber liquid can be calculated as:

$$Q_{bleed} = \frac{\dot{m}_{PM}}{f_{solids}\ \rho_{H2O}} \tag{2.37}$$

where Q_{bleed} = bleed rate, and
f_{solids} = mass fraction of solids in recirculation water.

The total flow rate of water required by the system, Q_T, is the sum of the water evaporated and the bleed water given by:

$$Q_{T(H2O)} = Q_{wv(evap)} + Q_{bleed} \tag{2.38}$$

The total water consumed annually is given by:

$$V_{T(H2O)} = Q_{T(H2O)}\ t \tag{2.39}$$

where $V_{T(H2O)}$ = annual volume of water consumed, and
t = scrubber operating time per year.

For a jet venturi system that uses a pneumatic spray system, the air usage must also be included as a consumable.

2.5.5.2 Electrical Power Usage

From the pressure drop across the system, the required fan brake horsepower can be calculated using the following equation from Section 1:

$$HP_{fan} = \frac{\Delta P\, Q_i}{6356\, \eta_{fan}} \tag{2.40}$$

where HP_{fan} = fan brake horsepower, hp,
η = efficiency of the fan, and
ΔP = pressure of the fan, in w.c.

The brake horsepower required for pump to recirculate the scrubbing liquid through the system is calculated similarly as:

$$HP_{pump} = \frac{\Delta P_{pump}\ {}^{L}\!/_{G}\ \frac{Q_i}{1{,}000}\ \gamma}{3952.6\ \eta_{pump}} \tag{2.41}$$

where HP_{pump} = pump brake horsepower, hp,
η_{pump} = efficiency of the fan,
ΔP_{pump} = pressure of the pump, feet w.c.,
L/G = liquid to gas ratio, gal/1000 ft^3,
Q_i = flow rate at inlet, acfm, and
γ = specific gravity of the scrubbing liquid.

For a jet venturi system, the pump or compressor requirements for the pneumatic or hydraulic nozzle system must also be included in the energy consumption calculations. Table 2.3 presents typical pump and fan requirements for various scrubbers.

Table 2.3: Pump and Fan Energy Requirements (1,000 cfm; η=90% at μm), [13]

Type of unit	Liquid flow (gpm)	Pressure (psig)	Pump hp	ΔP of gas (in.wg)	Fan hp	Relative hp/ 1,000cfm
Wet Cyclone	10	60	0.91	8	2.5	3.41
Venturi	10	20	0.233	15	3.94	4.27
Jet Venturi	40	70	3.28	0	0	3.28

2.6 Cost Analysis

The cost estimation methodology presented here provides a tool to estimate study-level venturi wet scrubber capital and annual costs. The cost equations and factors for venturi scrubbers are based on the method given in Estimating Costs of Pollution Control Equipment. [12] The reader should not be surprised if vendor quotes are obtained that differ from these estimates by as much as ±25 percent since they represent study level costs. Actual selection of the most cost-effective option should be based on a detailed engineering study and cost quotations from system suppliers. The costs are in 2002 dollars. Costs can be adjusted to other years using the Chemical Engineering Cost Index or the VAPCCI index for venturi scrubbers.

The cost equations apply to industrial sources of PM_{10} and $PM_{2.5}$ with air flow rates between 100 acfm and 200,000 acfm. Extrapolation to flow rates beyond those presented is not appropriate. Overall collection efficiency ranges from 97% to 99.9%. The waste gas is assumed to have nominal values for PM distribution, PM loading rates, temperatures, and moisture content.

2.6.1 Total Capital Investment

Total Capital Investment (TCI) includes costs associated with purchasing the venturi unit and direct and indirect costs associated with installing the unit. The equation for TCI is given by:

$$TCI = PEC + DC + IC \tag{2.42}$$

where PEC = purchased equipment costs
DC = direct installation costs
IC = indirect installation costs.

In general, installing a venturi wet scrubber does not require construction of buildings, site reparation, offsite facilities, land, and working capital. A more detailed discussion of TCI can be found in Section 1, Chapter 2 of this Manual.

2.6.1.1 Purchased Equipment Cost

The Purchased Equipment Cost (PEC) of a venturi wet scrubber system is the sum of the costs of the venturi equipment, instruments and controls, taxes, and freight. The last three items generally are taken as percentages of the equipment costs. Table 2.4 gives typical values for instruments and controls, taxes, and freight.

Table 2.4: Cost Estimates for Instrumentation, Taxes and Freight

Item	Percentage of Equipment Cost
Instruments and Controls	10%
Taxes	3%
Freight	5%

Venturi equipment cost equations were developed by performing a least squares regression of cost data provided by three vendors. Costs were provided for three types of package venturi systems; fixed throat venturi, jet venturi, and variable throat venturi. Package systems were assumed to include the following basic equipment:

- venturi,
- liquid injection system,
- cyclone, and
- mist eliminator.

Additional equipment required for the venturi but not included in the cost estimate include the following:

- recycle pump,
- ID fan,
- piping and valves, or
- basic instrumentation and controls

The equipment cost of the packaged venturi system varies in direct proportion to the waste gas flow rate. Note that this flow rate corresponds to the saturated waste gas, Q_{sat}, not the inlet waste gas flow rate [24]. The cost equations use saturated waste gas volume since the actual flow rate can vary widely based on temperature, humidity and pressure.

The cost of a venturi per cubic foot of gas treated decreases as volume of gas increases due to economy of scale. However, if the scrubber becomes too large to be shipped as a package unit and must be field erected, the cost per cubic foot of gas treated increases. A venturi unit requires field erection when the cyclone separator diameter exceeds the shippable diameter, generally 12.5 feet inner diameter. This is equivalent to approximately 90,000 to 100,000 acfm of saturated gas [24].

Table 2.5 presents the equipment cost equations for each type of package venturi system. The equations are for venturi systems constructed of carbon steel and Alloy C-276. Table 2.6 gives multipliers for other materials including 304 and 316 stainless steel, fiberglass reinforced plastic (FRP), rubber coated steel, and epoxy coated steel. The equipment costs are "free on board" (FOB) which means that no taxes or freight are included. The cost of additional equipment (pumps, fans, etc.,) is generally 80% to100% of the package venturi system cost. Figures 15 through 17 present the cost equations as a function of flowrate.

Table 2.5: Venturi Equipment Costs, 2002 Dollars

Type of Venturi Unit	Saturated Air Flow Rate Range (cfm)	Equipment Cost Equation, $	
		Carbon Steel	Alloy C-276
low energy	1,000 to 90,000	$= 150\ Q_{sat.}^{0.56}$	$= 900\ Q_{sat}^{0.5}$
high energy	1,000 to 90,000	$= 170\ Q_{sat}^{0.56}$	$=1{,}300\ Q_{sat}^{0.5}$
packaged jet venturi [1]	100 to 10,000	$= 4.5\ Q_{sat} + 19{,}000$	N/A
variable throat	1,000 to 90,000	= 1.1 to 1.15 of the fixed throat cost	

[1] The jet venturi costs includes recycle pump, ID fan, piping and valves, basic instrumentation, and mounting skid.

Table 2.6: Equipment Cost Factors for Venturi Units Constructed of Other Materials

Material	Equipment Cost Factor
Stainless Steel, 304L	1.08 - 1.16
Stainless Steel, 316L	1.25 - 1.40
Fiberglass Reinforced Plastic (FRP)	1.6 [1]
Rubber Lining	1.6 [1]
Epoxy Coating	1.1 [1]

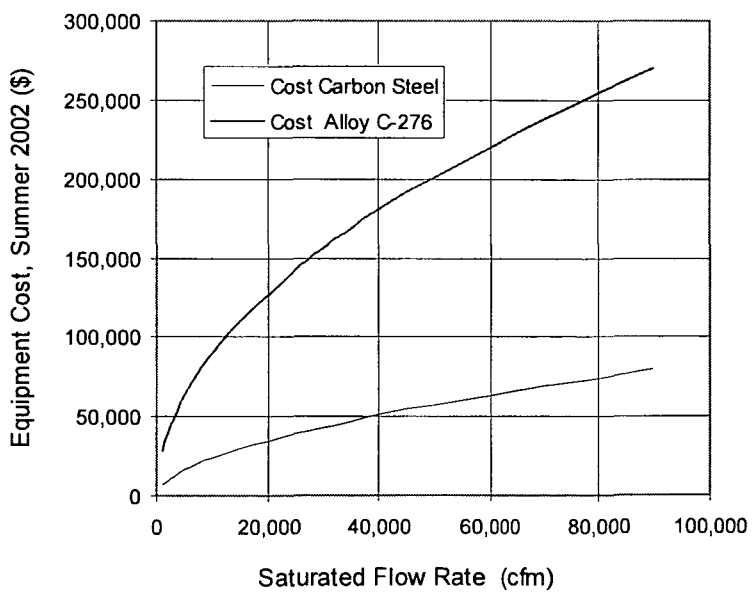


Figure 2.15: Equipment Costs for Low Energy Venturi

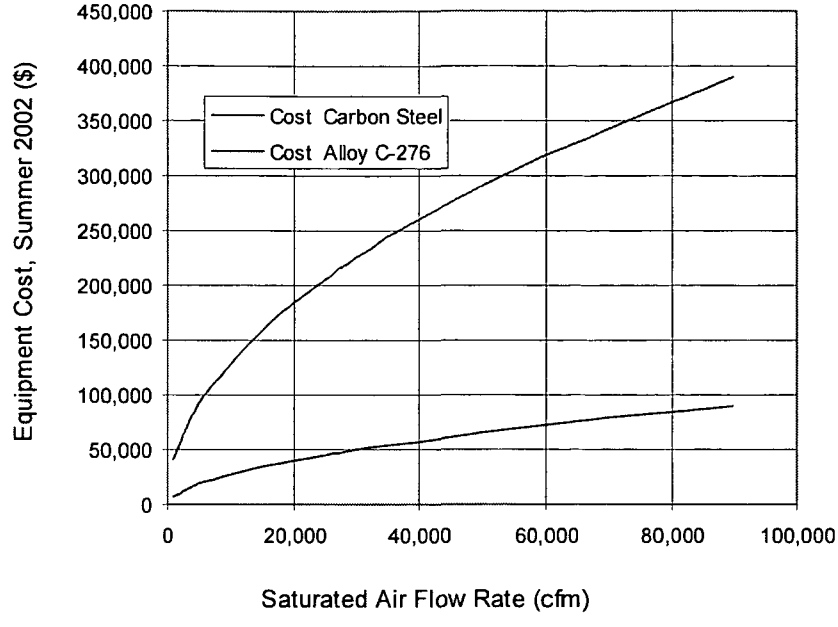


Figure 2.16: Equipment Costs for High Energy Venturi

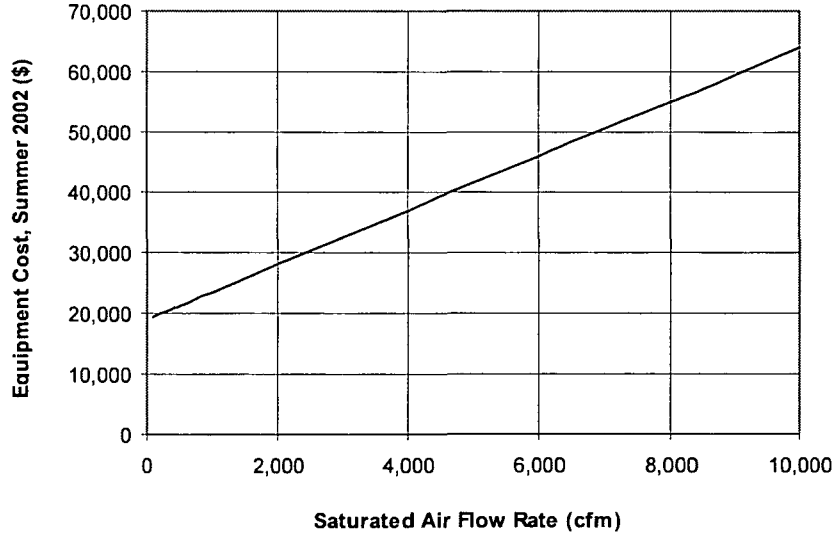


Figure 2.17: Equipment Costs for Jet Venturi

Higher pressure drops, require a greater thickness of the material of construction or stiffening of the material. The cost of the scrubber increases with the thickness of the material utilized. Venturi systems with pressure drops above 15 in w.c. generally require greater thickness or stiffening. For control of PM_{10}, the pressure drop is generally under 20 in w.c. while control of PM_{25} generally requires a pressure drop of 25 in w.c. or greater [24]. Table 2.7 presents typical pressure drops and materials of construction for various applications.

Table 2.7: Typical Venturi Scrubber Applications and Materials of Construction, [5, 24]

Application	Pressure Drop (in.wg)	Material of construction
Boilers		
Pulverized coal	15 - 40	316L stainless steel
Stoker coal	10 - 12	316L stainless steel
Bark	6 - 12	Carbon steel or stainless steelk
Combination	10 - 15	316L stainless steel
Recovery	30 - 40	Carbon, 316L stainless, or alloy steel
Incinerators		
Sewage sludge	18 - 20	316L stainless steel
Liquid waste	50 - 55	High nickel alloy
Solid waste		
Municipal	20 - 35	316L stainless steel or alloy steel
Pathological	20 - 35	316L stainless steel or alloy steel
Hospital	20 - 45	High nickel alloy
Kilns and calciners		
Lime	15 - 25	Carbon steel or stainless steel
Soda ash	20 - 40	Carbon steel or stainless steel
Potassium chloride	30	Carbon steel or stainless steel
Other		
Dryers	10 - 25	304 or 316L stainless steel
Crushers	6 - 20	Carbon steel
General spray dryer	15 - 30	Carbon steel or stainless steel

Note: Alloy steels may be required if corrosive halides are present

2.6.1.2 Direct and Indirect Installation Costs

Direct installation costs include materials and labor costs associated with installing the venturi unit. These costs include; auxiliary equipment (e.g., ductwork, compressor), foundations and supports, handling and erection, electrical, piping, insulation and painting. Indirect installation costs include engineering and supervision, construction and contractor fees, startup and testing, inventory capital, and any process and project contingency costs. Using the methodology presented in Section 1 of the *Manual*, Introduction and Cost Methodology, direct and indirect installation costs are estimated from a series of factors applied to the purchased equipment cost. The required factors are given in Table 2.8.

Table 2.8: Direct and Indirect Installation Costs for Venturi Scrubbers, [12]

Cost Item	Factor
Direct Costs	
Purchased equipment costs	
Venturi Packaged Unit	As estimated, A1
Auxiliary Costs	As estimated, A2
Equipment Costs	A = A1 + A2
Instrumentation	0.10 × A
Sales taxes	0.03 × A
Freight	0.05 × A
Purchased Equipment Cost, PEC	**B = 1.18 × A**
Direct installation costs	
Foundations & supports	0.06 × B
Handling & erection	0.40 × B
Electrical	0.01 × B
Piping	0.05 × B
Insulation for ductwork	0.03 × B
Painting	0.01 × B
Direct Installation Costs, DC	**0.56 × B**
Site preparation	As required, SP
Buildings	As required, Bldg.
Indirect Costs (installation)	
Engineering	0.10 × B
Construction and field expenses	0.10 × B
Contractor fees	0.10 × B
Start-up	0.01 × B
Performance test	0.01 × B
Model study	-
Contingencies	0.03 × B
Total Indirect Costs, IC	**0.35 × B**

As discussed in Section 2.3.7, auxiliary equipment required for operation of the venturi may include a pre-cooler, cyclone, hoods, stack, gas re-heater. Costs for hoods, ducts, stacks are presented in Section 2 of this Manual. Capital and O&M costs for cyclones are minimal and easily obtained from vendors. Costs for reheating are dependant on the temperature increase required for the waste gas.

Retrofit installations increase the costs of a venturi wet scrubber because of the common need to remove equipment to create space for the new scrubber. Additional ductwork to re-route the waste gas to the scrubber may also be required. The ducting path is often constrained by existing structures, additional supports are required, and the confined areas make erection more labor intensive and lengthy. Venturi scrubbers have a small footprint, therefore, retrofit costs are generally minimal. While retrofit costs are site-specific, a retrofit multiplier of 1.3 to 1.5 can be applied to the total capital investment to estimate these costs at the study level.

2.6.2 Annual Costs

Total annual cost (TAC) consist of direct costs, indirect costs, and recovery credits. No byproduct recovery credits are included because there are no salvageable byproducts generated from wet scrubbers. Therefore, TAC for venturi systems is given by:

$$TAC = DAC + IAC \tag{2.43}$$

where DAC = Direct Annual Costs, and
IAC = Indirect Annual Costs.

Each of these costs is discussed in the sections below. A more detailed discussion of annual costs can be found in Section 1, Chapter 2 of this *Manual*.

2.6.2.1 Direct Annual Costs

Direct annual costs (DAC) include variable and semivariable costs. Variable direct annual costs account for purchase of utilities, electrical power, and water. Semivariable direct annual costs include operating and supervisory labor and maintenance (labor and materials).

$$DAC = AC_{labor} + AC_{maint} + AC_{elect} + AC_{water} \tag{2.44}$$

where AC_{labor} = annual labor cost,
AC_{maint} = annual maintenance cost,
AC_{elect} = annual electricity cost, and
AC_{water} = annual water cost.

The quantity and cost of proprietary additives and waste disposal is application-specific, therefore, these costs are not estimated. Waste disposal costs depend on whether the waste liquid is de-watered and disposed as sludge or disposed in the municipal wastewater system. Waste disposal costs are greatly increased if the waste is classified as special or hazardous waste.

The labor costs are a function of the level of automation. Less labor is required for automatic controls but there are significantly higher capital costs for fully automated scrubber systems. Venturi scrubbers are assumed to require 2 to 8 hours of operating labor per shift [12]. More labor hours may be required for systems with highly variable flow rates, temperatures or pressures. Supervisory labor is assumed to be 15% of the operating labor and maintenance labor per shift, approximately 1 to 2 hours. The cost of materials required for maintenance is assumed to 100% of the maintenance labor cost. [12]

The amount of power and water utilized by the scrubber was estimated in Section 2.5.5. Using the estimated power consumption for the fan and pump, HP_{fan} and HP_{pump}, the annual cost of electricity is estimated from the following equation:

$$AC_{elect} = 0.7457\left(\frac{kW}{hp}\right)\left(HP_{fan} + HP_{pump}\right)\ t\ Cost_E \qquad (2.45)$$

where t = scrubber operating time per year, hours, and
$Cost_E$ = cost of electricity in dollars per kW ($/kw).

The cost of water is estimated from the total volume of water, $V_{T(H2O)}$ calculated in Equation 2.39:

$$AC_{H20} = V_{T(H2O)}\ Cost_{H2O} \qquad (2.46)$$

where $Cost_{H2O}$ = cost of water in dollars per gallon ($/gal).

2.6.2.2 Indirect Annual Cost

In general, IAC (fixed cost) includes property taxes, insurance, administrative charges, overhead, and the capital recovery cost. Section 1 of the *Manual* discusses these costs in detail.

Administrative costs, property tax, and insurance are assumed to be percentages of the TCI [12]. Overhead is assumed to be equal to 60% of the sum of operating, supervisory, and maintenance labor, and maintenance materials [12]. Capital recovery cost is based on the anticipated equipment lifetime and the annual interest rate employed. Table 2.9 gives suggested factors f these items.

An economic lifetime of 15 years is assumed for the wet scrubber system. For a 15-year life and an interest rate of 7 percent, the capital recovery factor, CRF, is equal to 0.1098. The system capital recovery cost is then estimated by:

$$CRF = 0.1098\ TCI \tag{2.47}$$

Table 2.9: Annual Cost Factors for Scrubbers [12]

Cost Item	Factor
Direct Annual Costs, DC	
Operating labor	
Operator	2 to 8 hours per shift
Supervisor	15% of operator
Maintenance	
Labor	1 to 2 hours per shift
Material	100% of maintenance labor
Utilities,	(consumption rate) x (hours/yr) x (unit cost)
Fan	
Pump	
Water	
Operating materials	Application specific
Wastewater disposal	Application specific
Indirect Annual Costs, IC	
Administrative charges	2% of Total Capital Investment
Property tax	1% of Total Capital Investment
Insurance	1% of Total Capital Investment
Overhead	60% of total labor and material costs
Capital recovery	0.1098 x Total Capital Investment
Total Annual Cost	**DC + IC**

2.7 Example Problem

Sludge incinerators frequently employ wet scrubbers for particulate emissions control. The furnace is generally a multiple-hearth or fluid-bed type. Gaseous emissions from the incinerator include carbon monoxide (CO), nitrogen oxides (NO_x), sulfur oxides (SO_x), volatile organic compounds (VOC), and hydrogen chloride (HCl). Particulate emissions include ash and heavy metals. Metal oxides will either be adsorbed onto the particulate matter or be free within the gas stream. The free oxides are generally less than 0.5 microns mean particle size and cannot be removed using conventional scrubbers. The moisture content from a sludge incinerator is very high, therefore wet scrubbing is generally employed.

This example is for a multi-hearth sludge incinerator. The inlet conditions to the venturi scrubber are:

volume flow rate	75,000 acfm
inlet temperature	450 °F
moisture content	20%
particulate loading	3 grains/scf
specific density of particulate	1.8

The design parameters for the scrubber are:

required collection efficiency	90% for < 1.0 micron PM

The particle size distribution for a log-normally distributed incinerator source is given in Table 2.10 and Figure 2.18 presents the plot of the distribution.

Table 2.10: Particle Size Distribution for an Sludge Incinerator Source

Particle Size Range (microns)	Median Particle Diameter (microns)	Mass Fraction	Cummulative Mass Fraction
0 - 1	0.50	0.005	0.005
1 - 2.5	1.75	0.195	0.200
2.5 - 4.5	3.50	0.400	0.600
4.5 - 7	5.75	0.300	0.900
7 - 12	9.50	0.080	0.980
> 12	20.00	0.020	1.000

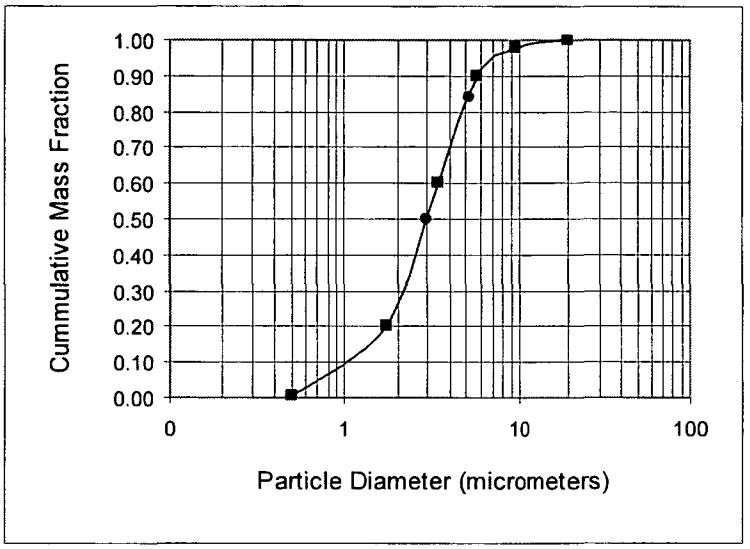


Figure 2.18: Particle Size Distribution for Sludge Incineratior

From Figure 2.18 the mass median particle diameter, d_{50}, and the 84th percentile mass particle diameter, d_{84}, can be read as:

$$d_{50} = 3$$

$$d_{84} = 5.1$$

Therefore, the standard deviation of the distribution as given by Equation 2.1 is:

$$\sigma = \frac{5.1}{3} = 1.7$$

Now we will calculate the properties of the waste gas at the outlet of the venturi scrubber assuming adiabatic saturation of the waste gas. Using the ideal gas relationship we calculate the volume flow rate of gas at standard conditions. We will use the following information:

molecular meight of water vapor = 18
molecular meight of air = 29
Volume of a lb-mole of air = 385 ft^3

The waste gas flow rate at standard temperature is:

$$75{,}000 \text{ acfm} \frac{(70°\text{F} + 460)}{(350°\text{F} + 460)} = 49{,}074 \text{ scfm}$$

The mass flow rate of dry air and mass of water vapor can be calculated from Equation 2.12 and the molecular weights.

$$\dot{m}_{wv} = 49{,}074 \text{ scfm} \times 25\% \times \frac{18}{385} = 574 \text{ lb/min of water vapor}$$

$$\dot{m}_{a} = 49{,}074 \text{ scfm} \times 75\% \times \frac{29}{385} = 2{,}772 \text{ lb/min of dry air}$$

The humidity ratio at standard temperature can now be calculated:

$$\omega = \frac{574 \text{ lb/min of dry air}}{2{,}772 \text{ lb/min of water vapor}} = 0.21$$

From the psychrometric chart given in Figure 2.9, we find the point on the chart for a dry bulb temperature of 350°F and a humidity ratio of 0.21. The line of constant enthalpy (purple) is followed to the left until we reach the 100% relative humidity line (red). This point is at a dry bulb temperature 160°F, a humidity ratio of 0.26, and a humid volume of 22ft³/lb.

The waste gas properties at the outlet of the venturi scrubber can now be calculated. The outlet waste gas flow rate is given by Equation 2.13:

$$Q_{m(out)} = 22 \frac{\text{ft}^3}{\text{lb}} \times 2{,}772 \frac{\text{lb}}{\text{min}} = 60{,}984 \text{ scfm}$$

The outlet humidity ratio gives the outlet mass flow rate of water vapor as given in Equation 2.14.

$$\dot{m}_{wv(out)} = 2{,}772 \frac{\text{lb}}{\text{min}} \times .26 = 721 \frac{\text{lb}}{\text{min}}$$

The water evaporated from the scrubbing fluid due to adiabatic saturation of the waste gas stream is calculated from Equation 2.15 as:

$$\dot{m}_{wv(evap)} = 721\frac{\text{lb}}{\text{min}} - 574\frac{\text{lb}}{\text{min}} = 147\frac{\text{lb}}{\text{min}}$$

The volume of makeup water for the recirculation system is then given by Equation 2.16 as:

$$Q_{wv(evap)} = \frac{147\ \text{lb/min}}{62.4\ \text{lb/ft}^3} = 2.4\ \text{cfm} = 18\ \text{gpm}$$

The next step is to size the scrubber. The first parameter to estimate is the pressure drop across the scrubber. We will employ the Calvert Cut Diameter approach. We use Figure 2.13 since the scrubber a venturi and B = 2.0.

First we need to determine the overall collection efficiency. Assuming the following collection efficiency requirements for each size range, the fraction collection efficiency is calculated from the mass fraction multiplied by the required collection efficiency (Equation 2.4):

Particle Size Range (microns)	Mass Fraction	Required Collection Efficiency	Fractional Collection Efficiency
0 - 1	0.005	0.900	0.0045
1 - 2.5	0.195	0.950	0.185
2.5 - 4.5	0.400	0.980	0.392
4.5 - 7	0.300	0.990	0.297
7 - 12	0.080	1.000	0.080
> 12	0.020	1.000	0.020
Overall Collection Efficiency			**0.979**

Note that particles greater than 5 microns are assumed to be captured at 100% efficiency for a venturi scrubber.

Reading the graph in Figure 2.13 for the following point:

$$\sigma = 1.7$$
$$Pt = 1\text{-}\eta_d = 0.02$$

We obtain a value for the cut ratio and then calculate cut diameter:

$$\frac{d_{cut}}{d_{50}} \approx 0.2$$

$$d_{cut} = 0.2 \times 3 = 0.6$$

From Figure 2.14, a venturi scrubber with a pressure drop of approximately 15 in of water column with a scrubber power of 4.5 hp per 1,000 ft^3/min is sufficient. Using Figure 2.6, the approximate venturi scrubber pressure drop is estimated at 12 in. of water column.

Now we can use this estimated pressure drop to size the rest of the venturi. Using Figure 3.10 and assuming a L/G of 10 gal per 1,000 acfm we obtain an approximate throat velocity of 320 ft/s. Using equation 2.17, we can estimate the velocity at the throat. We need the density of air at saturation which is the inverse of the humid volume:

$$HP_{fan} = \frac{15 \text{ in water} \times 75{,}000 \text{acfm}}{6356 \times 0.6} = 294 \text{ hp}$$

The brake horsepower of the pump is calculated from Equation 2.36. First we must calculate the specific gravity of the scrubbing liquid. Assuming the scrubbing liquid is water and the specific gravity of the PM is 1.8 we can estimate the specific gravity of the slurry as:

$$\gamma_{slurry} = \frac{\%solids + \%water}{\frac{\%solids}{\gamma_{PM}} + \frac{\%water}{\gamma_{water}}} = \frac{25+75}{\frac{25}{1.8} + \frac{75}{1.0}} = 1.125$$

We have a throat length of 13.5 ft and a diverging section of 18 ft, resulting in a total length of 31.5 ft. Therefore we can assume the pump must be sized for 40 ft of water column. This gives the power for the pump as:

$$HP_{pump} = \frac{40 \text{ ft} \times 10 \frac{\text{gal}}{1000 \text{ acfm}} \times \frac{75{,}000 \text{ acfm}}{1000} \times 1.125}{3952.6 \times 0.5} = 17 \text{ hp}$$

We now move on to estimating capital costs for the venturi scrubber. The venturi must be sized for the saturated flow rate of 60,984 acfm. The pressure drop required is 15 in of water, therefore, a low energy venturi is sufficient. The material of construction must be Stainless Steel, 304 L since the waste gas stream contains corrosives. Applying the materials multiplier to the equipment cost equation gives:

$$Venturi\ EC = 1.10 \left[150 \times 61{,}000^{0.56}\right] = \$78{,}950$$

This value must be increased by 80% to 100% to account for the cost of pumps, an upgraded ID fan, piping, valves, instrumentation and control. A moderate level of automation is assumed to be required, therefore the basic equipment cost is increased by 90% to account for the additional equipment:

$$Total\ EC = \$78{,}950\,(1.0 + 0.9) = 150{,}000$$

Adding additional equipment, taxes, and frieght to the basic equipment costs results in the PEC given by:

$$PEC = \$150{,}000\,(1.0 + 0.03 + 0.05) = \$162{,}000$$

Direct and indirect installation costs are given in Table 2.8. Using these factors, we can calculate TCI as:

$$TCI = \$162{,}000\,(1.0 + 0.56 + 0.35) = \$309{,}420$$

Assuming the system is a retrofit of low difficulty, we must increase the TCI by a factor of 1.3

$$Retrofit\ TCI = 1.3\,(\$309{,}420) = \$402{,}250$$

Total annual costs include direct annual costs and indirect annual costs. First we calculate the cost of labor and materials.

$$Operating\ Labor = \frac{3hr}{shift} \times \frac{2shifts}{day} \times \frac{300days}{yr} \times \frac{\$20.00}{hr} = \$39{,}600\ per\ year$$

$$Supervisory\ Labor = 0.15 \times 39{,}600 = \$5{,}940\ per\ year$$

$$Maintenance\ Labor = \frac{1hr}{shift} \times \frac{2shifts}{day} \times \frac{300days}{year} \times \frac{\$20.00}{hr} = \$13{,}200\ per\ year$$

$$AC_{labor} = \$39{,}000 + \$5{,}940 + \$13{,}200 = \$58{,}740\ per\ year$$

$$AC_{materials} = 100\%\ (\$132{,}000) = \$13{,}200\ per\ year$$

Next we calculate the cost of utilities.

$$AC_{elect} = 0.7457\frac{kW}{hp}\times(294hp+17hp)\times\frac{330days}{yr}\times\frac{16hr}{day}\times\frac{\$0.07}{kWh} = \$85{,}720\ per\ year$$

$$AC_{H2O} = 28gpm\times\frac{330days}{yr}\times\frac{16hr}{day}\times\frac{60min}{hr}\frac{\$0.2}{1{,}000gal} = \$1{,}770\ per\ year$$

So the total direct annual costs is the sum of these costs:

$$DAC = \$58{,}740 + \$13{,}200 + \$85{,}720 + 1{,}770 = \$159{,}430\ per\ year$$

Indirect annual costs include overhead, administrative charges, property taxes, insurance and the capital recovery factor. The following table details these costs.

Indirect Annual Cost	Factor	Cost per year	
Overhead Costs 60%	(AC_{labor}+ $AC_{materials}$)	0.60 ($58,740+$13,200) =	$43,160
Administrative Costs	2% TCI	0.02 ($402,250) =	$8,050
Property Taxes	1% TCI	0.01 ($402,250) =	$4,020
Insurance	1% TCI	0.01 ($402,250) =	$4,020
Capital Recovery	0.1098 TCI	0.1098 ($402,250) =	$44,170
Total IAC			**$103,420**

The total annual cost is the sum of direct and indirect annual costs given by:

$$TAC = \$159{,}430 + \$103{,}420 = \$262{,}850\ per\ year$$

2.8 Acknowledgments

We gratefully acknowledge the following companies for contributing data to this section:

- Bionomic Industries, Inc (Mahwah, NJ)
- Croll-Reynolds (Westfield, NJ)
- Misonix, Inc. (Farmingdale, NY)

References

[1] Schifftner, K.C. and H.E. Hesketh, Wet Scrubbers (2nd Edition), Technomic Publishing, Lancaster, PA, 1996.

[2] Davis, W.T. Ed., Air Pollution Engineering Manual (2nd Edition), Air and Waste Management, John Wiley & Sons, Inc., New York, 2000.

[3] Environmental Protection Agency, U.S. (EPA), "Stationary Source Control Techniques Document for Fine Particulate", EPA Document No. EPA-452/R-97-001, Office of Air Quality Planning and Standards, Research Triangle Park, NC, September 30, 1997.

[4] Wark, K., C.F. Warner, and W.T. Davis, Air Pollution: Its Origin and Control (3rd Edition), Addison-Wesley, Reading, MA, 1998.

[5] Cooper, C.D. and F.C. Alley, Air Pollution Control: A Design Approach, Waveland Press, Inc., Prospect Heights, IL, 1994.

[6] Environmental Protection Agency, U.S. (EPA), Air Pollution Technology Fact Sheet, Office of Air Quality Planning and Standards, Research Triangle Park, NC, 1990.

[7] Avallone, E. and T. Baumeister, Mark's Standard Handbook for Mechanical Engineers, McGraw Hill, New York, 1996.

[8] The Scrubber Manual (Revised), The McIlvaine Company, Northbrook, IL, 1995.

[9] Advanced Air Technologies, Inc., technical information on scrubbers, www.aatinc.com/scrubbing_zone.htm, 2002.

[10] William L. Heuman, Industrial Air Pollution Control Systems, McGraw-Hill, Washington D.C., 1997.

[11] "Air Pollution Control Systems for Boilers and Incinerators", U.S. Army Technical Manual, TM 5-815-1/AFR 19-6, May 9, 1998.

[12] Vatavuk, WIlliam, Estimating Costs of Air Pollution Control, Lewis Publishers, Chelsea Michigan, 1990.

[13] Cheremisinoff, Paul. N., Ed., Air Pollution Control and Design for Industry, Marcel Dekker, Inc., New York, 1993.

[14] Theodore, Louis and Anthony Buonicore, Ed., Air Pollution Control Equipment: Selection, Design, Operation, and Maintenance, Prentice-Hall, Inc. Englewood Cliffs, New Jersey, 1982.

[15] Calvert, S., J. Goldsmith, G. Leith, and D. Metha, "Scrubber Handbook", NTIS, Springfield, VA, 1972

[16] Hesketh, H.E., "Fine Particulate Collection Efficiency Related to Pressure Drop, Scrubbant and Particle Properties, and Contact Mechanisims, " *Journal of Air Pollution Control Association*, vol. 24, no. 10, 1974.

[17] Lapple, C.E. and H.J. Kamack, "Performance of Wet Dust Scrubbers", Chemical Engineering Progress, vol. 51, March 1955.

[18] Semrau, K.T., "Dust Scrubber Design - A Critique on the State of the Art,", " *Journal of Air Pollution Control Association*, vol.13, December 1963.

[19] Semrau, K.T., "Practical Process Design of Particulate Scrubbers", in *Industrial Air Pollution Engineering*, V. Casaseno, ed., McGraw Hill, New York, 1980.

[20] "Scrubber Selection Guide", Calvert Environmental Equipment, San Diego, 1988.

[21] Schnelle, Karl B. Jr. and Charles A. Brown, *Air Pollution Control Technology Handbook*, CRC Press, Washington D.C., 2002.

[22] Schifftner, Kenneth C. *Air pollution Control Equipment Selection Guide*, Lewis Publishers, Washington D.C., 2002.

[23] Calvert, S., How to choose a particulate scrubber, *Chem. Eng.*, 54-68, August 29, 1977.

[24] Personal correspondence from Ken Schiftner of Bionomic Industries, Inc., to Paula Hemmer, E.H. Pechan and Associates, Inc., June, 2002.

TECHNICAL REPORT DATA

(Please read Instructions on reverse before completing)

1. REPORT NO.	2.	3. RECIPIENT'S ACCESSION NO.
452/B-02-001		

4. TITLE AND SUBTITLE	5. REPORT DATE
The EPA Air Pollution Control Cost Manual	January, 2002
	6. PERFORMING ORGANIZATION CODE

7. AUTHOR(S)	8. PERFORMING ORGANIZATION REPORT NO.
Daniel Charles Mussatti	

9. PERFORMING ORGANIZATION NAME AND ADDRESS	10. PROGRAM ELEMENT NO.
U.S. Environmental Protection Agency Office of Air Quality Planning and Standards Air Quality Standards and Strategies Division Innovative Strategies and Economics Group Research Triangle Park, NC 27711	11. CONTRACT/GRANT NO.

12. SPONSORING AGENCY NAME AND ADDRESS	13. TYPE OF REPORT AND PERIOD COVERED
Director Office of Air Quality Planning and Standards Office of Air and Radiation U.S. Environmental Protection Agency Research Triangle Park, NC 27711	Final
	14. SPONSORING AGENCY CODE EPA/200/04

15. SUPPLEMENTARY NOTES

Updates and revises EPA 453/b-96-001, OAQPS Control Cost Manual, fifth edition (in English only)

16. ABSTRACT

In Spanish, this document provides a detailed methodology for the proper sizing and costing of numerous air pollution control devices for planning and permitting purposes. Includes costing for volatile organic compounds (VOCs); particulate matter (PM); oxides of nitrogen (NOx); SO2, SO3, and other acid gasses; and hazardous air pollutants (HAPs).

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Economics Cost Engineering cost Sizing Estimation Design	Air Pollution control Incinerators Absorbers Adsorbers Filters Condensers Electrostatic Precipitators Scrubbers	

18. DISTRIBUTION STATEMENT	19. SECURITY CLASS *(Report)*	21. NO. OF PAGES
Release Unlimited	Unclassified	1,400
	20. SECURITY CLASS *(Page)* Unclassified	22. PRICE

EPA/452/B-02-001

Chapter 3

Electrostatic Precipitators

James H. Turner
Phil A. Lawless
Toshiaki Yamamoto
David W. Coy
Research Triangle Institute
Research Triangle Park, NC 27709

John D. McKenna, John C. Mycock & Arthur B. Nunn
ETS, Inc.
Roanoke, VA 24018-4394

Gary P. Greiner
John D. McKenna
ETS, Inc.
Roanoke, VA 24018-4394

William M. Vatavuk
Innovative Strategies and Economics Group, OAQPS
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

September 1999

Contents

3.1 Process Description 3-4
3.1.1 Introduction 3-4
3.1.2 Types of ESPs 3-4
3.1.2.1 Plate-Wire Precipitators 3-4
3.1.2.2 Flat Plate Precipitators 3-8
3.1.2.3 Tubular Precipitators 3-9
3.1.2.4 Wet Precipitators 3-9
3.1.2.5 Two-Stage Precipitators 3-9
3.1.3 Auxiliary Equipment 3-10
3.1.4 Electrostatic Precipitation Theory 3-12
3.1.4.1 Electrical Operating Point 3-12
3.1.4.2 Particle Charging 3-15
3.1.4.3 Particle Collection 3-17
3.1.4.4 Sneakage and Rapping Reentrainment 3-19

3.2 ESP Design Procedure 3-21
3.2.1 Specific Collecting Area 3-21
3.2.1.1 SCA Procedure with Known Migration Velocity 3-22
3.2.1.2 Full SCA Procedure 3-25
3.2.1.3 Specific Collecting Area for Tubular Precipitators 3-31
3.2.2 Flow Velocity 3-31
3.2.3 Pressure Drop Calculations 3-33
3.2.4 Particle Characteristics 3-34
3.2.5 Gas Characteristics 3-35
3.2.6 Cleaning 3-36
3.2.7 Construction Features 3-37

3.3 Estimating Total Capital Investment 3-37
3.3.1 Equipment Cost 3-38
3.3.1.1 ESP Costs 3-38
3.3.1.2 Retrofit Cost Factor 3-41
3.3.1.3 Auxiliary Equipment 3-42
3.3.1.4 Costs for Two-Stage Precipitators 3-42
3.3.2 Total Purchased Cost 3-44
3.3.3 Total Capital Investment (TCI) 3-44

3.4 Estimating Total Annual Costs 3-45
3.4.1 Direct Annual Costs 3-45
3.4.1.1 Operating and Supervisory Labor 3-45
3.4.1.2 Operating Materials 3-47
3.4.1.3 Maintenance 3-47
3.4.1.4 Electricity 3-48
3.4.1.5 Fuel 3-49
3.4.1.6 Water 3-49
3.4.1.7 Compressed Air 3-49
3.4.1.8 Dust Disposal 3-49

3.4.1.9 Wastewater Treatment 3-50
3.4.1.10 Conditioning Costs 3-50
3.4.3 Recovery Credits 3-51
3.4.4 Total Annual Cost 3-51
3.4.5 Example Problem 3-51
3.4.5.1 Design SCA 3-52
3.4.5.2 ESP Cost 3-55
3.4.5.3 Costs of Auxiliaries 3-56
3.4.5.5 Annual Costs-Pressure Drop 3-56
3.4.5.6 Total Annual Cost 3-58

3.5 Acknowledgments 3-59

References 3-60

3.1 Process Description

3.1.1 Introduction

An electrostatic precipitator (ESP) is a particle control device that uses electrical forces to move the particles out of the flowing gas stream and onto collector plates. The particles are given an electrical charge by forcing them to pass through a corona, a region in which gaseous ions flow. The electrical field that forces the charged particles to the walls comes from electrodes maintained at high voltage in the center of the flow lane. Figure 3.1 is an example of electrostatic precipitator components.

Once the particles are collected on the plates, they must be removed from the plates without reentraining them into the gas stream. This is usually accomplished by knocking them loose from the plates, allowing the collected layer of particles to slide down into a hopper from which they are evacuated. Some precipitators remove the particles by intermittent or continuous washing with water.

3.1.2 Types of ESPs

ESPs are configured in several ways. Some of these configurations have been developed for special control action, and others have evolved for economic reasons. The types that will be described here are (1) the plate-wire precipitator, the most common variety; (2) the flat plate precipitator, (3) the tubular precipitator; (4) the wet precipitator, which may have any of the previous mechanical configurations; and (5) the two-stage precipitator. See Figure 6.14 for examples of typical flate-plate and plate-wire ESP configurations.

3.1.2.1 Plate-Wire Precipitators

Plate-wire ESPs are used in a wide variety of industrial applications, including coal-fired boilers, cement kilns, solid waste incinerators, paper mill recovery boilers, petroleum refining catalytic cracking units, sinter plants, basic oxygen furnaces, open hearth furnaces, electric arc furnaces, coke oven batteries, and glass furnaces.

In a plate-wire ESP, gas flows between parallel plates of sheet metal and high-voltage electrodes. These electrodes are long wires weighted and hanging between the plates or are supported there by mast-like structures (rigid frames). Within each flow path, gas flow must pass each wire in sequence as flows through the unit.

The plate-wire ESP allows many flow lanes to operate in parallel, and each lane can be quite tall. As a result, this type of precipitator is well suited for handling large volumes of gas. The need for rapping the plates to dislodge the collected material has caused the plate to be divided

into sections, often three or four in series with one another, which can be rapped independently. The power supplies are often sectionalized in the same way to obtain higher operating voltages, and further electrical sectionalization may be used for increased reliability. Dust also deposits on the discharge electrode wires and must be periodically removed similarly to the collector plate.

The power supplies for the ESP convert the industrial ac voltage (220 to 480 V) to pulsating dc voltage in the range of 20,000 to 100,000 V as needed. The supply consists of a step-up transformer, high-voltage rectifiers, and sometimes filter capacitors. The unit may supply either half-wave or full-wave rectified dc voltage. There are auxiliary components and controls to allow the voltage to be adjusted to the highest level possible without excessive sparking and to protect the supply and electrodes in the event a heavy arc or short-circuit occurs.

The voltage applied to the electrodes causes the air between the electrodes to break down electrically, an action known as a "corona." The electrodes usually are given a negative polarity because a negative corona supports a higher voltage than a positive corona before sparking occurs. The ions generated in the corona follow electric field lines from the wires to the collecting plates. Therefore, each wire establishes a charging zone through which the particles must pass.

Particles passing through the charging zone intercept some of the ions, which become attached. Small aerosol particles ($<1\ \mu$m diameter) can absorb tens of ions before their total charge becomes large enough to repel further ions, and large particles ($>10\ \mu$m diameter) can absorb tens of thousands. The electrical forces are therefore much stronger on the large particles.

As the particles pass each successive wire, they are driven closer and closer to the collecting walls. The turbulence in the gas, however, tends to keep them uniformly mixed with the gas. The collection process is therefore a competition between the electrical and dispersive forces. Eventually, the particles approach close enough to the walls so that the turbulence drops to low levels and the particles are collected.

If the collected particles could be dislodged into the hopper without losses, the ESP would be extremely efficient. The rapping that dislodges the accumulated layer also projects some of the particles (typically 12 percent for coal fly ash) back into the gas stream. These reentrained particles are then processed again by later sections, but the particles reentrained in the last section of the ESP have no chance to be recaptured and so escape the unit.

Practical considerations of passing the high voltage into the space between the lanes and allowing for some clearance above the hoppers to support and align electrodes leave room for part of the gas to flow around the charging zones. This is called "sneakage" and amounts to 5 to 10 percent of the total flow. Antisneakage baffles usually are placed to force the sneakage flow to mix with the main gas stream for collection in later sections. But, again, the sneakage flow around the last section has no opportunity to be collected.

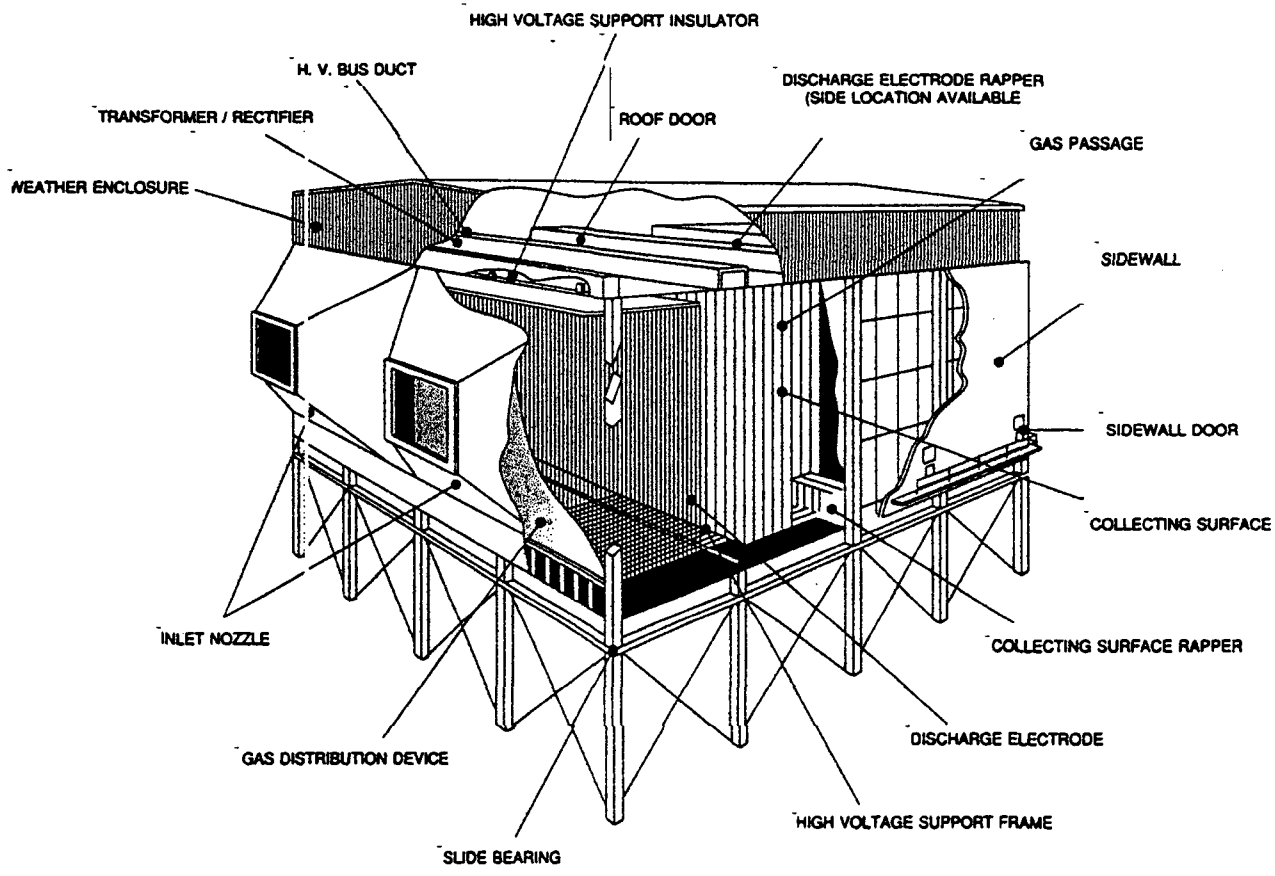


Figure 3.1: Electrostatic Precipitator Components
(Courtesy of the Institute for Clean Air Companies)

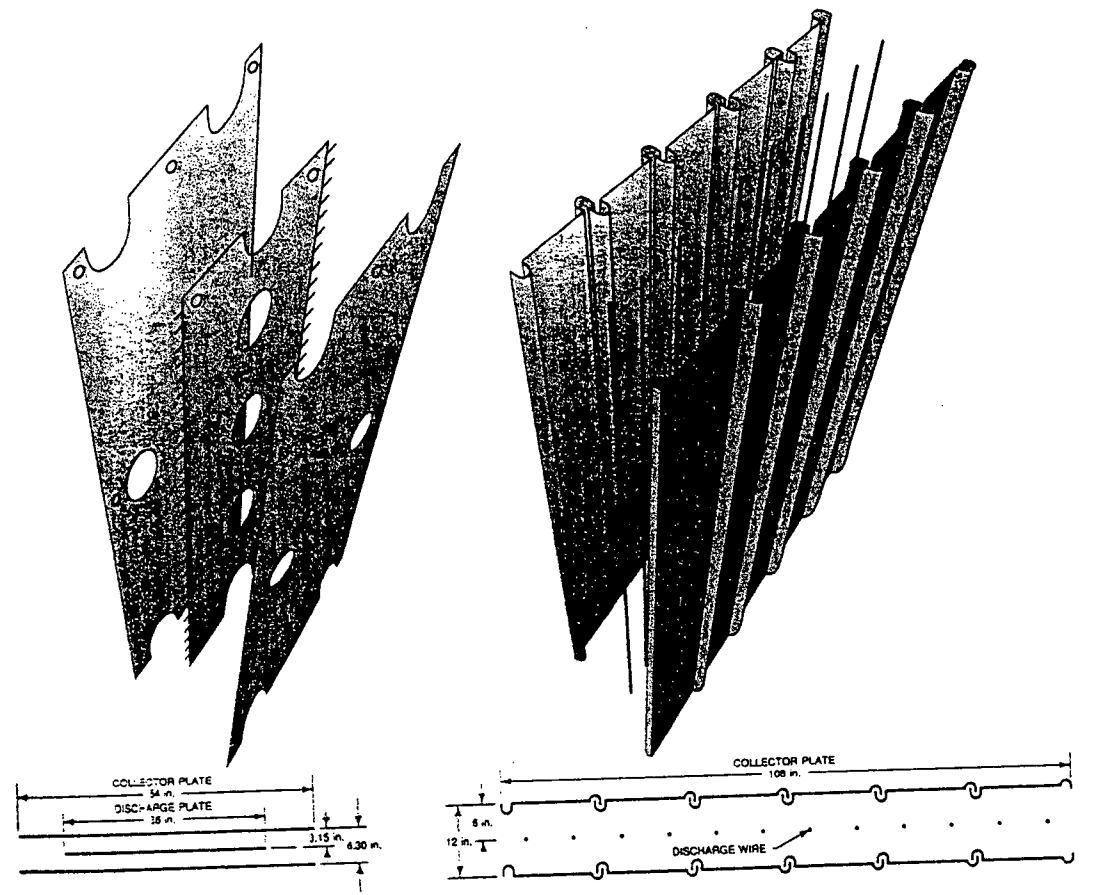


Figure 3.2: Flate-plate and Plate-wire ESP Configurations
(Courtesy of United McGill Corporation)

These losses play a significant role in the overall performance of an ESP. Another major factor is the resistivity of the collected material. Because the particles form a continuous layer on the ESP plates, all the ion current must pass through the layer to reach the ground-plates. This current creates an electric field in the layer, and it can become large enough to cause local electrical breakdown. When this occurs, new ions of the wrong polarity are injected into the wire-plate gap where they reduce the charge on the particles and may cause sparking. This breakdown condition is called "back corona."

Back corona is prevalent when the resistivity of the layer is high, usually above 2×10^{11} ohm-cm. For lower resistivities, the operation of the ESP is not impaired by back coronas, but resistivities much higher than 2×10^{11} ohm-cm considerably reduce the collection ability of the unit because the severe back corona causes difficulties in charging the particles. At resistivities below 10^8 ohm-cm, the particles are held on the plates so loosely that rapping and nonrapping reentrainment become much more severe. Care must be taken in measuring or estimating resistivity because it is strongly affected by variables such as temperature, moisture, gas composition, particle composition, and surface characteristics.

3.1.2.2 Flat Plate Precipitators

A significant number of smaller precipitators (100,000 to 200,000 acfm) use flat plates instead of wires for the high-voltage electrodes. The flat plates (United McGill Corporation patents) increase the average electric field that can be used to collect the particles, and they provide an increased surface area for the collection of particles. Corona cannot be generated on flat plates by themselves, so corona-generating electrodes are placed ahead of and sometimes behind the flat plate collecting zones. These electrodes may be sharp-pointed needles attached to the edges of the plates or independent corona wires. Unlike place-wire or tubular ESPs, this design operates equally well with either negative or positive polarity. The manufacturer has chosen to use positive polarity to reduce ozone generation.

A flat plate ESP operates with little or no corona current flowing through the collected dust, except directly under the corona needles or wires. This has two consequences. The first is that the unit is somewhat less susceptible to back corona than conventional units are because no back corona is generated in the collected dust, and particles charged with both polarities of ions have large collection surfaces available. The second consequence is that the lack of current in the collected layer causes an electrical force that tends to remove the layer from the collecting surface; this can lead to high rapping losses.

Flat plate ESPs seem to have wide application for high-resistivity particles with small (1 to 2 μm) mass median diameters (*MMDs*). These applications especially emphasize the strengths of the design because the electrical dislodging forces are weaker for small particles than for large ones. Fly ash has been successfully collected with this type of ESP, but low-flow velocity appears to be critical for avoiding high rapping losses.

3.1.2.3 Tubular Precipitators

The original ESPs were tubular like the smokestacks they were placed on, with the high-voltage electrode running along the axis of the tube. Tubular precipitators have typical applications in sulfuric add plants, coke oven by-product gas cleaning (tar removal), and, recently, iron and steel sinter plants. Such tubular units are still used for some applications, with many tubes operating in parallel to handle increased gas flows. The tubes may be formed as a circular, square, or hexagonal honeycomb with gas flowing upwards or downwards. The length of the tubes can be selected to fit conditions. A tubular ESP can be tightly sealed to prevent leaks of material, especially valuable or hazardous material.

A tubular ESP is essentially a one-stage unit and is unique in having all the gas pass through the electrode region. The high-voltage electrode operates at one voltage for the entire length of the tube, and the current varies along the length as the particles are removed from the system. No sneakage paths are around the collecting region, but corona nonuniformities may allow some particles to avoid charging for a considerable fraction of the tube length.

Tubular ESPs comprise only a small portion of the ESP population and are most commonly applied where the particulate is either wet or sticky. These ESPs, usually cleaned with water, have reentrainment losses of a lower magnitude than do the dry particulate precipitators.

3.1.2.4 Wet Precipitators

Any of the precipitator configurations discussed above may be operated with wet walls instead of dry. The water flow may be applied intermittently or continuously to wash the collected particles into a sump for disposal. The advantage of the wet wall precipitator is that it has no problems with rapping reentrainment or with back coronas. The disadvantage is the increased complexity of the wash and the fact that the collected slurry must be handled more carefully than a dry product, adding to the expense of disposal.

3.1.2.5 Two-Stage Precipitators

The previously described precipitators are all parallel in nature, *i.e.*, the discharge and collecting electrodes are side by side. The two-stage precipitator invented by Penney is a series device with the discharge electrode, or ionizer, preceding the collector electrodes. For indoor applications, the unit is operated with positive polarity to limit ozone generation.

Advantages of this configuration include more time for particle charging, less propensity for back corona, and economical construction for small sizes. This type of precipitator is generally used for gas flow volumes of 50,000 acfm and less and is applied to submicrometer sources emitting oil mists, smokes, fumes, or other sticky particulates because there is little electrical force

to hold the collected particulates on the plates. Modules consisting of a mechanical prefilter, ionizer, collecting-plate cell, after-filter, and power pack may be placed in parallel or series-parallel arrangements. Preconditioning of gases is normally part of the system. Cleaning may be by water wash of modules removed from the system up to automatic, in-place detergent spraying of the collector followed by air-blow drying.

Two-stage precipitators are considered to be separate and distinct types of devices compared to large, high-gas-volume, single-stage ESPs. The smaller devices are usually sold as pre-engineered, package systems.

3.1.3 Auxiliary Equipment

Typical auxiliary equipment associated with an ESP system is shown schematically in Figure 3.3. Along with the ESP itself, a control system usually includes the following auxiliary equipment: a capture device (i.e., hood or direct exhaust connection); ductwork; dust removal equipment (screw conveyor, etc.); fans, motors, and starters; and a stack. In addition, spray coolers and mechanical collectors may be needed to precondition the gas before it reaches the ESP. Capture devices are usually hoods that exhaust pollutants into the ductwork or are direct exhaust couplings attached to a combustor or process equipment. These devices are usually refractory lined, water cooled, or simply fabricated from carbon steel, depending on the gas-stream temperatures. Refractory or water-cooled capture devices are used where the wall temperatures exceed 800°F; carbon steel is used for lower temperatures. The ducting, like the capture device, should be water cooled, refractory, or stainless steel for hot processes and carbon steel for gas temperatures below approximately 1150°F (duct wall temperatures <800°F). The ducts should be sized for a gas velocity of approximately 4,000 ft/min for the average case to prevent particle deposition in the ducts. Large or dense particles might require higher velocities, but rarely would lower velocities be used. Spray chambers may be required for processes where the addition of moisture, or decreased temperature or gas volume, will improve precipitation or protect the ESP from warpage. For combustion processes with exhaust gas temperatures below approximately 700°F, cooling would not be required, and the exhaust gases can be delivered directly to the precipitator.

When much of the pollutant loading consists of relatively large particles, mechanical collectors, such as cyclones, may be used to reduce the load on the ESP, especially at high inlet concentrations. The fans provide the motive power for air movement and can be mounted before or after the ESP. A stack, normally used, vents the cleaned stream to the atmosphere. Screw conveyors or pneumatic systems are often used to remove captured dust from the bottom of the hoppers.

Wet ESPs require a source of wash water to be injected or sprayed near the top of the collector plates either continuously or at timed intervals. The water flows with the collected particles

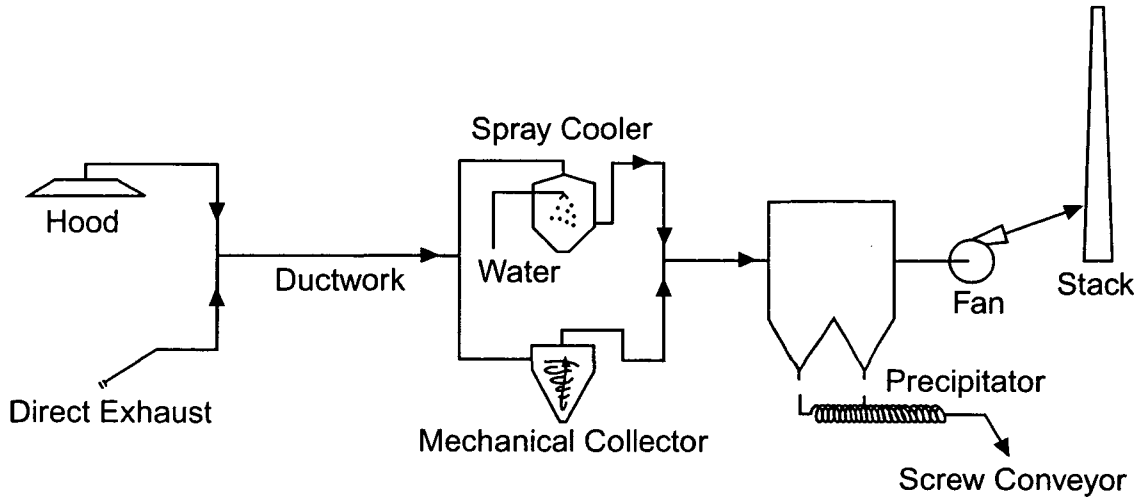


Figure 3.3: Control Device and Typical Auxiliary Equipment

into a sump from which the fluid is pumped. A portion of the fluid may be recycled to reduce the total amount of water required. The remainder is pumped directly to a settling pond or passed through a dewatering stage, with subsequent disposal of the sludge.

Gas conditioning equipment to improve ESP performance by changing dust resistivity is occasionally used as part of the original design, but more frequently it is used to upgrade existing ESPs. The equipment injects an agent into the gas stream ahead of the ESP. Usually, the agent mixes with the particles and alters their resistivity to promote higher migration velocity, and thus higher collection efficiency. However, electrical properties of the gas may change, rather than dust resistivity. For instance, cooling the gas will allow more voltage to be applied before sparking occurs. Significant conditioning agents that are used include SO_3, H_2SO_4, sodium compounds, ammonia, and water, but the major conditioning agent by usage is SO_3. A typical dose rate for any of the gaseous agents is 10 to 30 ppm by volume.

The equipment required for conditioning depends on the agent being used. A typical SO_3 conditioner requires a supply of molten sulfur. It is stored in a heated vessel and supplied to a burner, where it is oxidized to SO_2. The SO_2 gas is passed over a catalyst for further oxidation to SO_3. The SO_3 gas is then injected into the flue gas stream through a multi-outlet set of probes that breach a duct. In place of a sulfur burner to provide SO_2, liquid SO_2 may be vaporized from a storage tank. Although their total annual costs are higher, the liquid SO_2 systems have lower capital costs and are easier to operate than the molten sulfur based systems.

Water or ammonia injection requires a set of spray nozzles in the duct, along with pumping and control equipment.

Sodium conditioning is often done by coating the coal on a conveyor with a powder compound or a water solution of the desired compound. A hopper or storage tank is often positioned over the conveyor for this purpose.

3.1.4 Electrostatic Precipitation Theory

The theory of ESP operation requires many scientific disciplines to describe it thoroughly. The ESP is basically an electrical machine. The principal actions are the charging of particles and forcing them to the collector plates. The amount of charged particulate matter affects the electrical operating point of the ESP. The transport of the particles is affected by the level of turbulence in the gas. The losses mentioned earlier, sneakage and rapping reentrainment, are major influences on the total performance of the system. The particle properties also leave a major effect on the operation of the unit.

The following subsections will explain the theory behind (1) electrical operating points in the ESP, (2) particle charging, (3) particle collection, and (4) sneakage and rapping reentrainment. General references for these topics are White [1] or Lawless and Sparks [2].

3.1.4.1 Electrical Operating Point

The electrical operating point of an ESP section is the value of voltage and current at which the section operates. As will become apparent, the best collection occurs when the highest electric field is present, which roughly corresponds to the highest voltage on the electrodes. In this work, the term "section" represents one set of plates and electrodes in the direction of flow. This unit is commonly called a "field", and a "section" or "bus section" represents a subdivision of a "field" perpendicular to the direction of flow. In an ESP model and in sizing applications, the two terms "section" and "field" are used equivalently because the subdivision into bus sections should have no effect on the model. This terminology has probably arisen because of the frequent use of the word "field" to refer to the electric field.

The lowest acceptable voltage is the voltage required for the formation of a corona, the electrical discharge that produces ions for charging particles. The (negative) corona is produced when an occasional free electron near the high-voltage electrode, produced by a cosmic ray, gains enough energy from the electric field to ionize the gas and produce more free electrons. The electric field for which this process is self-sustained has been determined experimentally. For round wires, the field at the surface of the wire is given by:

$$E_c = 3.126 \times 10^6 d_r \left[1 + 0.0301 \left(\frac{d_r}{r_w} \right)^{0.5} \right] \qquad (3.1)$$

where

E_e	=	corona onset field at the wire surface (V/m)
d_r	=	relative gas density, referred to 1 atm pressure and 20°C (dimensionless)
r_w	=	radius of the wire, meters (m)

This is the field required to produce "glow" corona, the form usually seen in the laboratory on smooth, clean wires. The glow appears as a uniform, rapidly moving diffuse light around the electrode. After a period of operation, the movement concentrates into small spots on the wire surface, and the corona assumes a tuft-like appearance. The field required to produce "tuft" corona, the form found in full-scale ESPs, is 0.6 times the value of E_c.

The voltage that must be applied to the wire to obtain this value of field, V_c, is found by integrating the electric field from the wire to the plate. The field follows a simple "$1/r$" dependence in cylindrical geometry. This leads to a logarithmic dependence of voltage on electrode dimensions. In the plate-wire geometry, the field dependence is somewhat more complex, but the voltage still shows the logarithmic dependence. V_c is given by:

$$V_c = E_c r_w \ln \left(\frac{d}{r_w} \right) \qquad (3.2)$$

where

V_c	=	corona onset voltage (V)
d	=	outer cylinder radius for tubular ESP (m) $4/B$ x (wire-plate separation) for plate-wire ESP (m)

No current will flow until the voltage reaches this value, but the amount of current will increase steeply for voltages above this value. The maximum current density (amperes/square meter) on the plate or cylinder directly under the wire is given by:

$$j = \mu \in \frac{V^2}{L^3} \tag{3.3}$$

where

j	=	maximum current density (A/m^2)
μ	=	ion mobility m^2/Vs) ($meter^2$/volt second)
$\in$	=	free space permittivity (8.845×10^{-12} F/m)(Farad/meter)
V	=	applied voltage (V)
L	=	shortest distance from wire to collecting surface (m)

For tuft corona, the current density is zero until the corona onset voltage is reached, when it jumps almost to this value of j within a few hundred volts, directly under a tuft.

The region near the wire is strongly influenced by the presence of ions there, and the corona onset voltage magnitude shows strong spatial variations. Outside the corona region, it is quite uniform.

The electric field is strongest along the line from wire to plate and is approximated very well, except near the wire, by:

$$E_{max} = \frac{V}{L} \tag{3.4}$$

where

E_{max}	=	maximum field strength (V/m)

When the electric field throughout the gap between the wire and the plate becomes strong enough, a spark will occur, and the voltage cannot be increased without severe sparking occurring. The field at which sparking occurs is not sharply defined, but a reasonable value is given by:

$$E_s = 6.3 \times 10^5 \left(\frac{273}{T} P \right)^{1.65} \tag{3.5}$$

where

E_s	=	sparking field strength (V/m)
T	=	absolute temperature (K)
P	=	gas pressure (atm)

This field would be reached at a voltage of, for example, 35,000 V for a plate-wire spacing of 11.4 cm (4.5 in.) at a temperature of 149°C (300°F). The ESP will generally operate near this voltage in the absence of back corona. E_{max} will be equal to or less than E_s.

Instead of sparking, back corona may occur if the electric field in the dust layer, resulting from the current flow in the layer, reaches a critical value of about 1 x 10^6 V/m. Depending on conditions, the back corona, may enhance sparking or may generate so much current that the voltage cannot be raised any higher. The field in the layer is given by:

$$E_l = j\rho \tag{3.6}$$

where

E_l = electric field in dust layer (V/m)
ρ = resistivity of the collected material (ohm-m)

3.1.4.2 Particle Charging

Charging of particles takes place when ions bombard the surface of a particle. Once an ion is close to the particle, it is tightly bound because of the image charge within the particle. The "image charge" is a representation of the charge distortion that occurs when a real charge approaches a conducting surface. The distortion is equivalent to a charge of opposite magnitude to the real charge, located as far below the surface as the real charge is above it. The notion of the fictitious charge is similar to the notion of an image in a mirror, hence the name. As more ions accumulate on a particle, the total charge tends to prevent further ionic bombardment.

There are two principal charging mechanisms: diffusion charging and field charging. Diffusion charging results from the thermal kinetic energy of the ions overcoming the repulsion of the ions already on the particle. Field charging occurs when ions follow electric field lines until they terminate on a particle. In general, both mechanisms are operative for all sizes of particles. Field charging, however, adds a larger percentage of charge on particles greater than about 2μm in diameter, and diffusion charging adds a greater percentage on particles smaller than about 0.5μm.

Diffusion charging, as derived by White [1], produces a logarithmically increasing level of charge on particles, given by:

$$q(t) = \left(\frac{rkT}{e}\right) \ln(1 + r) \tag{3.7}$$

where

$q(t)$	=	particle charge (C) as function of time, t, in seconds
r	=	particle radius (m)
k	=	Boltzmann's constant (j/K)
T	=	absolute temperature (K)
e	=	electron charge ($1.67 \times 10^{-19} C$)
τ	=	dimensionless time given by:

$$\tau = \frac{\pi r v \ N e^2 \theta}{kT} \tag{3.8}$$

where

v	=	mean thermal speed of the ions (m/s)
N	=	ion number concentration near the particle (No./m^3) 2 = real time (exposure
θ	=	real time (exposure time in the charging zone) (s)

Diffusion charging never reaches a limit, but it becomes very slow after about three dimensionless time units. For fixed exposure times, the charge on a particle is proportional to its radius.

Field charging also exhibits a characteristic time-dependence, given by:

$$q(t) = \frac{q_s \theta}{\theta + \tau'} \tag{3.9}$$

where

q_s	=	saturation charge, charge at infinite time (C)
θ	=	real time (s)
τ'	=	another dimensionless time unit

The saturation charge is given by:

$$q_s = 12 \pi \epsilon r^2 E \tag{3.10}$$

where

ϵ	=	free space permittivity (F/m)
E	=	external electric field applied to the particle (V/m)

The saturation charge is proportional to the square of the radius, which explains why field charging is the dominant mechanism for larger particles. The field charging time constant is given by:

$$r' = \frac{4 \epsilon}{Ne\mu} \tag{3.11}$$

where

μ = ion mobility

Strictly speaking, both diffusion and field charging mechanisms operate at the same time on all particles, and neither mechanism is sufficient to explain the charges measured on the particles. It has been found empirically that a very good approximation to the measured charge is given by the sum of the charges predicted by Equations 3.7 and 3.9 independently of one another:

$$q_{tot} = q_d(t) + q_f(t) \tag{3.12}$$

where

$q_{tot}(t)$ = particle charge due to both mechanisms
$q_d(t)$ = particle charge due to diffusion charging
$q_f(t)$ = particle charge due to field charging

3.1.4.3 Particle Collection

The electric field in the collecting zone produces a force on a particle proportional to the magnitude of the field and to the charge:

$$F_e = qE \tag{3.13}$$

where

F_e = force due to electric field (N)
q = charge on particle (C)
E = electric field (V/m)

Because the field charging mechanism gives an ultimate charge proportional to the electric field, the force on large particles is proportional to the square of the field, which shows the advantage for maintaining as high a field as possible.

The motion of the particles under the influence of the electric field is opposed by the viscous drag of the gas. By equating the electric force and the drag force component due to the electric field (according to Stokes' law), we can obtain the particle velocity:

$$v(q,E,r) = \frac{q(E,r)\ E\ C(r)}{6\pi\ \eta\ r} \tag{3.14}$$

where

$v(q,E,r)$	=	particle velocity (m/s)
$q(E,r)$	=	particle charge (C)
$C(r)$	=	Cunningham correction to Stokes' law (dimensionless)
η	=	gas viscosity (kg/ms)

The particle velocity, is the rate at which the particle moves along the electric field lines, *i.e.*, toward the walls.

For a given electric field, this velocity is usually at a minimum for particles of about 0.5 μm diameter. Smaller particles move faster because the charge does not decrease very much, but the Cunningham factor increases rapidly as radius decreases. Larger particles have a charge increasing as r^2 and a viscous drag only increasing as r^1; the velocity then increases as r.

Equation 3.14 gives the particle velocity with respect to still air. In the ESP, the flow is usually very turbulent, with instantaneous gas velocities of the same magnitude as the particles velocities, but in random directions. The motion of particles toward the collecting plates is therefore a statistical process, with an average component imparted by the electric field and a fluctuating component from the gas turbulence.

This statistical motion leads to an exponential collection equation, given by:

$$N(r) = N_0(r)\ exp\left(\frac{-vr}{v_0}\right) \tag{3.15}$$

where

$N(r)$	=	particle concentration of size r at the exit of the collecting zone (No./m^3)
$N_0(r)$	=	particle concentration of size r at the entrance of the zone (No./m^3)
$v(r)$	=	size-dependent particle velocity (m/s)
v_o	=	characteristic velocity of the ESP (m/s), given by:

$$v_0 = \frac{Q}{A} = \frac{1}{SCA} \tag{3.16}$$

where

Q	=	volume flow rate of the gas (m^3/s)
A	=	plate area for the ESP collecting zone (m^2)
SCA	=	specific collection area (A/Q) (s/m)

When this collection equation is averaged over all the particle sizes and weighted according to the concentration of each size, the Deutsch equation results, with the penetration (fraction of particles escaping) given by:

$$p = exp(-w_e\ SCA) \tag{3.17}$$

where

p	=	penetration (fraction)
w_e	=	effective migration velocity for the particle ensemble (m/s)

The collection efficiency is given by:

$$Eff\,(\%) = 100\ (1 - p) \tag{3.18}$$

and is the number most often used to describe the performance of an ESP.

3.1.4.4 Sneakage and Rapping Reentrainment

Sneakage and rapping reentrainment are best considered on the basis of the sections within an ESP. Sneakage occurs when a part of the gas flow bypasses the collection zone of a section. Generally, the portion of gas that bypasses the zone is thoroughly mixed with the gas that passes through the zone before all the gas enters the next section. This mixing cannot always be

assumed, and when sneakage paths exist around several sections, the performance of the whole ESP is seriously elected. To describe the effects of sneakage and rapping reentrainment mathematically we first consider sneakage by itself and then consider the effects of rapping as an average over many rapping cycles.

On the assumption that the gas is well mixed between sections, the penetration for each section can be expressed as:

$$p_s = S_N + \left[(1 - S_N) \times p_c(Q')\right] \tag{3.19}$$

where

p_s = section's fractional penetration
S_N = fraction of gas bypassing the section (sneakage)
$p_c(Q')$ = fraction of particles penetrating the collection zone, which is functionally dependent on Qt, the gas volume flow in the collection zone,reduced by the sneakage (m^3/s)

The penetration of the entire ESP is the product of the section penetrations. The sneakage sets a lower limit on the penetration of particles through the section.

To calculate the effects of rapping, we first calculate the amount of material captured on the plates of the section. The fraction of material that was caught is given by:

$$\frac{m}{m_0} = 1 - p_s = 1 - S_N - \left[(1 - S_N) \times p_c(Q')\right] \tag{3.20}$$

where

m/m_o = mass fraction collected from the gas stream

This material accumulates until the plates are rapped, whereupon most of the material falls into the hopper for disposal, but a fraction of it is reentrained and leaves the section. Experimental measurements have been conducted on fly ash ESPs to evaluate the fraction reentrained, which averages about 12 percent.

The average penetration for a section including sneakage and rapping reentrainments, is:

$$p_s = S_N + \left[(1 - S_N) \times p_c(Q')\right] + RR\ (1 - S_N)\left[1 - p_c(Q')\right] \tag{3.21}$$

where

RR = fraction reentrained

This can be written in a more compact form as:

$$p_s = LF + \left[(1 - LF) \times p_c(Q')\right] \tag{3.22}$$

by substituting LF (loss factor) for $S_N + RR(1 - S_N)$. These formulas can allow for variable amounts of sneakage and rapping reentrainment for each section, but there is no experimental evidence to suggest that it is necessary.

Fly ash precipitators analyzed in this way have an average S_N of 0.07 and an RR of 0.12. These values are the best available at this time, but some wet ESPs, which presumably have no rapping losses, have shown S_N values of 0.05 or less. These values offer a means for estimating the performance of ESPs whose actual characteristics are not known, but about which general statements can be made. For instance, wet ESPs would be expected to have $RR = 0$, as would ESPs collecting wet or sticky particles. Particulate materials with a much smaller mass mean diameter, MMD, than fly ash would be expected to have a lower RR factor because they are held more tightly to the plates and each other. Sneakage factors are harder to account for; unless special efforts have been made in the design to control sneakage, the 0.07 value should be used.

3.2 ESP Design Procedure

3.2.1 Specific Collecting Area

Specific collecting area (SCA) is a parameter used to compare ESPs and roughly estimate their collection efficiency. SCA is the total collector plate, area divided by gas volume flow rate and has the units of s/m or s/ft. Since SCA is the ratio of A/Q, it is often expressed as $m^2/(m^3/s)$ or ft^2/kacfm, where kacfm is thousand acfm. SCA is also one of the most important factors in determining the capital and several of the annual costs (for example, maintenance and dust disposal costs) of the ESP because it determines the size of the unit. Because of the various ways in which SCA can be expressed, Table 3.1 gives equivalent SCAs in the different units for what would be considered a small, medium, and large SCA.

The design procedure is based on the loss factor approach of Lawless and Sparks [2] and considers a number of process parameters. It can be calculated by hand, but it is most conveniently used with a spreadsheet program. For many uses, tables of effective migration velocities can be used to obtain the SCA required for a given efficiency. In the following subsection, tables have

Table 3.1: Small, Medium, and Large SCAs as Expressed by Various Units

Units	Small	Medium	Large
ft^2/kacfm[a]	100	400	900
s/m	19.7	78.8	177
s/ft	6	24	54

[a] 5.080 ft^2/kacfm = 1 (s/m)

been calculated using the design procedure for a number of different particle sources and for differing levels of efficiency. If a situation is encountered that is not covered in these tables, then the full procedure that appears in the subsequent subsection should be used.

3.2.1.1 SCA Procedure with Known Migration Velocity

If the migration velocity is known, then Equation 3.17 can be rearranged to give the *SCA*:

$$SCA = \frac{-\ln(p)}{w_e}$$

A graphical solution to Equation 3.23 is given in Figure 3.4. The migration velocities have been calculated for three main precipitator types: plate-wire, flat plate, and wet wall ESPs of the plate-wire type. The following three tables, keyed to design efficiency as an easily quantified variable, summarize the migration velocities under various conditions:

- In Table 3.2, the migration velocities are given for a plate-wire ESP with conditions of no back corona and severe back corona; temperatures appropriate for each process have been assumed.

- In Table 3.3, the migration velocities calculated for a wet wall ESP of the plate-wire type assume no back corona and no rapping reentrainment.

- In Table 3.4, the flat plate ESP migration velocities are given only for no back corona conditions because they appear to be less affected by high-resistivity dusts than the plate-wire types.

It is generally expected from experience that the migration velocity will decrease with increasing efficiency. In Tables 3.2 through 3.4, however, the migration velocities show some

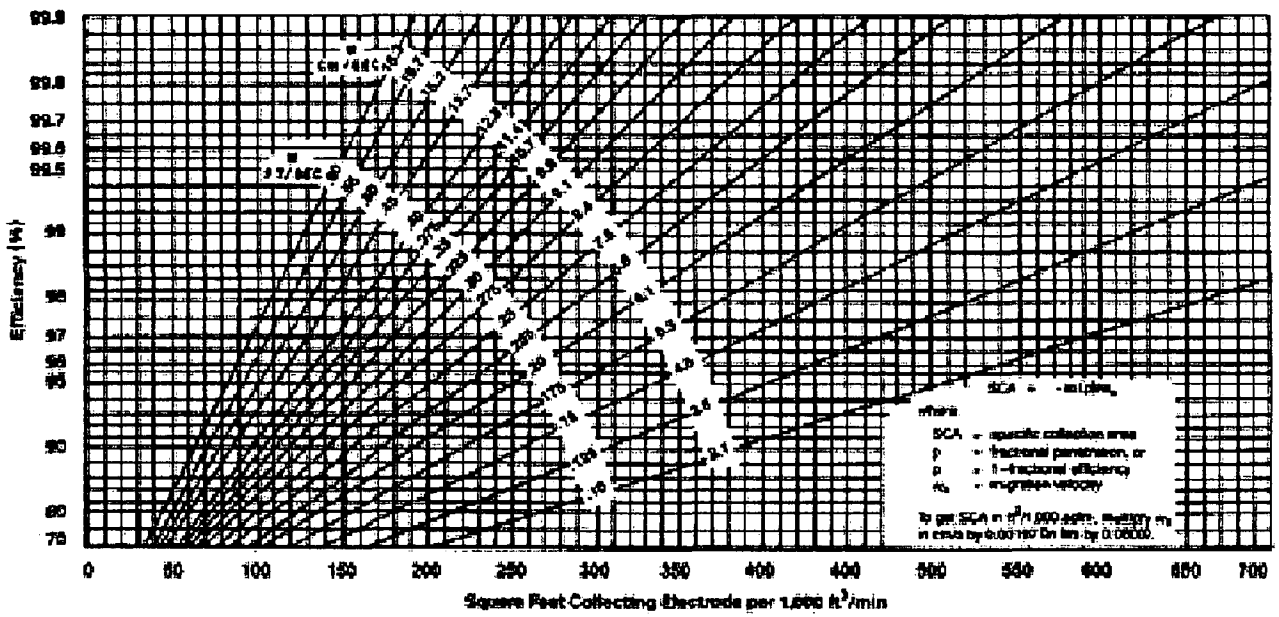


Figure 3.4: Chart for Finding SCA

Table 3.2: Plate-Wire ESP Migration Velocities (cm/s)[a]

Particle Source		Design Efficiency, % 95	99	99.5	99.9
Bituminous coal fly ash[b]	(no BC)	12.6	10.1	9.3	8.2
	(BC)	3.1	2.5	2.4	2.1
Sub-bituminous coal fly as in tangential-fired boiler[b]	(no BC)	17.0	11.8	10.3	8.8
	(BC)	4.9	3.1	2.6	2.2
Other coal[b]	(no BC)	9.7	7.9	7.9	7.2
	(BC)	2.9	2.2	2.1	1.9
Cement Kiln[c]	(no BC)	1.5	1.5	1.8	1.8
	(BC)	0.6	0.6	0.5	0.5
Glass plant[d]	(no BC)	1.6	1.6	1.5	1.5
	(BC)	0.5	0.5	0.5	0.5
Iron/steel sinter plant dust with mechanical precollector[b]	(no BC)	6.8	6.2	6.6	6.3
	(BC)	2.2	1.8	1.8	1.7
Kraft-paper recovery boiler[b]	(no BC)	2.6	2.5	3.1	2.9
Incinerator fly ash[e]	(no BC)	15.3	11.4	10.6	9.4
Copper reverberatory furnace[f]	(no BC)	6.2	4.2	3.7	2.9
Copper converter[g]	(no BC)	5.5	4.4	4.1	3.6
Copper roaster[h]	(no BC)	6.2	5.5	5.3	4.8
Coke plant combustion stack[i]	(no BC)	1.2[j]	-	-	-

BC = Back Corona

[a] To convert cm/s to fps, multiply cm/s by 0.0328m but computational precedures uses SI units. To convert cm/s to m/s, multiply by 0.01. Assumes same particle size as given in full computational procedure.

[b] At 300°F. Depending on individual furnace/boiler conditions, chemical nature of the fly ash, and availability of naturally occurring conditioning agents (e.g., moisture in the gas stream). Migration velocities may vary considerably from these values. Likely values are in the range form back corona to no back corona. BC = back corona.

[c] At 600°F.

[d] At 500°F.

[e] At 250°F.

[f] 450°F to 570°F

[g] 500°F to 700°F

[h] 600°F to 660°F

[i] 360°F to 450°F
Data available only for inlet concentrations in the range of 0.02 to 0.2 g/s m^3 and for efficiencies less than 90%.

Table 3.3: Wet Wall Plate - Wire ESP Migration Velocities
(No back corona, cm/s)[a]

Particle Source[b]	Design Efficiency, %			
	95	99	99.5	99.9
Bituminous coal fly ash	31.4	33.0	33.5	24.9
Sub-bituminous coal fly ash in tangential-fired boiler	40.0	42.7	44.1	31.4
Other coal	21.1	21.4	21.5	17.0
Cement kiln	6.4	5.6	5.0	5.7
Glass plant	4.6	4.5	4.3	3.8
Iron/steel snter plant dust with mechanical precollector	14.0	13.7	13.3	11.6

[a] To convert cm/s to ft/s, multiply cm/s by 0.0328. Computational precedure uses SI units; to convert cm/s to m/s, ultiply cm/s by 0.01. Assumes same particle size is given in full computational procedure.
[b] All sources asumed at 200°F.

fluctuations. This is because the number of sections must be increased as the efficiency increases, and the changing sectionalization affects the overall migration velocity. This effect is particularly noticeable, for example, in Table 3.4 for glass plants. When the migration velocities in the tables are used to obtain *SCA*s for the different efficiencies in the tables , the *SCA*s will increase as the officiency increases.

3.2.1.2 Full SCA Procedure

The full procedure for determining the *SCA* for large plate-wire, flat plate, and (with restrictions) tubular dry ESPs is given here. This procedure does not apply to the smaller, two-stage precipitators because these are packaged modules generally sized and sold on the basis of the waste gas volumetric flow rate. Nor does this procedure apply to determining the *SCA* for wet ESPs. The full procedure consists of the 15 steps given below:

Step 1 – Determine the design collection efficiency, Eff (%). Efficiency is the most commonly used term in the industry and is the reference value for guarantees. However, if it has not been specified, it can be computed as follows:

$$Eff\,(\%) = 100 \times \left(1 - \frac{\text{outlet load}}{\text{inlet load}} \right) \qquad (3.18a)$$

Table 3.4: Flat Plate ESP Migratoin Velocities[a]

Particle Source	Design Efficiency, % 95	99	99.5	99.9
Biguminous coal fly ash[c]	13.2	15.1	18.6	16.0
Sub-bituminous coal fly ash in tangential-fired boiler[c]	28.6	18.2	21.2	17.7
Other coal[c]	15.5	11.2	151	13.5
Cement kiln[d]	2.4	2.3	3.2	3.1
Glass plant[e]	1.8	1.9	2.6	2.6
Iron/steel sinter plant dust with mechanical precollector[c]	13.4	12.1	13.1	12.4
Kraft-paper recovery boiler[c]	5.0	4.7	6.1	5.3
Incinerator fly ash[f]	25.2	16.9	21.1	18.3

[a] Assumes same particle size as given in full computational procedure. These values give the grounded collector plate SCA, from which the collector plate area is derived. In flat plate ESPs, the discharge or high-voltage plate area is typically 40% of the ground-plate area. The flat plate manufacturer usually counts all the plate area (collector plates plus discharge plates in meeting an SCA specificiation, which means that the velocities tabulated above must be divided by 1.4 to be used on the manufacturer's basis. BC = back corona.

[b] To convert cm/s to ft/s, multiply cm/s by 0.0328. computational procedure uses SI units; to convert cm/s to m/s, multiply cm/s by 0.01.

[c] At 300°F.

[d] At 600°F.

[e] At 500°F.

[f] At 250°F.

Step 2 – Compute design penetration, p:

$$p = 1 - \left(\frac{Eff}{100}\right) \tag{3.17a}$$

Step 3 – Compute or obtain the operating temperature, T_k, °K. Temperature in Kelvin is required in the calculations which follow.

Step 4 – Determine whether severe back corona is present. Severe back corona usually occurs for dust resistivities above 2×10^{11} ohm-cm. Its presence will greatly increase the size of the ESP required to achieve a certain efficiency.

Step 5 – Determine the *MMD* of the inlet particle distribution MMD_i (μm). If this is not known, assume a value from the following table:

Table 3.5

Source	MMD_1(μm)
Bituminous coal	16
Sub-bituminous coal, tangential boiler	21
Sub-bituminous coal, other boiler types	10 to 15
Cement kiln	2 to 5
Glass plant	1
Wood burning boiler	5
Sinter plant,	50
with mechanical precollector	6
Kraft process recovery	2
Incinerators	15 to 30
Copper reverberatory furnace	1
Copper converter	1
Coke plant combustion stack	1
Unknown	1

Step 6 - Assume value for sneakage, S_N, and rapping reentrainment, *RR*, fromthe following tables:

Table 3.6

ESP Type	S_N
Plate-wire	0.07
Wet wall	0.05
Flat plate	0.10

Table 3.7

ESP/Ash Type	RR
Coal fly ash, or not known	0.14
Wet wall	0.0
Flat plate with gas velocity > 1.5 m/s (not glass or cement	0.15
Glass or cement	0.10

Step 7 – Assume values for the most penetrating size, *MMDp*, and rapping puff size, MMD_r:

$$MMD_p = 2\ \mu m \tag{3.24}$$
$$MMD_r = 5\ m \text{ for ash with } MMD_i > 5\ \mu m \tag{3.25}$$
$$MMD_r = 3\ m \text{ for ash with } MMD_i < 5\ \mu m \tag{3.26}$$

where

MMD_p = *MMD* of the size distribution emerging from a very efficient collecting zone
MMD_p = *MMD* of the size distribution of rapped/reentrained material.

Step 8 – Use or compute the following factors for pure air:

$$\epsilon_0 = 8.845 \times 10^{-12} \text{ free space permittivity} \left(\frac{F}{m}\right) \tag{3.27}$$

$$\eta = 1.72 \times 10^{-5} \left(\frac{T_k}{273}\right)^{0.71} \text{ gas viscosity } \left(\frac{kg}{m \cdot s}\right) \tag{3.28}$$

$$E_{bd} = 630{,}000\left(\frac{273}{T_k}\right)^{1.65} \text{ electric field at sparking } \left(\frac{\text{V}}{\text{m}}\right) \quad (3.29)$$

$$LF = S_N + RR\,(1 - S_N) \text{ loss factor (dimensionless)} \quad (3.30)$$

For plate-wire ESPs:

$$E_{avg} = \frac{E_{ba}}{1.75} \text{ average field with no back corona} \quad (3.31)$$

$$E_{avg} = 0.7 \times \frac{E_{ba}}{1.75} \text{ average field with severe back corona} \quad (3.32)$$

For flat plate ESPs:

$$E_{avg} = E_{bd} \times \frac{5}{6.3} \text{ average field, no back corona, positive polarity} \quad (3.33)$$

$$E_{avg} = 0.7 \times E_{bd} \times \frac{5}{6.3} \text{ average field, severe back corona, positive polarity} \quad (3.34)$$

Step 9 – Assume the smallest number of sections for the ESP, n, such that $LF^n < p$. Suggested values of *n* are:

These values are for an LF of 0.185, corresponding to a coal fly ash precipitator. The values are approximate, but the best results are for the smallest allowable *n*.

Table 3.8

Efficiency (%)	*n*
<96.5	2
<99	3
<99.8	4
<99.9	5
<99.9	6

Step 10 – Compute the average section penetration, p_s:

$$p_s = p^{\frac{1}{n}} \tag{3.35}$$

Step 11 – Compute the section collection penetration, p_c:

$$p_c = \frac{p_s - LF}{1 - LF} \tag{3.36}$$

If the value of n is too small, then this value will be negative and n must be increased.

Step 12 – Compute the particle size change factors, D and MMD_{rp}, which are constants used for computing the change of particle size from section to section:

$$\begin{aligned} D = p_s &= S_N + P_c\,(1 - S_N) + RR\,(1 - S_N)\,(1 - p_c) \\ &= MMD_{rp} = RR\,(1 - S_N)\,(1 - p_c)\,\frac{MMD_r}{D} \end{aligned} \tag{3.37}$$

Step 13 - Compute a table of particle sizes for sections 1 through n:

Table 3.9

Section	MMDs
1	MMD1 = MMDi
2	MMD2 = {MMD1 x S_N + [1 - p_c) x MMD_p + p_c x MMD1] x p_c}/D + MMDrp
3	MMD3 = {MMD2 x S_N + [1 - p_c) x MMD_p + p_c x MMD2] x p_c}/D + MMDrp
.	.
.	.
.	.
n	MMDn = {MMDn - 1 x S_N + {1 - p_c) x MMD_p + p_c x MMDn - 1] x p_c}/D + MMDrp

Step 14 - Calculate the SCA for sections 1 through n, using MMD_n, η, ϵ, E_{avg}, and p_e:

$$SCA_1 = -\left(\frac{\eta}{\epsilon}\right) \times (1 - S_N) \times \frac{\ln(p_c)}{E_{avg}^2 \times MMD_1 \times 10^{-6}} \tag{3.38}$$

$$SCA_n = -\left(\frac{\eta}{\epsilon}\right) \times (1 - S_N) \times \frac{\ln(p_c)}{E_{avg}^2 \times MMD_n \times 10^{-6}} \tag{3.39}$$

where the factor 10^{-6} converts micrometers to meters. Note that the only quantity changing in these expressions is MMD_x; therefore, the following relation can be used:

$$SCA_{n+1} = SCA_n \times \frac{MMD_n}{MMD_{n+1}} \tag{3.40}$$

Step 15 – Calculate the total *SCA* and the English *SCA, ESCA*:

$$SCA\left(\frac{\text{s}}{\text{m}}\right) = \sum_{i=1}^{n} SCA_i \tag{3.41}$$

$$ESCA\left(\frac{\text{ft}^2}{\text{kacfm}}\right) = 5.080 \times SCA\left(\frac{\text{s}}{\text{m}}\right) \tag{3.42}$$

This sizing procedure works best for p_c values less than the value of LF, which means the smallest value of n. Any ESP model is sensitive to the values of particle diameter and electric field. This one shows the same sensitivity, but the expressions for electric field are based on theoretical and experimental values. The *SCA* should not be strongly affected by the number of sections chosen; if more sections are used, the *SCA* of each section is reduced.

3.2.1.3 Specific Collecting Area for Tubular Precipitators

The procedure given above is suitable for large plate-wire or flat plate ESPs, but must be used with restrictions for tubular ESPs. Values of $S_N = 0.015$ and $RR = 0$ are assumed, and only one section is used.

Table 3.10 gives migration velocities that can be used with Equation 3.23 to calculate *SCA*s for several tubular ESP applications.

3.2.2 Flow Velocity

A precipitator collecting a dry particulate material runs a risk of nonrapping (continuous) reentrainment if the gas velocity becomes too large. This effect is independent of *SCA* and has

been learned through experience. For fly ash applications, the maximum acceptable velocity is about 1.5 m/s (5 ft/s) for plate-wire ESPs and about 1 m/s (3 ft/s) for flat plate ESPs. For low resistivity applications, design velocities of 3 ft/s or less are common to avoid nonrapping reentrainment. The frontal area of the ESP (W x H), e.g., the area normal to the direction of gas flow, must be chosen to keep gas velocity low and to accommodate electrical requirements (e.g., wire-to-plate spacing) while also ensuring that total plate area requirements are met. This area can be configured in a variety of ways. The plates can be short in height, long in the direction of flow, with several in parallel (making the width narrow). Or, the plates can be tall in height, short in the

Table 3.10: Tubular ESP Migration Velocities (cm/s)[b]

Particle Source		Design Efficiency, % 90	95
Cement kiln	(no BC)	2.2-5.4	2.1-5.1
	(BC)	1.1-2.7	1.0-2.6
Glass plant	(no BC)	1.4	1.3
	(BC)	0.7	0.7
Kraft-paper recovery boiler	(no BC)	4.7	4.4
Incinerator 15 μm MMD	(no BC)	40.8	39.
Wet, at 200°F MMD(μm)			
1		3.2	3.1
2		6.4	6.2
5		16.1	15.4
10		32.2	30.8
20		64.5	61.6

BC = Back corona

[a] These rates were calculated on the basis of: $S_N = 0.015$, RR = 0, one seciton only. These are in agreement with operating tubular ESPs; extension of results to more than one section is not recommended.

[b] To convert cm/s to ft/s, multiply cm/s by 0.0328.

direction of flow, with many in parallel making the width large). After selecting a configuration, the gas velocity can be obtained by dividing the volume flow rate, Q, by the frontal area of the ESP:

$$v_{gas} = \frac{Q}{WH} \tag{3.43}$$

where:

v_{gas}	=	gas velocity (m/s)
W	=	width of ESP entrance (m)
H	=	height of ESP entrance (m)

When meeting the above restrictions, this value of velocity also ensures that turbulence is not strongly developed, thereby assisting in the capture of particles.

3.2.3 Pressure Drop Calculations

The pressure drop in an ESP is due to four main factors:

- Diffuser plate (usually present)—(perforated plate at the inlet)
- Transitions at the ESP inlet and outlet
- Collection plate baffles (stiffeners) or corrugations
- Drag of the flat collection plate

The total pressure drop is the sum of the individual pressure drops, but any one of these sources may dominate all other contributions to the pressure drop. Usually, the pressure drop is not a design-driving factor, but it needs to be maintained at an acceptably low value. Table 3.11 gives typical pressure drops for the four factors. The ESP pressure drop, usually less than about 0.5 in. H_2O, is much lower than for the associated collection system and ductwork. With the conveying velocities used for dust collected in ESPs, generally 4,000 ft/min or greater, system pressure drops are usually in the range of 2 to 10 in H_2O, depending upon the ductwork length and configuration as well as the type(s) of preconditioning device(s) used upstream.

The four main factors contributing to pressure drop are described briefly below.

The diffuser plate is used to equalize the gas flow across the face of the ESP. It typically consists of a flat plate covered with round holes of 5 to 7 cm diameter (2 to 2.5 in.) having an open area of 50 to 65 percent of the total. Pressure drop is strongly dependent on the percent open area, but is almost independent of hole size.

Table 3.11: Components of ESP Pressure Drop

Component	Typical Pressure Drop (in. H_2O)	
	Low	High
Diffuser	0.010	0.09
Inlet transition	0.07	0.14
Outlet transition	0.007	0.015
Baffles	0.0006	0.123
Collection plates	0.0003	0.008
Total	**0.09**	**0.38**

The pressure drop due to gradual enlargement at the inlet is caused by the combined effects of flow separation and wall friction and is dependent on the shape of the enlargement. At the ESP exit, the pressure drop caused by a short, well-streamlined gradual contraction is small.

Baffles are installed on collection plates to shield the collected dust from the gas flow and to provide a stiffening effect to keep the plates aligned parallel to one another. The pressure drop due to the baffles depends on the number of baffles, their protrusion into the gas stream with respect to electrode-to-plate distance, and the gas velocity in the ESP.

The pressure drop of the flat collection plates is due to friction of the gas dragging along the flat surfaces and is so small compared to other factors that it may usually be neglected in engineering problems.

3.2.4 Particle Characteristics

Several particle characteristics are important for particle collection. It is generally assumed that the particles are spherical or spherical enough to be described by some equivalent spherical diameter. Highly irregular or elongated particles may not behave in ways that can be easily described.

The first important characteristic is the mass of particles in the gas stream, *i.e.*, the particle loading. This quantity usually is determined by placing a filter in the gas stream, collecting a known volume of gas, and determining the weight gain of the filter. Because the ESP operates over a wide range of loadings as a constant efficiency device, the inlet loading will determine the outlet loading directly. If the loading becomes too high, the operation of the ESP will be altered, usually for the worse.

The second characteristic is the size distribution of the particles, often expressed as the cumulative mass less than a given particle size. The size distribution describes how many particles of a given size there are, which is important because ESP efficiency varies with particle size. In practical terms, an ESP will collect all particles larger than 1.0 μm in diameter better than ones smaller than 10 μm. Only if most of the mass in the particles is concentrated above 10 μm would the actual size distribution above 10 μm be needed.

In lieu of cumulative mass distributions, the size distribution is often described by log-normal parameters. That is, the size distribution appears as a probabilistic normal curve if the logarithm of particle size used is the abscissa. The two parameters needed to describe a log-normal distribution are the mass median (or mean) diameter and the geometric standard deviation.

The *MMD* is the diameter for which one-half of the particulate mass consists of smaller particles and one-half is larger (see the Procedure, Step 5, of Subsection 3.2.1.2). If the *MMD* of a distribution is larger than about 3 μm, the ESP will collect all particles larger than the *MMD* at least as well as a 3 μm particle, representing one-half the mass in the inlet size distribution.

The geometric standard deviation is the equivalent of the standard deviation of the normal distribution: It describes how broad the size distribution is. The geometric standard deviation is computed as the ratio of the diameter corresponding to 84 percent of the total cumulative mass to the *MMD*; it is always a number greater than 1. A distribution with particles of all the same size (monodisperse) has a geometric standard deviation of 1. Geometric standard deviations less than 2 represent rather narrow distributions. For combustion sources, the geometric standard deviations range from 3 to 5 and are commonly in the 3.5 to 4.5 range.

A geometric standard deviation of 4 to 5, coupled with an *MMD* of less than 5 μm, means that there is a substantial amount of submicrometer material. This situation may change the electrical conditions in an ESP by the phenomenon known as "space charge quenching", which results in high operating voltages but low currents. It is a sign of inadequate charging and reduces the theoretical efficiency of the ESP. This condition must be evaluated carefully to be sure of adequate design margins.

3.2.5 Gas Characteristics

The gas characteristics most needed for ESP design are the gas volume flow and the gas temperature. The volume flow, multiplied by the design *SCA*, gives the total plate area required for the ESP. If the volume flow is known at one temperature, it may be estimated at other temperatures by applying the ideal gas law. Temperature and volume uncertainties will outweigh inaccuracies of applying the ideal gas law.

The temperature of the gas directly affects the gas viscosity, which increases with temperature. Gas viscosity is affected to a lesser degree by the gas composition, particularly the water vapor content. In lieu of viscosity values for a particular gas composition, the viscosity for air may be used. Viscosity enters the calculation of *SCA* directly, as seen in Step 14 of the design procedure.

The gas temperature and composition can have a strong effect on the resistivity of the collected particulate material. Specifically, moisture and acid gas components may be chemisorbed on the particles in a sufficient amount to lower the intrimic resistivity dramatically (orders of magnitude). For other types of materials, there is almost no effect. Although it is not possible to treat resistivity here, the designer should be aware of the potential sensitivity of the size of the ESP to resistivity and the factors influencing it.

The choice of the power supply size (current capacity and voltage) to be used with a particular application may be influenced by the gas characteristics. Certain applications produce gas whose density may vary significantly from typical combustion sources (density variation may result from temperature, pressure, and composition). Gas density affects corona starting voltages and voltages at which sparking will occur.

3.2.6 Cleaning

Cleaning the collected materials from the plates often is accomplished intermittently or continuously by rapping the plates severely with automatic hammers or pistons, usually along their top edges, except in the case of wet ESPs that use water. Rapping dislodges the material, which then falls down the length of the plate until it lands in a dust hopper. The dust characteristics, rapping intensity, and rapping frequency determine how much of the material is reentrained and how much reaches the hopper permanently.

For wet ESPs, consideration must be given to handling waste waters. For simple systems with innocuous dusts, water with particles collected by the ESP may be discharged from the ESP system to a solids-removing clarifier (either dedicated to the ESP or part of the plant wastewater treatment system) and then to final disposal. More complex systems may require skimming and sludge removal, clarification in dedicated equipment, pH adjustment, and/or treatment to remove dissolved-solids. Spray water from the ESP preconditioner may be treated separately from the water used to flood the ESP collecting plates, so that the cleaner of the two treated waters may be returned to the ESP. Recirculation of treated water to the ESP may approach 100 percent.

The hopper should be designed so that all the material in it slides to the very bottom, where it can be evacuated periodically, as the hopper becomes full. Dust is removed through a valve into a dust-handling system, such as a pneumatic conveyor. Hoppers often are supplied with auxiliary heat to prevent the formation of lumps or cakes and the subsequent blockage of the dust handling system.

3.2.7 Construction Features

The use of the term "plate-wire geometry" may be somewhat misleading. It could refer to three different types of discharge electrodes: weighted wires hung from a support structure at the top of the ESP, wire frames in which wires are strung tautly in a rigid support frame, or rigid electrodes constructed from a single piece of fabricated metal. In recent years, there has been a trend toward using wire frames or rigid discharge electrodes in place of weighted wire discharge electrodes (particularly in coal-fired boiler applications). This trend has been stimulated by the user's desire for increased ESP reliability. The wire frames and rigid electrodes are less prone to failure by breakage and are readily cleaned by impulse-type cleaning equipment.

Other differences in construction result from the choice of the ratio of gas passage (flow lane) width or discharge electrode to collecting electrode spacing. Typically, discharge to collecting electrode spacing varies from 11 to 19 cm (4.3 to 7.5 in.). Having a large spacing between discharge and collecting electrodes allows higher electric fields to be used, which tends to improve dust collection. To generate larger electric fields, however, power supplies must produce higher operating voltages. Therefore, it is necessary to balance the cost savings achieved with larger electrode spacing against the higher cost of power supplies that produce higher operating voltages.

Most ESPs are constructed of mild steel. ESP shells are constructed typically of 3/16 or 1/4 in. mild steel, plate. Collecting electrodes are generally fabricated from lighter gauge mild steel. A thickness of 18 gauge is common, but it will vary with size and severity of application.

Wire discharge electrodes come in varied shapes from round to square or barbed. A diameter of 2.5 mm (0.1 in.) is common for weighted wires, but other shapes used have much larger effective diameters, e.g., 64 mm (0.25 in.) square electrodes.

Stainless steel may be used for corrosive applications, but it is uncommon except in wet ESPs. Stainless steel discharge electrodes have been found to be prone to fatigue failure in dry ESPs with impact-type electrode cleaning systems.[3]

Precipitators used to collect sulfuric acid mist in sulfuric acid plants are constructed of steel, but the surfaces in contact with the acid mist are lead-lined. Precipitators used on paper mill black liquor recovery boilers are steam-jacketed. Of these two, recovery boilers have by far the larger number of ESP applications.

3.3 Estimating Total Capital Investment

Total capital investment (TCI) for an ESP system includes costs for the ESP structure, the internals, rappers, power supply, auxiliary equipment, and the usual direct and indirect costs associated with

installing or erecting new control equipment. These costs, in **second-quarter 1987** dollars, are described in the following subsections.[8]

3.3.1 Equipment Cost

3.3.1.1 ESP Costs

Five types of ESPs are considered: plate-wire, flat plate, wet, tubular, and two-stage. Basic costs for the first two are taken from Figure 3.5, which gives the flange-to-flange, field-erected price based on required plate area and a rigid electrode design. This plate area is calculated from the sizing information given previously for the four types. Adjustments are made for standard options listed in Table 3.12. Costs for wet/tubular ESPs are discussed under Recent Trends, below, and costs for two-stage precipitators are given in a later subsection.

The costs are based on a number of actual quotes. Least squares lines have been fitted to the quotes, one for sizes between 50,000 and 1,000,000 ft^2 and a second for sizes between 10,000 and 50,000 ft^2. An equation is given for each line. Extrapolation below 10,000 or above 1,000,000 ft^2 should not be used. The reader should not be surprised if quotes are obtained that differ from these curves by as much as ±25 percent. Significant savings can be had by soliciting multiple quotes. All Units include the ESP casing, pyramidal hoppers, rigid electrodes and internal collecting plates, transformer rectifier (TR) sets and microprocessor controls, rappers, and stub supports (legs) for 4 feet clearance below the hopper discharges. The lower curve is the basic unit without the standard options. The upper curve includes all of the standard options (see Table 3.12) that are normally utilized in a modern system. These options add approximately 45 percent to the basic cost of the flange-to-flange hardware. Insulation costs are for 3 in. of field-installed glass fiber encased in a metal skin and applied on the outside of all areas in contact with the exhaust gas stream. Insulation for ductwork, fan casings, and stacks must be calculated separately.

Impact of alternative electrode designs All three designs—rigid electrode, weighted wire, and rigid frame—can be employed in most applications. Any cost differential between designs will depend on the combination of vendor experience and site-specific factors that dictate equipment size factors. The rigid frame design will cost up to 25 percent more if the mast or plate height is restricted to the same used in other designs. Several vendors can now provide rigid frame collectors with longer plates, and thus the cost differential can approach zero.

The weighted wire design uses narrower plate spacings and more internal discharge electrodes. This design is being employed less; therefore, its cost is increasing and currently is

[8] For information on escalating these prices to more current dollars, refer to the EPA report *Escalation Indexes for Air Pollution Control Costs* and updates thereto, all of which are installed on the OAQPS Technology Transfer Network (CTC Bulletin Board).

approximately the same as that for the rigid electrode collector. Below about 15,000 ft^2 of plate area, ESPs are of different design and are not normally field erected, and the costs will be significantly different from values extrapolated from Figure 3.5.

Impact of materials of construction: metal thickness and stainless steel Corrosive or other adverse operating conditions may suggest the specification of thicker metal sections in the precipitator. Reasonable increases in metal sections result in minimal cost increases. For example, collecting plates are typically constructed of 18 gauge mild steel. Most ESP manufacturers can increase the section thickness by 25 percent without significant design changes or increases in manufacturing costs of more than a few percent.

Changes in type of material can increase purchase cost of the ESP from about 30 to 50 percent for type 304 stainless steel collector plates and precipitator walls, and up to several hundred percent for more expensive materials used for all elements of the ESP. Based on the type 304 stainless steel cost, the approximate factors given in Table 3.13 can be used for other materials. Appendix A provides more detail on the effects of material thickness and type.

Lower limits for condensible stack emissions (MACT and Risk Assessment), have resulted in increased demand for Wet ESTs.[7]

Recent trends As of 1987, most of the market was in the 50,000 to 200,000 ft^2 plate area size range. ESP selling prices had increased very little over the previous 10 years because of more effective designs, increased competition from European suppliers, and a shrinking utility market.

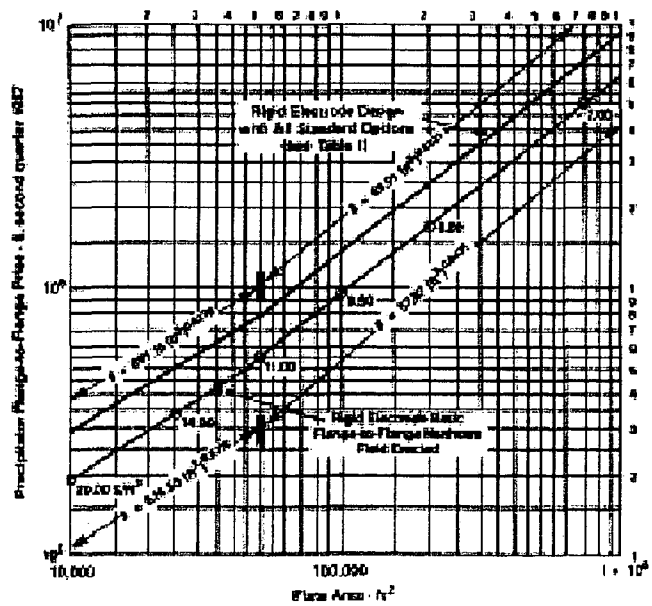


Figure 3.5: Dry-type ESP flange-to-flange purchase price vs. plate area.

Table 3.12: Standard Options for Basic Equipment

Option		Cost adder (%)
1	Inlet and outlet nozzles and diffuser plates	8 to 10
2	Hopper auxiliaries/heaters, level detectors	8 to 10
3	Weather enclosures and stair access	8 to 10
4	Structural supports	5
5	Insulation	8 to 10
	Total options 1 to 5	1.37 to 1.45 x Base

Table 3.13

Material	Factor	Reference(s)
Stainless steel, 316	1.3	[4,5,6]
Carpenter 20CB-3	1.9	[6]
Monel-400	2.3	[4,6]
Nickel-200	3.2	[6]
Titanium	4.5	[6]

Industry sources report that ESP costs (1999) have not changed significantly since 1987. [8][9] Design improvements have allowed wider plate spacings that reduce the number of internal components and higher plates and masts that provide additional plate area at a low cost. Wider plate spacing has reduced overall material and installation costs, easily compensating for any increases in material and labor costs.[7] Downflow tubular Wet ESPs use gravity to remove water and entrained particulate which has been collected onto the tubes, resulting in low operational costs. These units can be designed to intermittently wash the tubes while on line. Tubular units are typically delivered as "shop assembled" modules because the tubes are welded together in bundles. The size of the modules is limited by shipping considerations. Plate type units are usually shipped "knocked down". These units do have the same size limitations as tubular Wet ESPs, but do not require comparitively more labor for field installation. Use of high-grade alloys for corrosive applications (e.g., incinerators) increases material costs.[7] Microprocessor controls and energy management systems have lowered operating costs.

Table 3.14 lists costs (total and per acfm) for various gas volumes and removal efficiencies for Wet ESPs. For larger gas volumes, multiple modules may be used. The pricing is based on "shop assembled" modules.[7]

Table 3.14: Air Pollution Wet Electrostatic Precipitator

	80% Efficiency		85% Efficiency		90% Efficiency		95% Efficiency	
Saturated Volume (acfm)	Price ($x1000)	Price ($/acfm)	Price ($x1000)	Price ($/acfm)	Price ($x1000)	Price ($/acfm)	Price ($x1000)	Price ($/acfm)
10,000	315	31.5	327	32.7	339	33.9	365	36.5
15,000	342	22.8	355	23.7	378	25.5	408	27.2
20,000	369	18.5	385	19.3	412	20.6	451	22.6
25,000	398	16.0	423	17.0	448	18.0	---	---
30,000	427	14.3	441	14.7	---	---	---	---
35,000	442	12.7	---	---	---	---	---	---

Few, if any, hot-side ESPs (those used upstream from an air preheater on a combustion source) are being specified for purchase. Recognition that low sodium coals tend to build resistive ash layers on the collection plates, thus reducing ESP efficiency, has almost eliminated sales of these units. Of about 150 existing units, about 75 are candidates for conversion to cold-side units over the next 10 years.

Specific industry application has little impact on either ESP design or cost, with three exceptions: paper mills and sulfuric acid manufacturing plants, and coke by-product plants. Paper mill ESPs use drag conveyor hoppers that add approximately 10 percent to the base flange-to-flange equipment cost. For emissions control in sulfuric acid plants and coke byproduct ovens, wet ESPs are used. In sulfuric acid manufacture, wet ESPs are used to collect acid mist. These precipitators usually are small, and they use lead for all interior surfaces; hence, they normally cost \$65 to \$95/ft^2 of collecting area installed (mid-1987 dollars) and up to \$120/ft^2 in special situations. In addition, a wet circular ESP is used to control emissions from a coke oven off-gas detarring operation. These precipitators are made using high-alloy stainless steels and typically cost \$90 to \$120/ft^2, installed. Because of the small number of sales, small size of units sold, and dependency on site-specific factors, more definitive costs are not available.

3.3.1.2 Retrofit Cost Factor

Retrofit installations increase the costs of an ESP because of the common need to remove something to make way for the new ESP. Also, the ducting usually is much more expensive. The ducting path is often constrained by existing structures, additional supports are required, and the confined areas make erection more labor intensive and lengthy. Costs are site-specific; however, for estimating purposes, a retrofit multiplier of 1.3 to 1.5 applied to the total capital investment can be used. The multiplier should be selected within this range based on the relative difficulty of the installation.

A special case is conversion of hot-to-cold side ESPs for coal-fired boiler applications. The magnitude of the conversion is very site-specific, but most projects will contain the following elements:

- Relocating the air preheater and the ducting to it
- Resizing the ESP inlet and outlet duct to the new air volume and rerouting it
- Upgrading the ID (induced draft) fan size or motor to accommodate the higher static pressure and horsepower requirements
- Adding or modifying foundations for fan and duct supports
- Assessing the required *SCA* and either increasing the collecting area or installing an SO_3 gas-conditioning system
- Adding hopper heaters
- Upgrading the analog electrical controls to microprocessor-type controls
- Increasing the number of collecting plate rappers and perhaps the location of rapping

In some installations, it may be cost-effective to gut the existing collector totally, utilize only the existing casing and hoppers, and upgrade to modern internals.

The cost of conversion is a multimillion dollar project typically running at least 25 to 35 percent of the total capital investment of a new unit.

3.3.1.3 Auxiliary Equipment

The auxiliary equipment depicted in Figure 3.2 is discussed elsewhere in the Manual. Because dust-removal equipment (e.g., screw conveyers), hoods, precoolers, cyclones, fans, motors, and stacks are common to many pollution control systems, they are (or will be) given extended treatment in separate sections.

3.3.1.4 Costs for Two-Stage Precipitators

Purchase costs for two-stage precipitators, which should be considered separately from large-scale, single-stage ESPs, are given in Figure 3.6.[10] To be consistent with industry practice, costs are given as a function of flow rate through the system. The lower cost curve is for a two-cell unit without precooler, an installed cell washer, or a fan. The upper curve is for an engineered, package system with the following components: inlet diffuser plenum, prefilter, cooling coils with coating, coil plenums with access, water flow controls, triple pass configuration, system exhaust fan with accessories, outlet plenum, and in-place foam cleaning system with semiautomatic controls and programmable controller. All equipment is fully assembled mechanically and electrically, and it is mounted on a steel structural skid.

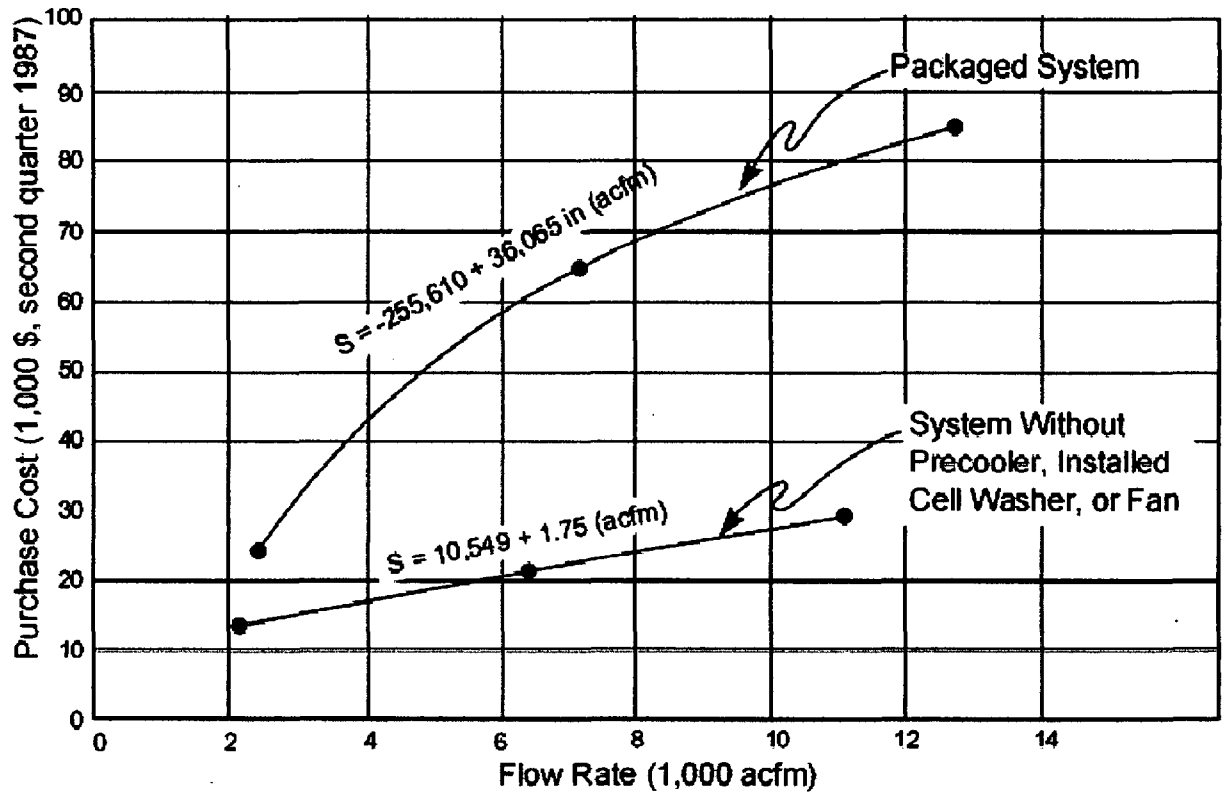


Figure 3.6: Purchase Costs for Two-stage, Two-cell Precipitators [40]

3.3.2 Total Purchased Cost

The total purchased cost of an ESP system is the sum of the costs of the ESP, options, auxiliary equipment, instruments and controls, taxes, and freight. The last three items generally are taken as percentages of the estimated total cost of the first three items. Typical values, from Section 1 of the Manual, are 10 percent for instruments and controls, 3 percent for taxes, and 5 percent for freight.

Costs of standard and other options can vary from 0 to more than 150 percent of bare ESP cost, depending on site and application requirements. Other factors that can increase ESP costs are given in Table 3.15.

Table 3.15: Items That Increase ESP Costs

Item	Factor	Applied
Rigid frame electrode with restricted plate height	1.0 to 1.25	ESP base coat
Type 304 stainless steel collector plates and precipitator walls	1.3 to 1.5	ESP base coat
All stainless steel construction	2 to 3	ESP base coat
ESP with drag conveyor hoppers (paper mill)	1.1	ESP base coat
Retrofit installations	1.3 to 1.5	ESP total capital investment (new facility installation)
Wet ESP		
Sulfuric acid mist	See 3.3.1.1	---
Sulfuric acid mist (special installation)	See 3.3.1.1	---
Coke oven off gas	See 3.3.1.1	---

3.3.3 Total Capital Investment (TCI)

Using the Section 1.2 methodology, TCI is estimated from a series of factors applied to the purchased equipment cost to obtain direct and indirect costs for installation. The TCI is the sum of these three costs. The required factors are given in Table 3.16. Because ESPs may vary from small units attached to existing buildings to large, separate structures, specific factors for site preparation or for buildings are not given. However, costs for buildings may be obtained from such references as Means Square Foot Costs 1987 [11]. Land, working capital, and off-site facilities are excluded from the table because they are not normally required. For very large installations, however, they may be needed and could be estimated on an as-needed basis.

Note that the factors given in Table 3.16 are for average installation conditions, e.g., no unusual problems with site earthwork, access, shipping, or interfering structures. Considerable variation may be seen with other-than-average installation circumstances. For two-stage precipitators purchased as packaged systems, several of the costs in Table 3.16 would be greatly reduced or eliminated. These include instruments and controls, foundations and supports, erection and handling, painting, and model studies. An installation factor of 0.20 B to 0.25 B would be more nearly appropriate for the two-stage ESPs.

3.4 Estimating Total Annual Costs

3.4.1 Direct Annual Costs

Direct annual costs include operating and supervisory labor, operating materials, replacement rappers and electrodes, maintenance (labor and materials), utilities, dust disposal, and wastewater treatment for wet ESPs. Most of these costs are discussed individually below. They vary considerably with location and time and, for this reason, should be obtained to suit the specific ESP system being costed. For example, current labor rates may be found in such publications as the Monthly Labor Review, published by the U.S. Department of Labor, Bureau of Labor Statistics.

3.4.1.1 Operating and Supervisory Labor

Proper operation of the ESP is necessary both to meet applicable particulate emission regulations and to ensure minimum costs. An ESP is an expensive piece of equipment. Even well-designed equipment will deteriorate rapidly if improperly maintained and will have to be replaced long before it should be necessary. Not only can proper operation and maintenance save the operator money, such an operation and maintenance program can also contribute to good relations with the governing pollution control agency by showing good faith in efforts to comply with air regulations.

Although each plant has its own methods for conducting an operation and maintenance program, experience has shown that plants that assign one individual the responsibility of coordinating all the pieces of the program operate better than those where different departments look after only a certain portion of the program. The separate departments have little knowledge of how their portion impacts the overall program. In other words, a plant needs one individual to coordinate the operation, maintenance, and troubleshooting components of its ESP program if it expects to have a relatively trouble-free operation. The coordinator typically is an engineer who reports to plant management and interfaces with the maintenance and plant process supervisors, the laboratory, and the purchasing department. For companies with more than one plant, he would be responsible for all ESPs. The portion of his total time that this individual spends an the ESP then becomes an operating expense for the ESP. This can be expressed as:

Table 3.16: Capital Cost Factors for ESPs[a] [26]

Cost Item	Factor
Direct Costs	
Purchased equipment costs	
ESP + auxiliary equipment	As estimated, A
Instrumentation	0.10 A
Sales taxes	0.03 A
Freight	0.05 A
Purchased equipment cost, PEC	B = 1.18 A
Direct installation costs	
Foundations & supports	0.04 B
Handling & erection	0.50 B
Electrical	0.08 B
Piping	0.01 B
Insulation for ductwork[a]	0.02 B
Painting	0.02 B
Direct installation costs	0.67 B
Site preparation	As required, SP
Buildings	As required, Bldg.
Total Direct Costs, DC[b]	1.67 B + SP + Bldg.
Indirect Costs (installation)	
Engineering	0.20 B
Construction and field expenses	0.20 B
Contractor fees	0.10 B
Start-up	0.01B
Performance test	0.01 B
Model study	0.02 B
Contingencies	0.03 B
Total Indirect Costs, IC	0.57 B
Total Capital Investment = DC + IC	2.24 B + SP + Bldg.

[a] If ductwork dimensions have been established, cost may be estimated based on \$10 to \$12/ft² of surface for field application. (Alternatively, refer to Section 2 of this Manual.) Fan housing and stacks may also be insulated. [42]

[b] For two stage precipitators, total installation direct costs are more nearly 0.20 to 0.25 + SP + Bldg.

$$AC = X\ (LCC) \tag{3.44}$$

where

AC	=	annual coordination cost ($/yr)
X	=	fraction of total time spent on ESP
LCC	=	individual annual labor cost for ESP coordinator ($/yr)

In addition to coordination costs, typical operating labor requirements are 1/2 to 2 hours per shift for a wide range of ESP sizes.[8] Small or well-performing units may require less time, and very large or troublesome units may require more time. Supervisory labor is taken as 15 percent of operating labor.

3.4.1.2 Operating Materials

Operating materials are generally not required for ESPs. An exception is the use of gas-preconditioning agents for dust resistivity control.

3.4.1.3 Maintenance

The reader should obtain Publication No. EPA/625/1-85/017, Operating and Maintenance Manual for ESPs,[13] for suggested maintenance practices. Routine ESP maintenance labor costs can be estimated using data provided by manufacturers. If such data are unavailable, the following procedure can be used. Based on data for a 100,000 ft^2 collector, maintenance labor is estimated to require 15 h/wk, 44 wk/yr. At a direct labor cost of $12.50/h (mid-1987 costs), an estimated annual maintenance labor cost of $8,250 or $0.0825/$ft^2$ of collector area is established. This relationship can be assumed to be linear above a 50,000 ft^2 collector-size and constant at $4,125 below this size. To the maintenance labor cost must be added the cost of maintenance materials. Based on an analysis of vendor information, annual maintenance materials are estimated as 1 percent of the flange-to-flange precipitator purchase cost:

$$MC = 0.01\ (FCC) + labor\ cost \tag{3.45}$$

where

MC	=	annual maintenance cost ($/yr)
FCC	=	ESP flange-to-flange purchase cost ($)
labor cost	=	$4,125 if A < 50,000 ft^2
		0.825 A if > 50,000ft^2

and

A	=	ESP plate area (ft^2)

3.4.1.4 Electricity

Power is required to operate system fans, transformer-rectifier (TR) sets, and cleaning equipment. Fan power for primary gas movement can be calculated as described in Section 2 of the Manual. After substituting into this equation a combined fan-motor efficiency of 0.65 and a specific gravity of 1.0, we obtain:

$$FP = 0.000181\,(Q)\,(\Delta P)\,(\theta') \tag{3.46}$$

where

FP	=	fan power requirement (kWh/yr)
Q	=	system flow rate (acfm)
ΔP	=	system pressure drop (in. H_2O)
θ'	=	annual operating time (h/yr)

Pump power for wet ESPs can be calculated from:

$$PP = \frac{0.746 Q_l Z S_g \theta'}{3{,}960\eta} \tag{3.47}$$

where

PP	=	pump power requirement (kWh/yr)
Q	=	water flow rate (gal/min)
Z	=	fluid head (ft)
S_g	=	specific gravity of water being pumped compared to water at 70°F and 29.92 in. Hg
θ'	=	annual operating time (h/yr)
η	=	pump-motor efficiency (fractional)

Energy for TR sets and motor-driven or electromagnetic rapper systems is the sum of the energy consumption for operating both items. Manufacturers' averaged data indicate that the following relationship can be used:

$$OP = 1.94 \times 10^{-3} A \theta^{1} \tag{3.48}$$

where

OP	=	annual ESP operating power (kWh/yr)
A	=	ESP plate area (ft^2)
θ^{l}	=	annual operating time (h/yr)

For installations requiring hopper heaters, hopper heater power can be similarly estimated:

$$HH = 2\,(HN)\,\theta' \qquad (3.49)$$

where

HH	=	annual hopper heater power consumption (kWh/yr)
HN	=	number of hoppers
θ'	=	annual operating time (h/yr)

For two-stage precipitators, power consumption ranges from 25 to 100 W/kacfm, with 40 W/kacfm being typical.

3.4.1.5 Fuel

If the ESP or associated ductwork is heated to prevent condensation, fuel costs should be calculated as required. These costs can be significant, but they may be difficult to predict. For methods of calculating heat transfer requirements, see Perry. [14]

3.4.1.6 Water

Cooling process gases for preconditioning can be done by dilution with air, evaporation with water, or heat exchange with normal equipment. Spray cooling requires consumption of plant water (heat exchangers may also require water), although costs are not usually significant. Section 1 of the Manual provides information on estimating cooling water costs. Water consumption in wet ESPs is estimated at 5 gal/min kacfm [15] for large single-stage units and 16 gal/min-kacfm for two-stage precipitators.[16]

3.4.1.7 Compressed Air

ESPs may use compressed air at pressures of about 60 to 100 psig for operating rappers. Equivalent power cost is included in Equation 3.9 for operating power.

3.4.1.8 Dust Disposal

If collected dust cannot be recycled or sold, it must be landfilled or disposed of in some other manner. Costs may typically run $20/ton or $30/ton for nonhazardous wastes exclusive of transportation (see Section 1 of the Manual). Landfilling of hazardous wastes may cost 10 times as much. The disposal costs are highly site-specific and depend on transportation distance to the landfill, handling rates, and disposal unloading (tipping) fees. If these factors are known, they lead to the relationship:

$$DD = 4.29 \times 10^{-6} G\theta' Q\,[T + (TM)\,D] \qquad (3.50)$$

where

DD	=	annual dust disposal cost ($/yr)
G	=	ESP inlet grain loading or dust concentration (gr /ft^3)
θ'	=	annual operating time (h/yr)
Q	=	gas flow rate through ESP (acfm)
T	=	tipping fee ($/ton)
TM	=	mileage rate ($/ton-mile)
D	=	dust hauling distance (miles)

3.4.1.9 Wastewater Treatment

As indicated above, the water usage for wet ESPs is about 5 gal/min kacfm.[15] Treatment cost of the resulting wastewater may vary from about $1.30 to $2.15/1,000 gal [16] depending on the complexity of the treatment system. More precise costs can be obtained from Gumerman *et al.*[18]

3.4.1.10 Conditioning Costs

Adaptation of information on utility boilers [19] suggests that SO_3 conditioning for a large ESP (2.6 x 10^6 acfm) costs from about $1.60/10^6 ft^3 of gas processed for a sulfur burner providing 5 ppm SO_3 to about $2.30/10^6 ft^3 (in first-quarter 1987 dollars) for a liquid SO_2 system providing 20 ppm of SO_3.

3.4.2 Indirect Annual Costs

Capital recovery, property taxes, insurance, administrative costs ("G&A"), and overhead are examples of indirect annual costs. The capital recovery cost is based on the equipment lifetime and the annual interest rate employed. (See Section 1 for a thorough discussion of the capital recovery cost and the variables that determine it.) For ESPs, the system lifetime varies from 5 to 40 years, with 20 years being typical. Therefore, as Section 1 of the Manual suggests, when figuring the system capital recovery cost, one should base it on the total capital investment. In other words:

$$CRC_s = TCI \times CRF_s \qquad (3.51)$$

where

CRC_s = capital recovery cost for ESP system ($/yr)
TCI = total capital investment ($)
CRF_s = capital recovery factor for ESP system (defined in Section 1.2)

For example, for a 20-year system life and a 7 percent annual interest rate, the *CRFs* would be 0.09439.

The suggested factor to use for property taxes, insurance, and administrative charges is 4 percent of the TCI. Overhead is calculated as 60 percent of the sum of operating, supervisory, coordination, and maintenance labor, as well as maintenance materials.

3.4.3 Recovery Credits

For processes that can reuse the dust collected in the ESP or that can sell the dust in a local market, such as fly ash sold as an extender for paving mixes, a credit should be taken. As used below, this credit (*RC*) appears as a negative cost.

3.4.4 Total Annual Cost

Total annual cost for owning and operating an ESP system is the sum of the components listed in Subsections 3.4.1 through 3.4.3, i.e.:

$$TAC = DC + IC - RC \tag{3.52}$$

where

TAC = total annual cost ($)
DC = direct annual cost ($)
IC = indirect annual cost ($)
RC = recovery credits (annual) ($)

3.4.5 Example Problem

Assume an ESP is required for controlling fly ash emissions from a coal-fired boiler burning bituminous coal. The flue gas stream is 50 kacfm at 325°F and has an inlet ash loading of 4 gr/ft^3. Analysis of the ash shows of 7 μm and a resistivity of less than 2 x 10^{11} ohm-cm. Assume that the ESP operates for 8,640 h/yr (360 d) and that an efficiency of 99.9 percent is required.

3.4.5.1 Design SCA

The SCA can be calculated from Equation 3.23. Assuming that a flat plate ESP design is chosen, the fly ash migration velocity is 16.0 cm/ s (see Table 3.3). Then:

$$SCA = \frac{-\ln(1-0.999)}{16.0} = 0.432\ \frac{s}{cm} = 43.2\ \frac{s}{m}$$

Converting to English units (see Step 15 in the procedure):

$$ESCA = 5.080 \times 43.2 = 219\ \frac{ft^2}{kacfm}$$

Total collector plate area is then:

$$219\ \frac{ft^2}{kacfm} \times 50\ kacfm = 10{,}950\ ft^2$$

To obtain a more rigorous answer, we can follow the steps of the procedure given in Subsection 3.2.1:

Step 1 – Design efficiency is required as 99.9.

Step 2 – Design penetration:

$$1 - \frac{99.9}{100} = 0.001$$

Step 3 – Operating temperature in Kelvin:

$$(325°F - 32°F) \times \frac{5}{9} + 273°C = 436°K$$

Step 4 – Because dust resistivity is less than 2×10^{11} ohm-cm (seeStep 4), no severe back corona is expected and back corona = 0.

Step 5 – The *MMD* of the fly ash is given as 7 μm.

Step 6 – Values for sneakage and rapping reentrainment (from table 3.1) are:

$$SN = 0.10$$
$$RR = 0.124 \text{ (assuming gas velocity <1-5 m/s)}$$

Step 7 – The most penetrating particle size, from Step 7 of the procedure is:

$$MMD_p = 2\,\mu\text{m}$$

The rapping puff size is:

$$MMD_r = 5\,\mu\text{m}$$

Step 8 – From the procedure (Subsection 3.2.1):

$$\epsilon_{\omega} = 8.845 \times 10^{-12}$$
$$\eta = 1.72 \times 10^{-5}(436/273)^{0.71} = 2.40 \times 10^{-5}$$
$$E_{bd} = 6.3 \times 10^{5}(273/436)^{1.65} = 2.91 \times 10^{5} \text{ V/m}$$
$$E_{avg} = E_{bd} \times 5/6.3 = 2.31 \times 10^{5}$$
$$LF = S_N + RR(1 - S_N) = 0.1 + 0.124(1 - 0.1) = 0.212$$

Step 9 – Choose the number of sections for $LF^n < p$, $p = 0.001$. Try four sections:

$$LF^n = 0.212^4 = 0.002$$

This value is larger than p. Try five sections:

$$LF^n = 0.212^5 = 0.000428$$

This value is smaller than p and is acceptable.

Step 10 – Average section penetration is:

$$p_s = p^{\frac{1}{n}} = 0.001^{\frac{1}{5}} = 0.251$$

Step 11 – Section collection penetration is:

$$p_c = \frac{p_s - LF}{1 - LF} = \frac{0.251 - 0.212}{1 - 0.212} = 0.0495$$

Step 12 – Particle size change factors are:

$$D = p_s + S_N + p_c\,(1 - S_N) + RR\,(1 - S_N)\,(1 - p_c)$$
$$= 0.10 + 0.0495\,(1 - 0.1) + 0.214\,(1 - 0.1)\,(1 - 0.0495)$$
$$= 0.251$$

$$MMD_{rp} = \frac{RR\,(1 - S_N)\,(1 - p_c)\,MMD_r}{D}$$
$$= \frac{0.124\,(1 - 0.1)\,(1 - 0.0495)\,(5)}{0.251}$$
$$= 2.11$$

Step 13 - Particle sizes for each section are:

Table 3.17: Mass Median Diameter

Section	MMD (μm)
1	$MMD_1 = MMD_i = 7$
2	$MMD_2 = \{MMD_1 \times S_N + [(1 - p_c) \times MMD_p + p_c \times MMD_1] \times p_c\}/D + MMD_{rp}$ = $\{7 \times 0.1 + [(1 - 0.0495) \times 2 + 0.0495 \times 7] \times 0.0495\}/0.251 + 2.11 = 5.34$
3	$MMD_3 = \{5.34 \times 0.1 + [(1-0.0495) \times 2 + 0.0495 \times 5.34] \times 0.0495\}/0.251 + 2.11 = 4.67$
4	$MMD_4 = \{4.67 \times 0.1 + [(1-0.0495) \times 2 + 0.0495 \times 4.67] \times 0.0495\}/0.251 + 2.11 = 4.39$
5	$MMD_5 = \{4.39 \times 0.1 + [(1-0.0495) \times 2 + 0.0495 \times 4.39] \times 0.0495\}/0.251 + 2.11 = 4.28$

Step 14 – SCAs for each section are:

Table 3.18: Specific Collection Area

Section	SCA (s/m)
1	$SCA_1 = -(\varsigma/\mathring{a}_0) \times (1 - S_N) \times \ln(p_c) / (E_{avg}^2 \times MMD_1 \times 10^{-6}) = 8.86$
2	$SCA_2 = SCA_1 \times MMD_1/MMD_2 = 8.86\ (7 / 5.34) = 11.61$
3	$SCA_3 = SCA_2 \times MMD_2/MMD_3 = 11.61\ (5.34 / 4.67) = 13.28$
4	$SCA_4 = SCA_3 \times MMD_3/MMD_4 = 13.28\ (4.67 / 4.39) = 14.13$
5	$SCA_5 = SCA_4 \times MMD_4/MMD_5 = 14.13\ (4.39 / 4.28) = 14.49$

Step 15 – Calculate the total *SCA*.

$$\text{Total SCA} = 19.65 + 25.76 + 29.46 + 31.34 + 32.15 = 138.36\ \frac{s}{m}$$

$$\text{English SCA} = 5.080 \times 138.36 = 702.87\ \frac{ft^2}{kacfm}$$

Note that the more rigorous procedure calls for an *SCA* that is considerably higher than the value found by using Equation 3.23. This discrepancy is caused by the considerably smaller particle size used in the example problem than is assumed for Table 3.3. In this case, the shorter method would lead to an unacceptably low cost estimate.

Total collector plate area is:

$$702.87\ \frac{ft^2}{kacfm} \times 50\ kacfm = 35{,}144\ ft^2$$

3.4.5.2 ESP Cost

From Figure 3.5, the basic flange-to-flange cost of the rigid electrode ESP is $438,060 (mid-1987 dollars). Assuming all standard options are purchased. The ESP cost rises to $635,189 (mid-1987 dollars).

3.4.5.3 Costs of Auxiliaries

Assume the following auxiliary costs have been estimated from data in other parts of the Manual:

Table 3.19: Auxiliary Equipment Costs

Ductwork	$16,000
Fan	16,000
Motor	7,500
Starter	4,000
Dampers	7,200
Pneumatic conveyer	4,000
Stack	8,000
Total	$62,700

3.4.5.4 Total Capital Investment

Direct costs for the ESP system, based on the factors in Table 3.16, are given in Table 3.20. (Again, we assume site preparation and building costs to be negligible.) TCI is $1,840,000 (rounded, mid-1987 dollars).

3.4.5.5 Annual Costs-Pressure Drop

Table 3.21 gives the direct and indirect annual costs, as calculated from the factors given in Section 3.4. Pressure drop (for energy costs) can be taken from Table 3.11 in Subsection 3.2.2. Using the higher values from the table, pressure drop for the inlet diffuser plate, inlet and outlet transitions, baffles, and plates is:

$$\Delta P = 0.09 + 0.14 + 0.015 + 0.123 + 0.008 = 0.38 \text{ in. } H_2O$$

Assume the ductwork contributes an additional 4.1 in. H_2O.[9] The total pressure drop is, therefore, 4.48 in. H_2O. As is typical, the ductwork pressure drop overwhelms the ESP pressure drop.

---

[9] For ductwork pressure drop data, refer to Section 2.1 ("Hoods, Ductwork, and Stacks") of the *Manual*.

Table 3.20: Capital Cost Factors for ESP System Example Problem

Cost Item	Cost, $
Direct Costs	
Purchased equipment costs	
Adsorber vessels and carbon	$635,189
Auxiliary equipment	---
Sum = A	$635,189
Instrumentation, 0.1 A	69,789
Sales taxes, 0.03 A	20,937
Freight, 0.05 A	34,894
Purchased equipment cost, B	$823,509
Direct installation costs	
Foundations & supports, 0.04 B	32,940
Handling & erection, 0.50 B	411,755
Electrical, 0.08 B	65,881
Piping, 0.01 B	8,235
Insulation for ductwork, 0.02 B	16,470
Painting, 0.02 B	16,470
Direct installation costs	$551,751
Site preparation	---
Facilities and Buildings	---
Total Direct Costs, DC	$1,375,260
Indirect Costs (installation)	
Engineering, 0.20 B	164,702
Construction and field expenses, 0.20 B	164,702
Contractor fees, 0.10 B	82,351
Start-up, 0.01 B	8,235
Performance test, 0.01 B	8,235
Model study, 0.02 B	16,470
Contingencies, 0.03 B	24,705
Total Indirect Costs, IC	$469,400
Total Capital Investment (rounded)	$1,840,000

3.4.5.6 Total Annual Cost

The total annual cost, calculated in Table 3.21, is $511,000 (rounded). Had the particle sizes being captured been larger, the ESP cost would have been considerably less. Also, for a much larger gas flow rate, the $/acfm treated cost would have been more favorable. Reviewing components of the TAC, dust disposal is the largest single item. Care should be taken in determining this cost and the unit disposal cost ($/ton). Finding a market for the dust, for example, as an extender in asphalt or a dressing for fields, even at giveaway prices, would reduce TAC dramatically.

Table 3.21: Annual Costs for Carbon Absorber System Example Problem

Cost Item	Calculations	Cost
Direct Annual Costs, DC		
Operating Labor		
Operator	3 hr/day x 360 days/yr x $12/hr	$12,960
Supervisor	15% of operator = 0.15 x 7,820	1,944
Coordinator	1/3 of operator = 1/3 x 12,960	4,320
Operating materials		—
Maintenance		
Labor	$4,125 for collector area < 50,000 ft^3	4,125
Material	1% of purchase equipment costs = 0.01 x 823,509	8,235
Utilities		
Electricity-fan	0.000181 x 50,000 acfm x 4.48 in. H_2O x 8,640 hr/yr x $0.06 kWh	21,018
Electricity-operating	1.94×10^{-3} x 35,144 ft^2 x 8,640 hr x $0.06/kWh	35,344
Waste Dispoal	at $20/ton tipping fee at 2 miles and $0.50/ton-mile for essentially 100% collection efficiency: 4.29 x 10^{-6} x 4 gr/ft^3 x 8,640 hr/yr x 50,000 acfm x (20 + 0.50 x 2) $/ton	155,676
	Total DC	$243,622
Indirect Annual Costs, IC		
Overhead	60% of sum of operating labor, maintenance labor, & maintenance materials : = 0.6(12,960 + 1,944 + 4,320 + 4,125 + 8,235)	18,950
Administrative charges	2% of Total Capital Investment = 0.02($1,844,660)	36,893
Property tax	1% of Total Capital Investment = 0.01($1,844,660)	18,447
Insurance	1% of Total Capital Investment = 0.01($1,844,660)	18,447
Capital recovery[a]	0.1175 ($1,844,660)	216,748
	Total IC	$309,485
Total Annual Cost (rounded)		$553,000

[a] The capital recovery cost factor, CRF, is a function of the fabric filter or equipment life and the opportunity cost of the capital (i.e., interest rate). For this example, for a 20 year equipment life and a 10% interest rate, CRF = 0.1175.

3.5 Acknowledgments

We gratefully acknowledge C. G. Noll, United McGill Corp. (Columbus, OH), for extensive review and the following companies for contributing data to this chapter:

- Research-Cottrell
- Joy Industrial Equipment Co., Western Precipitation Division (Los Angeles, CA)
- Environmental Elements Corp. (Baltimore, MD)

References

[1] White, H. J., *Industrial Electrostatic Precipitation*, Addison-Wesley, Reading, MA, 1963.

[2] Lawless, P. A., and L. E. Sparks, "A Review of Mathematical Models for ESPs and Comparison of Their Successes," *Proceedings of Second International Conference on Electrostatic Precipitation*, S. Masuda, ed., Kyoto, 1984, pp. 513-522.

[3] Bump, R. L. (Research Cottrell, Inc.), "Evolution and Design of Electrostatic Precipitator Discharge Electrodes," paper presented at the APCA Annual Meeting, New Orleans, LA, June 1982.

[4] Correspondence: Richard Selznick (Baron Blakeslee, Inc., Westfield, NJ) to William M. Vatavuk, April 23, 1986.

[5] Correspondence: James Jessup (M&W Industries, Inc., Rural Hall, NC) to William M. Vatavuk, May 16, 1986.

[6] Matley, Jay (ed.), *Modern Cost Engineering*, McGraw-Hill, New York, 1984, p. 142.

[7] Correspondence: Joseph Bruno (AirPol, Inc.) to SAIC, August 26, 1999.

[8] Correspondence: Sam Dunkle (Fisher-Klosterman, Inc.) to SAIC, August 26, 1999.

[9] Correspondence: James Reynolds (Croll-Reynolds Clean Air Tecnologies) to SAIC, September 1, 1999.

[10] Personal communication: Robert Shipe, Jr. (American Air Filter Co., Louisville, KY), and S. A. Sauerland (United Air Specialists, Inc., Cincinnati, OH), to Roger Ellefson (JACA Corp., Fort Washington, PA), June 1987.

[11] R.S. Means Company, Inc., *Means Square Foot Costs 1987*, Kingston, MA.

[12] Telecon: Gary Greiner (ETS, Inc., Roanoke, VA) to James H. Turner, October 1986.

[13] PEDCo Environmental, Inc., *Operating and Maintenance Manual for ESPs*, Publication No. EPA/625/1-85/017, Office of Research and Development, Air and Energy Engineering Research Lab, Research Triangle Park, NC, September 1985.

[14] Perry,, R. H., *et al.*, *Perry's Chemical Engineers Handbook* (Sixth Edition), McGraw-Hill, New York, 1984.

[15] Bakke, E., "Wet Electrostatic Precipitators for Control of Submicron Particles," *Proceedings of the Symposium on Electrostatic Precipitators for the Control of Fine Particles*, Pensacola, FL, September 30 to October 2, 1974, Publication No. EPA 650/ 2-75-016, 1975.

[16] Beltran Associates, Inc., "Poly-Stage Precipitator for Stack and Duct Emissions," November 1978.

[17] Vatavuk, W.M., and R.B. Neveril, "Estimating Costs of Air-Pollution Control Systems, Part XVII: Particle Emissions Control," *Chemical Engineering* (adapted), April 2, 1984, pp. 97-99.

[18] Gumerman, R.C., B.E. Burris, and S.P. Hansen, *Estimation of Small System Water Treatment Costs*, Publication No. EPA/600/2-84/184a, NTIS No. PB85-161644, 1984.

[19] Gooch, J.P., *A Manual on the Use of Flue Gas Conditioning for ESP Performance Enhancement*, Electric Power Research Institute Report No. CS-4145, 1985.

Appendix A

Effects of Material Thickness and Type on ESP Costs

The impact of material thickness and composition of collecting plates and the ESP casing can be estimated using the following equations and Figure 6.17:

Plates:

$$I = \frac{\left[\left(\frac{W_t}{2} \times FS\right) - 0.90\right] M + SP}{SP} \tag{3.54}$$

Casing:

$$I = \frac{\left[\left(\frac{W_t}{10} \times FS\right) - 0.58\right] M + SP}{SP} \tag{3.55}$$

where

I	=	incremental increase of flange-to-flange selling price ($/ft^2)
W_t	=	weight of steel (lb/ft^2)
FS	=	fabricated steel selling price ($/lb) (normally assume approximately 2 times material cost)
M	=	manufacturer's markup factor of fabricated cost (direct labor, wages, and material cost before general and administrative expense and profit) to selling price (normally 2 to 3)
SP	=	flange-to-flange selling price from Figure 6.17 ($/ft^2)

Most vendors can produce ESPs with collecting plate material thicknesses from 16 to 20 gauge and casing material thicknesses from 1/8 through 1/4 in. without affecting the 2 times material cost = fabricated cost relationship. Thus, the impact of increasing the collecting plates from 18 to 16 gauge and the casing from 3/16 to 1/4 in. plate on a 72,000 ft^2 collector having a selling price of $10/ft^2 and assuming a markup factor of 2 is as follows:

Plates:

$$I = \frac{\left[\left(\frac{2.5}{2} \times 0.90\right) - 0.90\right] 2 + 10}{10} \qquad (3.55)$$
$$= 1.045 = 4.5 \text{ percent increase}$$

Casing:

$$I = \frac{\left[\left(\frac{10.21}{10} \times 0.76\right) - 0.58\right] 2 + 10}{10} \qquad (3.56)$$
$$= 1.039 = 3.9 \text{ percent increase}$$

Equations 3.53 and 3.54 were developed using the following assumptions:

$$I = \frac{\text{Material selling price increase} + \text{Standard ESP selling price}}{\text{Standard ESP selling price}}$$

Because Figure 3.5 identifies the standard ESP selling price /ft^2 of collecting area, the material selling price increase = (New material cost - Standard material cost)M. Then it follows that:

$$\text{Material selling price} = \frac{\text{lb steel}}{\text{ft}^2 \text{ collecting area}} \times \text{Fabricated cost in } \frac{\$}{\text{lb}} \times M \qquad (3.57)$$

The ESP dimensions given in Figure 3.7 include:

- Casing area = 30 ft 30 ft x 8 = 7,200 ft^2 (assume 4 walls, 1 top, 2 hopper sides, 2 triangular hopper ends π 8 equivalent sides)
- Collecting plate area =

$$30 \text{ ft} \times 30 \text{ ft} \times 2 \frac{\text{sides}}{\text{plate}} \times \frac{30 \text{ ft}}{s} \text{ plates} = \frac{54{,}000}{s} \text{ft}^2 = 72{,}000 \text{ ft}^2 \text{ for } s = 0.75 \text{ ft} \qquad (3.59)$$

where s = plate spacing (ft)

Thus, there are:

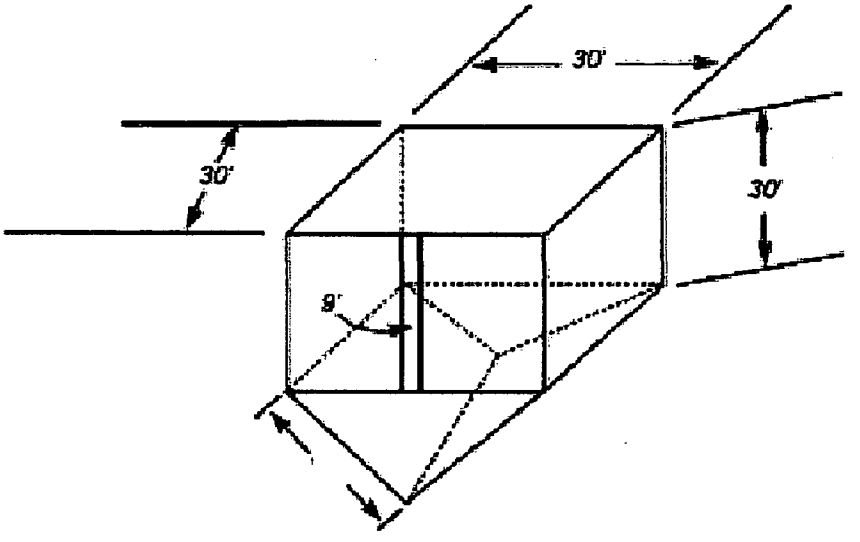


Figure 6.19:ESP Dimensions

- 7.50/s ft^2 of collecting area per 1 ft^2 of casing and
- 2 ft^2 of collecting area per 1 ft^2 of collecting plate

Material cost per ft^2 collecting area is:

$$\text{Plates} = \frac{\text{lb steel}}{2\ \text{ft}^2} \times \frac{\$}{\text{lb}} \tag{3.60}$$

$$\text{Casing} = \frac{\frac{\text{lb steel}}{\text{ft}^2}}{\frac{7.50}{s}} \times \frac{\$}{\text{lb}} \tag{3.61}$$

For a standard ESP with 18 gauge collecting plates and 3/16 in. plate casing. Assuming:

Material cost for 18 gauge mild steel	=	\$0.45/lb
Material cost for 3/16 in. plate mild steel	=	\$0.38/lb
Material cost to fabricated cost factor	=	2

These costs yield fabricated material costs of:

Plates:

$$\frac{2 \text{ lb}}{2 \text{ ft}^2} \times \frac{\$0.45}{\text{lb}} \times 2 = \frac{\$0.90}{\text{ft}^2} \text{ of collecting area} \tag{3.62}$$

Casing:

$$\frac{\frac{7.66 \text{ lb}}{\text{ft}^2}}{\frac{7.50}{s}} \times \frac{\$0.38}{\text{lb}} \times 2 = \$0.78 \frac{s}{\text{ft}^2} \text{ of collecting area} \tag{3.63}$$

At a typical 9 in. plate spacing the casing cost would be $\$0.58/\text{ft}^2$ of collecting area.

which gives us Equations 3.53 and 3.54. Note that the value 0.58 will change significantly if a plate spacing other than 9 in. is chosen.

$$\text{Selling price impact} = \frac{\left(\text{Cost of new material} - \text{Cost of old material}\right) M + \text{Original overall selling price}}{\text{Original overall selling price}} \tag{3.64}$$

Thus, for a less than 5 percent increase in flange-to-flange cost, all the precipitator exposed wall sections can be increased by more than 25 percent to provide increased life under corrosive conditions. Section thickness increases that are greater than those just discussed would probably result in significant cost increases because of both increased material costs and necessary engineering design changes.

The impact of changing from mild steel to 304 stainless steel assuming material costs of \$1.63/lb for 18 gauge collecting plates, \$1.38/lb for the 3/16 in. casing, and a markup factor of 3 is as follows:

Plates:

$$I = \frac{\left[\left(\frac{2}{2} \times 1.63\right) - 0.9\right] 3 + 10}{10} \tag{3.65}$$
$$= 21.9 \text{ percent increase}$$

Casing:

$$I = \frac{\left[\left(\frac{7.66}{10} \times 1.38\right) - 0.58\right] 3 + 10}{10} \qquad (3.66)$$

$$= 14.3 \text{ percent increase}$$

To these material costs must be added extra fabrication labor and procurement costs that will increase the ESP flange-to-flange cost by a factor of 2 to 3. Note that a totally stainless steel collector would be much more expensive because the discharge electrodes, rappers, hangers, etc., would be also converted to stainless. The preceding equations can be used for other grades of stainless steel or other materials of construction by inserting material costs obtained from local vendors on a $/lb basis.

TECHNICAL REPORT DATA

(Please read Instructions on reverse before completing)

1. REPORT NO.	2.	3. RECIPIENT'S ACCESSION NO.
452/B-02-001		

4. TITLE AND SUBTITLE	5. REPORT DATE
The EPA Air Pollution Control Cost Manual	January, 2002
	6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S)	8. PERFORMING ORGANIZATION REPORT NO.
Daniel Charles Mussatti	
9. PERFORMING ORGANIZATION NAME AND ADDRESS	10. PROGRAM ELEMENT NO.
U.S. Environmental Protection Agency Office of Air Quality Planning and Standards Air Quality Standards and Strategies Division Innovative Strategies and Economics Group Research Triangle Park, NC 27711	11. CONTRACT/GRANT NO.
12. SPONSORING AGENCY NAME AND ADDRESS	13. TYPE OF REPORT AND PERIOD COVERED
Director Office of Air Quality Planning and Standards Office of Air and Radiation U.S. Environmental Protection Agency Research Triangle Park, NC 27711	Final
	14. SPONSORING AGENCY CODE EPA/200/04

15. SUPPLEMENTARY NOTES

Updates and revises EPA 453/b-96-001, OAQPS Control Cost Manual, fifth edition (in English only)

16. ABSTRACT

In Spanish, this document provides a detailed methodology for the proper sizing and costing of numerous air pollution control devices for planning and permitting purposes. Includes costing for volatile organic compounds (VOCs); particulate matter (PM); oxides of nitrogen (NOx); SO2, SO3, and other acid gasses; and hazardous air pollutants (HAPs).

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Economics Cost Engineering cost Sizing Estimation Design	Air Pollution control Incinerators Absorbers Adsorbers Filters Condensers Electrostatic Precipitators Scrubbers	

18. DISTRIBUTION STATEMENT	19. SECURITY CLASS *(Report)*	21. NO. OF PAGES
Release Unlimited	Unclassified	1,400
	20. SECURITY CLASS *(Page)*	22. PRICE
	Unclassified	

EPA Form 2220-1 (Rev. 4-77) PREVIOUS EDITION IS OBSOLETE