



Draft EIS

Volume 1B: Control Technologies

NESEA

EPA-453/D-92-016b

**Hazardous Air Pollutant Emissions
from Process Units in the
Synthetic Organic Chemical
Manufacturing Industry--
Background Information
for Proposed Standards**

Volume 1B: Control Technologies

Emission Standards Division

U.S. Environmental Protection Agency
Region 5, Library (PL-121)
77 West Jackson Boulevard, 12th Floor
Chicago, IL 60604-3590

**U.S. Environmental Protection Agency
Office of Air and Radiation
Office of Air Quality Planning and Standards
Research Triangle Park, North Carolina 27711**

November 1992

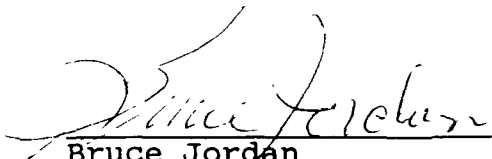
(DISCLAIMER)

This Report has been reviewed by the Emission Standards Division of the Office of Air Quality Planning and Standards, EPA, and approved for publication. 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 Library Services Office (MD-35), U.S. Environmental Protection Agency, Research Triangle Park, N.C. 27711, or from the National Technical Information Service, 5285 Port Royal Road, Springfield, VA 22161.

ENVIRONMENTAL PROTECTION AGENCY

Background Information for Proposed Standards
Hazardous Air Pollutant Emissions from Process Units
in the Synthetic Organic Chemical Manufacturing Industry
Volume 1B: Control Technologies

Prepared by:


Bruce Jordan
Director, Emission Standards Division
U.S. Environmental Protection Agency
Research Triangle Park, N.C. 27711

Nov 30, 1992
(Date)

1. The proposed standards would regulate emissions of organic hazardous air pollutants (HAP's) emitted from chemical manufacturing processes of the Synthetic Organic Chemical Manufacturing Industry (SOCMI). Only those chemical manufacturing processes that are part of major sources under Section 112(d) of the CAA would be regulated. The recommended standards would reduce emissions of 149 of the organic chemicals identified in the CAA list of 189 hazardous air pollutants.
2. Copies of this document have been sent to the following Federal Departments: Labor, Health and Human Services, Defense, Office of Management and Budget, Transportation, Agriculture, Commerce, Interior, and Energy; the National Science Foundation; and the Council on Environmental Quality. Copies have also been sent to members of the State and Territorial Air Pollution Program Administrators; the Association of Local Air Pollution Control Officials; EPA Regional Administrators; and other interested parties.
3. The comment period for this document is 90 days from the date of publication of the proposed standard in the Federal Register. Ms. Julia Stevens may be contacted at 919-541-5578 regarding the date of the comment period.
4. For additional information contact:

Dr. Janet Meyer
Standards Development Branch (MD-13)
U.S. Environmental Protection Agency
Research Triangle Park, N.C. 27711
Telephone: 919-541-5299
5. Copies of this document may be obtained from:

U.S. EPA Library (MD-35)
Research Triangle Park, N.C. 27711

National Technical Information Service (NTIS)
5285 Port Royal Road
Springfield, VA 22161
Telephone: 703-487-4650

OVERVIEW

Emission standards under Section 112(d) of the Clean Air Act apply to new and existing sources in each listed category of hazardous air pollutant emission sources. This background information document (BID) provides technical information used in the development of the Hazardous Organic National Emission Standard for Hazardous Air Pollutants (NESHAP), which will affect the Synthetic Organic Chemical Manufacturing Industry (SOCMI). The BID consists of three volumes: Volume 1A, National Impacts Assessment (EPA-453/D-92-016a); Volume 1B, Control Technologies (EPA-453/D-92-016b); and Volume 1C, Model Emission Sources (EPA-453/D-92-016c).

Volume 1A presents a description of the affected industry and the five kinds of emission points included in the impacts analysis: process vents, transfer loading operations, equipment leaks, storage tanks, and wastewater collection and treatment operations. Volume 1A also describes the methodology for estimating nationwide emissions, emission reductions, control costs, other environmental impacts, and increases in energy usage resulting from a potential NESHAP; and presents three illustrative sets of potential national impacts and a summary of the economic analysis. While Volume 1A provides the overview of how information on model emission sources and control technology cost were used to estimate national impacts, Volumes 1B and 1C contain detailed information on the estimation of control technology performance and costs and model emission source development.

Volume 1B discusses the applicability, performance, and costs of combustion devices; collection systems and recovery devices; storage tank improvements; and control techniques for equipment leak emissions. These control technologies were the basis of the Hazardous Organic NESHAP impacts analysis. These control technologies are applicable to emission points in the SOCMI and in other source categories. Methods for estimating

capital costs and annual costs (including operation and maintenance costs) of each control technology are presented.

Volume 1C presents descriptions of each kind of emission point included in the impacts analysis and the development of model emission sources to represent each kind of emission point for use in the impacts analysis. The emission reductions, other environmental impacts, and energy impacts associated with application of the control technologies described in Volume 1B to the model emission sources is discussed. For illustrative purposes, the environmental, energy, and cost impacts that would results from control of several example model emission sources are presented.

TABLE OF CONTENTS

<u>Section</u>	<u>Page</u>
LIST OF TABLES	ix
LIST OF FIGURES	xi
ACRONYM AND ABBREVIATION LIST	xii
1.0 INTRODUCTION	1-1
2.0 EMISSION CONTROL TECHNOLOGIES	2-1
2.1 Combustion Control Devices	2-1
2.1.1 Flares	2-1
2.1.2 Thermal Incinerators	2-7
2.1.3 Catalytic Incinerators	2-13
2.1.4 Industrial Boilers and Process Heaters	2-16
2.2 Collection Systems and Recovery Devices	2-20
2.2.1 Vapor Collection Systems for Loading Racks	2-21
2.2.2 Condensers	2-27
2.2.3 Steam Strippers	2-33
2.2.4 Carbon Adsorbers	2-43
2.2.5 Absorbers	2-48
2.3 Storage Tank Improvements for Emission Reduction	2-52
2.3.1 Description of Tank Improvements	2-53
2.3.2 Factors Affecting Control Efficiency	2-54
2.3.3 Applicability of Storage Tank Improvements	2-62
2.4 Equipment Leak Emission Sources and Emission Control Techniques	2-63
2.4.1 Equipment Description and Specifications	2-65
2.4.2 Closed Vent Systems and Control Devices	2-77
2.4.3 Work Practices	2-78
2.5 References	2-84
3.0 COST ANALYSIS	3-1
3.1 Cost Methodology for Combustion Systems	3-1

TABLE OF CONTENTS
(Continued)

<u>Section</u>	<u>Page</u>
3.1.1 Cost Methodology for Flare Systems . .	3-1
3.1.2 Cost Methodology for Incinerator Systems	3-7
3.2 Cost Methodology for Collection Systems and Recovery Devices	3-21
3.2.1 Cost Methodology for Vapor Collection Systems for Loading Racks	3-21
3.2.2 Cost Methodology for Condensers . . .	3-24
3.2.3 Cost Methodology for Steam Stripping .	3-31
3.3 Cost Methodology for Storage Tank Improvements	3-40
3.3.1 Design Considerations Affecting Cost .	3-40
3.3.2 Development of Capital Costs	3-43
3.3.3 Development of Total Annual Costs . .	3-44
3.4 Cost Methodology for Equipment Leak Control Technologies	3-45
3.4.1 Control Equipment	3-45
3.4.2 Leak Detection and Repair Techniques .	3-50
3.4.3 Capital Costs	3-57
3.4.4 Annual Costs	3-57
3.5 References	3-61

LIST OF TABLES

<u>Table</u>		<u>Page</u>
2-1	Controlled and Uncontrolled Internal Floating Roof Deck Fittings	2-55
2-2	Effectiveness of Internal Floating Roofs on an Example Tank	2-57
2-3	Internal Floating Roof Rim Seal Systems, Seal Loss Factors, and Control Efficiencies .	2-60
3-1	Flare General Design Specifications	3-2
3-2	Bases and Factors for Annual Costs for Flares	3-6
3-3	Incinerator General Design Specifications . .	3-8
3-4	Scrubber General Design Specifications . . .	3-12
3-5	Capital Cost Factors for Thermal Incinerators	3-18
3-6	Bases and Factors for Annual Costs for Thermal Incinerators	3-20
3-7	Total Annual Cost for Loading Rack Vapor Collection Systems	3-25
3-8	Equipment Cost Equations for Packaged Refrigerated Condenser Systems	3-27
3-9	Bases and Factors for Annual Costs for Refrigerated Condenser Systems	3-29
3-10	Equipment Cost Equations for a Steam Stripping Unit	3-33
3-11	Cost Methodology for Estimating Total Capital Investment for a Steam Stripping System . . .	3-36
3-12	Cost Methodology for Estimation of Total Annual Cost for a Steam Stripping System . .	3-41
3-13	Base Cost Data for Equipment Leak Control Devices	3-47
3-14	Base Costs and Assumptions for a Leak Detection Program	3-54

LIST OF TABLES
(Continued)

<u>Table</u>		<u>Page</u>
3-15	Equations for Determining Costs and Number of Leaks for a Leak Detection Program	3-55
3-16	Derivation of Annualized Costs for Control of Equipment Leaks	3-58

LIST OF FIGURES

<u>Figure</u>		<u>Page</u>
2-1	Steam-Assisted Elevated Flare System	2-3
2-2	Discrete Burner, Thermal Incinerator	2-9
2-3	Distributed Burner, Thermal Incinerator	2-11
2-4	Catalytic Incinerator	2-15
2-5	Vapor Balancing System	2-24
2-6	Refrigerated Condenser System for VOC Vapor Recovery	2-28
2-7	Schematic Diagram of a Shell and Tube Surface Condenser	2-30
2-8	Continuous Steam Stripper System	2-36
2-9	Two-Stage Regenerative Adsorption System	2-45
2-10	Packed Tower Absorption Process	2-50
3-1	Summary of Total Capital Investment versus Wastewater Feed Rate for Steam Stripping Unit	3-37
3-2	Unit Operating Costs Versus Wastewater Feed Rate for Steam Stripping Unit	3-42

ACRONYM AND ABBREVIATION LIST

API	American Petroleum Insititute
Btu	British thermal unit(s)
cfm	cubic foot (feet) per minute
cm	centimeter
CO ₂	carbon dioxide
EPA	U.S. Environmental Protection Agency
fpm	foot (feet) per minute
ft	foot (feet)
gal	gallon(s)
gm	gram(s)
gm-mole	gram-mole(s)
H ₂	hydrogen
HAP	hazardous air pollutant
HON	hazardous organic NESHAP
hp	horsepower
hr	hour(s)
in.	inch(es)
kg	kilogram(s)
kPa	kilopascal
ℓ	liter(s)
lb	pound(s)
lb-mole	pound mole(s)
LEL	lower explosive limit
MACT	maximum achievable control technology
Mg	megagram(s)
mm	millimeter(s)
MMBtu	million British thermal unit(s)
MJ	megajoule(s)
MW	megawatt(s)
N ₂	nitrogen
NESHAP	National Emission Standards for Hazardous Air Pollutants

NPDES	National Pollutant Discharge Elimination System
NSPS	New Source Performance Standard(s)
OAQPS	Office of Air Quality Planning and Standards
OCCM	OAQPS Control Cost Manual
OSHA	Occupational Safety and Health Administration
POTW	publicly owned treatment works
ppmv	part(s) per million by volume
psia	pound(s) per square inch absolute
RACT	reasonably available control technology
RD/PRV	rupture disk/pressure relief valve
scf	standard cubic foot (feet)
scfm	standard cubic foot (feet) per minute
scm	standard cubic meter
sec	second(s)
SOCMI	synthetic organic chemical manufacturing industry
TLV	threshold limit value
VOC	volatile organic compound
VOL	volatile organic liquid
μm	micrometer(s)
W	watt(s)
W/m ²	watts per square meter
yr	year(s)
\$	dollar(s)

1.0 INTRODUCTION

The hazardous organic national emission standards for hazardous air pollutants (NESHAP) will address the following five sources of emissions:

- Equipment leaks;
- Transfer operations;
- Process vents;
- Storage vessels (tanks); and
- Wastewater treatment operations.

This volume of the background information document (BID) for the hazardous organic NESHAP (HON) presents the technologies considered for control of organic hazardous air pollutant (HAP) emissions from these sources.

The control technologies described here are common in the synthetic organic chemical manufacturing industry (SOCMI) and have been studied by the U.S. Environmental Protection Agency (EPA) during past rulemaking. The discussions of efficiency, applicability, and costs presented in this volume are a compilation of information gathered during those earlier studies.

The emission sources affected by the HON have the potential to emit organic HAP's, which are part of the larger criteria pollutant class of volatile organic compounds (VOC's). The control technologies discussed in this volume reduce emissions of VOC's and therefore organic HAP's. Because data on the performance of the control technologies are based on reduction of VOC's, these discussions present reductions in terms of VOC's. Similar emission reduction efficiencies would be expected for the smaller group of organic HAP's.

For each of the control technologies discussed in Chapter 2, the design and operation of the control device or system is described, including an explanation of the physical or chemical processes that destroy the VOC or remove it from the emission stream. The second part of each discussion describes the factors affecting the efficiency of the control device, including vent stream characteristics and control device operating parameters. The last part of each discussion addresses the applicability of the technologies; for example, to what types of SOCM sources can the technology be successfully applied?

Most of the control technologies discussed in Chapter 2 may be applied to many different types of emission sources. For this reason, the discussions of the technologies were intended to be independent of emission source. However, because vapor collection systems for loading racks, storage tank improvements, and equipment leak controls are specific to an emission source, the discussions of these technologies also include a description of the emission source.

Chapter 2 contains descriptions of the following technologies:

- Combustion devices (flares, thermal and catalytic incinerators, and industrial boilers and process heaters);
- Collection systems and recovery devices (vapor collection systems for loading racks, condensers, steam strippers, carbon adsorbers, and absorbers);
- Storage tank improvements; and
- Control techniques for equipment leak emissions.

For combustion devices and collection and recovery systems, descriptions of the technologies include a discussion of the factors affecting their efficiency and their applicability to different waste streams.

For storage tanks, the discussion focuses on installing a floating roof and upgrading an existing floating roof. The types and mechanisms of different VOC losses are described with the controls applicable to the particular loss types.

The discussion of equipment leak emissions describes the various sources of equipment leak emissions, that is, the mechanisms by which different components become emitters of VOC's. Alternative (leakless) components, closed vent systems, and leak detection and repair (LDAR) programs are discussed as methods for reducing equipment leak emissions.

Chapter 3 explains the procedures for calculating the cost of applying these different emission reduction technologies. The cost methodology includes discussion of design considerations affecting costs, development of capital costs, and development of annual costs. Of the combustion technologies, flares and incinerator systems (including an acid gas scrubber) are addressed. Cost methodologies for collection systems and recovery devices are presented for loading rack vapor collection systems, condensers, and steam strippers. For storage tanks, cost methodologies are described for installing a new floating roof and upgrading an existing floating roof. Finally, for equipment leaks, cost methodologies are included for the control equipment applicable to certain source types and for LDAR programs.

The appendices to this volume contain example cost calculations for applying control technologies to the five emission sources covered by the HON.

2.0 EMISSION CONTROL TECHNOLOGIES

This chapter discusses technologies for control of organic HAP's from sources in the SOCMI. Because organic HAP's are VOC's, technologies that have been used to control VOC emissions were evaluated for control of organic HAP's. These technologies include combustion devices, collection and recovery devices, improvements to reduce emissions from storage tanks, and equipment and work practices that reduce emissions from equipment leaks.

2.1 COMBUSTION CONTROL DEVICES

This section discusses devices that control emissions of VOC's by means of combustion. Combustion control devices, unlike noncombustion control devices, alter the chemical structure of the VOC. Destruction of the VOC through combustion is complete if all VOC's are converted to carbon dioxide and water. Incomplete combustion results in some of the VOC remaining unaltered or being converted to other organic compounds such as aldehydes or acids.

The combustion control devices discussed in the following four subsections include flares, thermal incinerators, catalytic incinerators, and boilers and process heaters. The discussion for each device treats its typical design and operation, factors affecting destruction efficiency, and applicability.

2.1.1 Flares

2.1.1.1 Description of Flares. Flaring is an open combustion process in which the oxygen necessary for combustion is provided by the air around the flame. Good combustion in a flare is governed by flame temperature, residence time of organic species in the combustion zone, turbulent mixing of the organic species to complete the

oxidation reaction, and the amount of oxygen available for free radical formation.

Many flare systems are operated with baseload gas recovery systems to recover VOC's from the flare header system for reuse. The recovered VOC's may be used as feedstock in other processes or as a fuel in process heaters, boilers, or other combustion devices. When baseload gas recovery systems are applied, the flare is typically used to combust process upset and emergency gas releases that the baseload system is not designed to recover. The operation of a baseload gas recovery system may offer an economic advantage over operation of a flare alone if sufficient quantities of usable VOC's can be recovered.

Flares are generally categorized in two ways: (1) by the height of the flare tip (i.e., ground-level or elevated) and (2) by the method of enhancing mixing at the flare tip (i.e., steam-assisted, air-assisted, pressure-assisted, or unassisted). Elevating the flare can prevent potentially dangerous conditions at ground level where the open flame (i.e., an ignition source) is near a process unit. Further, the products of combustion can be dispersed above working areas to reduce the effects of noise, heat radiation, smoke, and objectionable odors.

This discussion focuses on steam-assisted elevated flares, the most common type used in the chemical industry. Ground flares are discussed only in how they differ from elevated flares.

The basic elements of a steam-assisted elevated flare system are shown in Figure 2-1. The vent stream is sent to the flare through the collection header (1). The vent stream entering the header can vary widely in volumetric flow rate, moisture content, VOC concentration, and heat content. The knockout drum (2) removes water or condensed hydrocarbons that can extinguish the flame or cause irregular combustion or smoking in the flare combustion zone. Vent streams are also typically routed through a flame arrestor (3) before going to the flare. This prevents possible flame flashback, which is

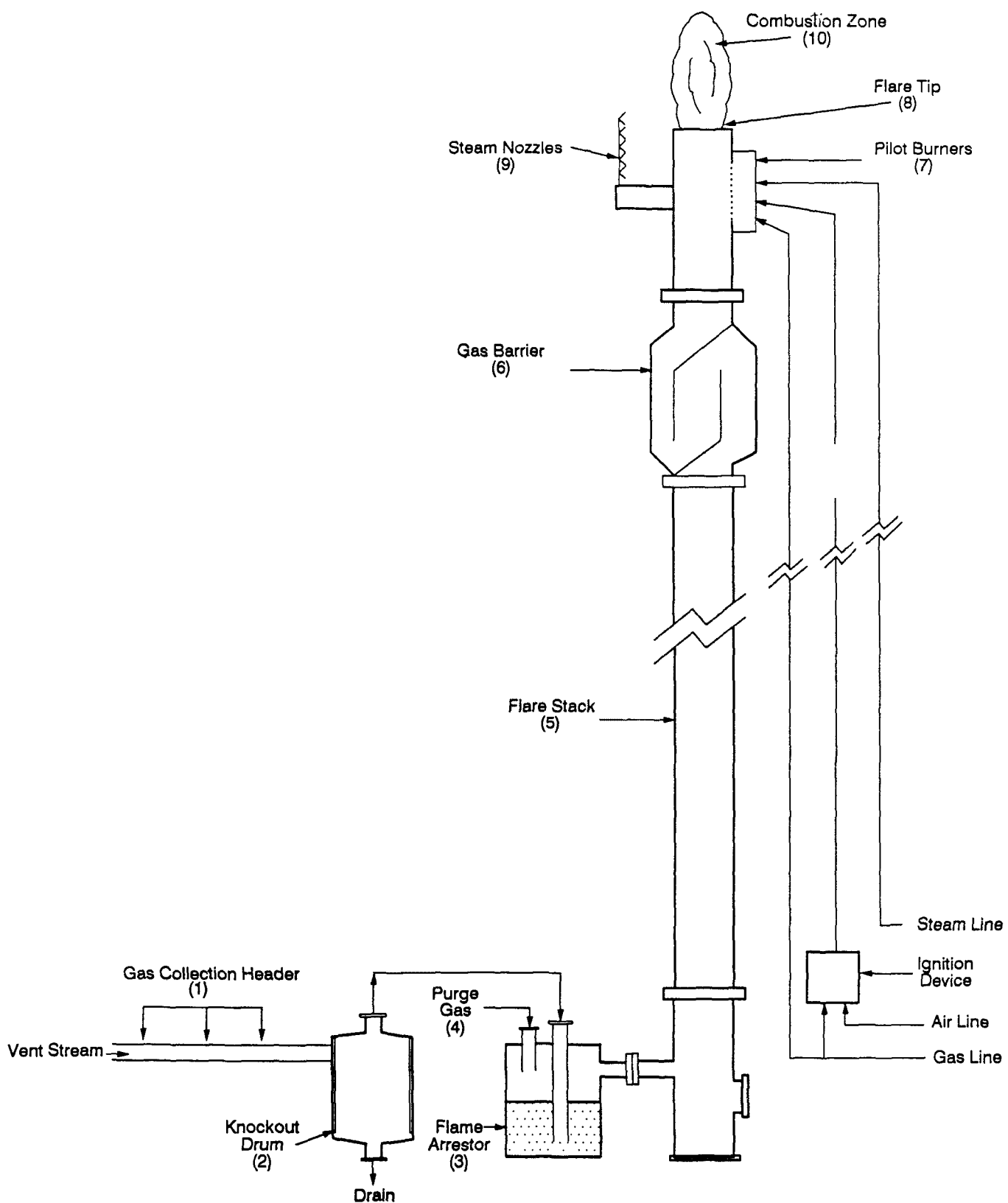


Figure 2-1. Steam-assisted elevated flare system.

caused when the vent stream flow rate to the flare is too low and the flame front pulls down into the stack.¹ Purge gas (N_2 , CO_2 , or natural gas) (4) also helps to prevent flashback in the flare stack (5) caused by low vent stream flow rate. To reduce the amount of purge gas required, a gas barrier (6) or a stack seal is typically used just below the flare head to impede the flow of air into the flare gas network.

The vent stream enters at the base of the flame where it is heated by already burning fuel and pilot burners (7) at the flare tip (8). Pilot burners positioned around the outer perimeter of the flare tip ensure reliable ignition of the vent stream. The flare tip is designed to give environmentally acceptable combustion of the vent gas over the flare system's capacity range. 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. The flare tip is designed to avoid a detached flame (a space between the stack and flame with incomplete combustion), which is caused by an excessively high flow rate.

The vent stream flows into the combustion zone (10), where the exterior of the microscopic gas pockets are oxidized. The rate of reaction is limited by the mixing of the vent stream and oxygen from the air. If the gas pocket has sufficient oxygen and residence time in the combustion zone, complete combustion (all VOC's are converted to carbon dioxide and water) will occur. 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.

A diffusion flame typical of elevated flares receives its combustion oxygen by diffusion of air into the flame from the surrounding atmosphere. The high flow rate of the vent stream into the flare requires more combustion air at a faster rate than simple gas diffusion can supply. Thus, many flares are designed with "assisted" air supply and mixing. For example, high-velocity steam injection nozzles (9) can be added to

increase gas turbulence for mixing in the flame boundary zones, thereby drawing in more combustion air and improving combustion efficiency. This steam injection promotes smokeless flare operation by minimizing the cracking reaction that forms carbonaceous soot. The significant disadvantages of using steam are increased noise and cost. The steam requirement depends on the composition of the vent stream flared, the velocity of the steam from the injection nozzle, and the diameter of the flare tip. Although some gases can be flared smokelessly without any steam, typically 0.01 to 0.6 kg of steam per kilogram of vented gas is required.

Some flares use forced air instead of steam to provide combustion air and the mixing needed for smokeless operation. These flares consist of two coaxial flow channels. The combustible gases flow in the center channel, and the combustion air (provided by a fan in the bottom of the flare stack) flows in the annulus. The principal advantage of air-assisted flares is that they can be used where steam is not available. Air assist is rarely used on large flares because air flow is difficult to control when the gas flow is intermittent. About 90.8 hp of blower capacity is required for each 100 lb/hr of gas flared.²

Ground flares are usually enclosed and have multiple burner heads that are staged to operate based on the volume of vent stream directed to the flare. The energy of the vent stream itself (because of the high-pressure drop across the nozzle) is usually adequate to provide the mixing necessary for smokeless operation, and air or steam assist is not needed. A fence or other enclosure reduces noise, heat, and light from the flare and provides some wind protection.

Ground flares have less capacity than elevated flares and are less widely used. Typically they are used to combust continuous, constant flow vent streams, whereas steam-assisted elevated flares are used to dispose of large amounts of gas released in emergencies.³ Stable combustion can be obtained with lower Btu-content vent streams than is possible with open

flare designs (50 to 60 Btu/scf has been reported¹) probably due to their isolation from wind effects.

2.1.1.2 Factors Affecting Destruction Efficiency of Flares. The destruction efficiency of flares is a function of many factors: flammability, auto-ignition temperature, and heat content of the vent stream; and mixing at the flare tip.¹

The flammability limits of the vent stream flared influence ignition stability and flame extinction. Flammability limits are 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. That is, 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. It is easier to initiate and maintain combustion of gases with wide limits of flammability (for instance, H₂).

The auto-ignition temperature of a vent stream affects combustion because gas mixtures must be at a sufficient temperature and concentration to burn. A gas with a low auto-ignition temperature will ignite more easily than a gas with a high auto-ignition temperature.

The heat content of the vent stream is a measure of the heat available from the combustion of the VOC in the vent stream. The heat content of the vent stream affects the flame structure and stability. A gas with a lower heat content produces a cooler flame that does not favor combustion kinetics and is more easily extinguished. The lower flame temperature will also reduce buoyant forces, which reduces mixing.

Poor mixing at the flare tip or poor flare maintenance can cause smoking (particulate matter release). Vent streams with high carbon-to-hydrogen ratios (greater than 0.35) have a greater tendency to smoke and require better mixing to burn smokelessly.⁴ For this reason, one generic steam-to-vent-stream ratio is not appropriate for all vent streams. The

steam required depends on the vent stream carbon-to-hydrogen ratio. A high ratio requires more steam to prevent a smoking flare.

The efficiency of a flare in reducing VOC emissions can be variable. For example, smoking flares are far less efficient than properly operated and maintained flares. Flares have been shown to have high VOC destruction efficiencies, under proper operating conditions. Ninety-eight percent combustion efficiency can be achieved by steam-assisted flares with exit flow velocities less than 18 m/sec (60 ft/sec) and combustion gases with heat contents over 11 MJ/scm (300 Btu/scf) and by flares operated without assist with exit flow velocities less than 18 m/sec (60 ft/sec) and burning gases with heat contents over 8 MJ/scm (200 Btu/scf).⁴

2.1.1.3 Applicability of Flares. Flares can be dedicated to a specific vent stream. Flares can also be designed to control both normal process releases and emergency upsets. The latter involves the release of large volumes of gases. Often, large diameter flares designed to handle emergency releases are also used to control continuous vent streams from various process operations. Typically in refineries, many vent streams are combined in a common gas header to fuel boilers and process heaters. However, excess gases, fluctuations in flow rate in the fuel gas line, and emergency releases are sometimes sent to a flare.

Flares can be used to control almost any VOC stream and can handle fluctuations in VOC concentration, flow rate, heat content, and inerts content. Flaring is appropriate for continuous, batch, and variable flow vent stream application. Some streams, such as those containing halogenated or sulfur-containing compounds, are usually not flared because they corrode the flare tip or cause formation of secondary pollutants (such as acid gases or sulfur dioxide).

2.1.2 Thermal Incinerators

2.1.2.1 Description of Thermal Incinerators. Like other combustion control devices, thermal incinerators operate on the principle that any VOC heated to a high enough temperature

in the presence of sufficient oxygen will be oxidized to carbon dioxide and water. The theoretical combustion temperature for thermal oxidation depends on the properties of the VOC to be combusted. Some VOC's are oxidized at temperatures much higher than others.

Thermal incineration processes are influenced by residence time, mixing, and combustion temperature. An efficient thermal incinerator system must provide:

- A chamber temperature high enough to enable oxidation of the organic compounds to proceed rapidly to completion;
- Sufficient turbulence for good mixing of the hot combustion products from the burner, the combustion air, and the organic compounds; and
- Sufficient residence time at the theoretical combustion temperature for oxidation to reach completion.

A thermal incinerator is usually a refractory-lined chamber containing a burner (or set of burners) at one end. As shown in Figure 2-2, discrete dual fuel burners (1) and inlets for the vent stream (2) and combustion air (3) are arranged in a premixing chamber (4) to mix the hot products from the burners thoroughly with the process vent streams. The mixture of hot combusting gases then passes into the main combustion chamber (5).⁵ This chamber is sized to allow the mixture enough time at the elevated temperature for oxidation to reach completion (residence times of 0.3 to 1.0 sec are common). Energy can be recovered from the hot flue gases in a heat recovery section (6) to make the process more energy efficient.⁶ Preheating combustion air or offgas is a common method of energy recovery; however, it is sometimes more economical to generate steam for use in the chemical process. Insurance regulations require that if the vent stream is preheated, the VOC concentration must be maintained below 25 percent of the lower explosive limit (LEL) to minimize the potential for explosion hazards.

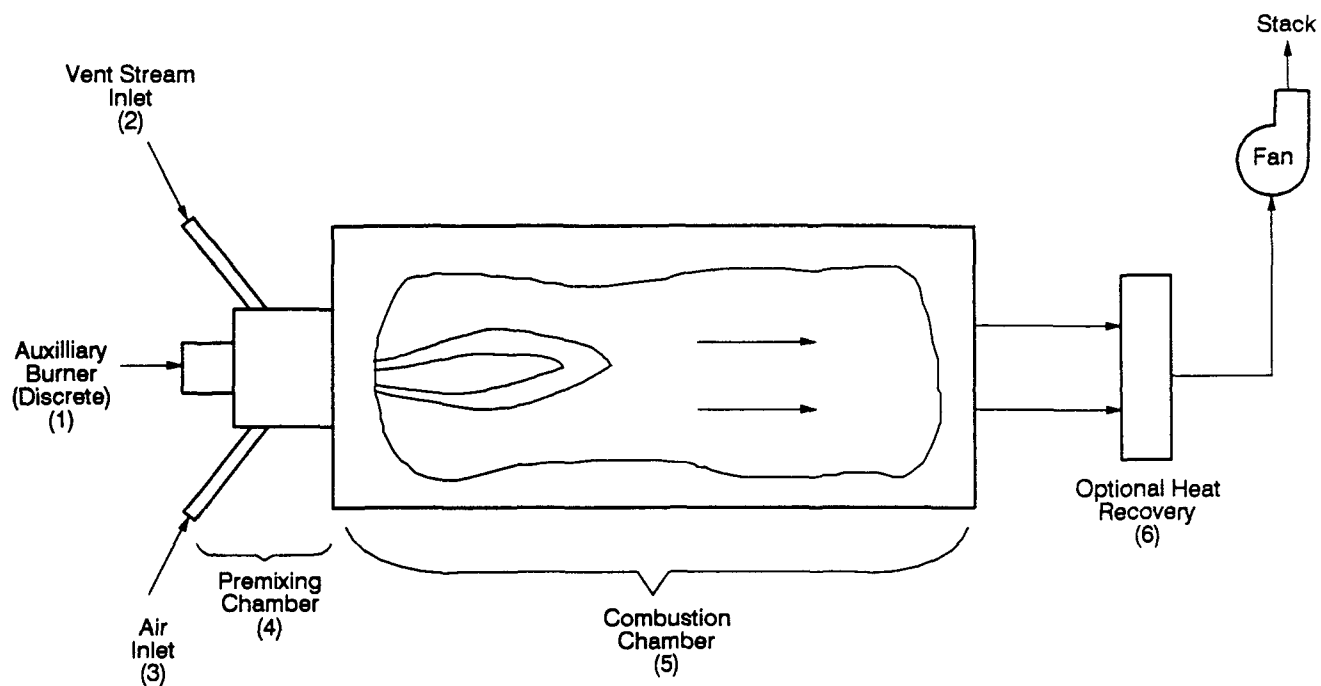


Figure 2-2. Discrete burner, thermal incinerator.

Thermal incinerators designed specifically for VOC destruction using natural gas as the supplemental fuel may also use a grid-type (distributed) gas burner (Figure 2-3). The tiny gas flame jets (1) on the grid surface (2) ignite the vapors as they pass through the grid. The grid acts as a baffle to promote mixing of the gases entering the incinerator chamber (3). This arrangement ensures that all vapors burn at a lower chamber temperature, using less fuel. This system makes possible a shorter reaction chamber, while maintaining high efficiency.

Combustion devices are always operated with some quantity of excess air to ensure a sufficient supply of oxygen. The amount of excess air used varies with the fuel and burner type but should be kept as low as possible to minimize fuel consumption while maintaining combustion efficiency. Using too much excess air wastes fuel because the additional air must be heated to the combustion chamber temperature. Excess air also increases flue gas volume and can increase the size and cost of the system. Package, single-unit thermal incinerators can be built to control streams with flow rates in the range of 0.14 scm/sec (300 scfm) to about 24 scm/sec (50,000 scfm). The smallest commercially available packaged incinerators are designed to handle flow rates of 500 scfm.⁷ Assuming a turndown ratio of 10 to 1,⁸ the minimum flow rate to an incinerator is 50 scfm.

2.1.2.2 Acid Gas Scrubbing. Thermal incinerators used to reduce emissions of halogenated VOC's may require additional control equipment to remove corrosive combustion products (acid gas). The halogenated VOC streams are usually scrubbed after combustion to prevent equipment corrosion that results from contact with these acid gases. The flue gases are quenched to lower their temperature and are then routed through absorption equipment such as packed towers or liquid jet scrubbers. Section 2.2.5 discusses the operation and application of absorption equipment.

2.1.2.3 Thermal Incinerator Efficiency. Variations in chamber temperature, residence time, inlet VOC concentration,

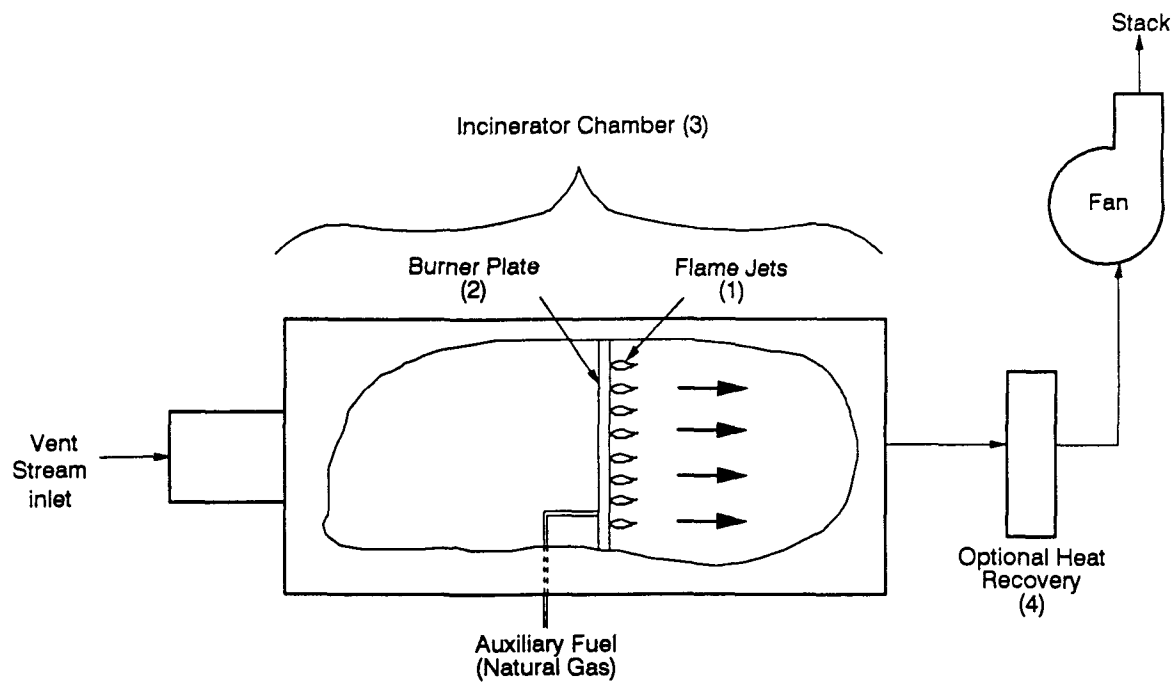


Figure 2-3. Distributed burner, thermal incinerator.⁶

compound type, and flow regime (mixing) affect the VOC destruction efficiency of a thermal incinerator. Performance tests demonstrate that thermal incinerators can achieve 98 percent destruction efficiency for most VOC's at combustion chamber temperatures ranging from 700 to 1,300 °C (1,300 to 2,370 °F) and residence times of 0.5 to 1.5 sec. For vent streams with VOC concentrations below approximately 2,000 ppmv, all new thermal incinerators can achieve outlet VOC concentrations of 20 ppmv or lower.⁹ These data indicate that significant variations in destruction efficiency occurred for C₁ to C₅ alkanes and olefins, aromatics (benzene, toluene, and xylene), oxygenated compounds (methyl ethyl ketone and isopropanol), chlorinated organic compounds (vinyl chloride), and nitrogen-containing compounds (acrylonitrile and ethylamines) at chamber temperatures below 760 °C (1,400 °F). A thermal incinerator properly designed and operated to produce the described conditions in the combustion chamber should be capable of higher than 98 percent destruction efficiencies for any nonhalogenated VOC.

At temperatures above 760 °C (1,400 °F), oxidation occurs more rapidly than gas diffusion mixing. The VOC destruction efficiency then becomes dependent upon the fluid mechanics (mixing) in the oxidation chamber. The flow regime must ensure rapid, thorough mixing of the vent stream, combustion air, and hot combustion products from the burner. This raises the temperature of the VOC-laden stream and maintains it at the combustion temperature, in the presence of excess oxygen, for sufficient time to complete oxidation.

Other parameters affecting incinerator performance are the heat content of the vent stream, the water content of the stream, and the amount of excess combustion air (the amount of air above the stoichiometric air needed for combustion). Combustion of a vent stream with a heat content less than 1.9 MJ/scm (50 Btu/scf) usually requires burning supplemental fuel to maintain the desired combustion temperature. Using recuperative heat exchangers to preheat combustion air can lessen or eliminate supplemental fuel requirements. Vent

streams with a heat content above 1.9 MJ/scm (50 Btu/scf) can support combustion but may need supplemental fuel for flame stability.

The maximum achievable VOC destruction efficiency decreases with decreasing inlet VOC concentration because combustion is slower at lower inlet concentrations. Therefore, a VOC weight percentage reduction based on the mass rate of VOC exiting the control device versus the mass rate of VOC entering the device is appropriate for vent streams with VOC concentrations above approximately 2,000 ppmv (corresponding to 1,000 ppmv VOC in the incinerator inlet stream because air dilution is typically 1:1), however, for lower inlet concentrations, an outlet concentration is more appropriate than a percent reduction. As previously stated, 98 percent VOC reduction is achievable by incineration for vent streams with VOC concentrations above 2000 ppmv, while an incinerator outlet concentration of 20 ppmv is achievable for vent streams with VOC concentrations below 2000 ppmv.

2.1.2.4 Applicability of Thermal Incinerators. Thermal incinerators are technically feasible control devices for most vent streams. They can be used for vent streams with any concentration and type of VOC, and they can be designed to handle minor fluctuations in flow rate. However, potential excessive fluctuations in flow rate (i.e., process upsets) might prevent the use of thermal incinerators and would require the use of a flare. The presence of halogens requires additional equipment such as scrubbers for acid gas removal.

2.1.3 Catalytic Incinerators

2.1.3.1 Description of Catalytic Incinerators. A catalyst promotes oxidation of some VOC's at a lower temperature than that required for thermal incineration. The catalyst increases the rate of the chemical reaction without becoming permanently altered itself. Catalysts typically used for VOC incineration include platinum and palladium. Other formulations used include metal oxides for emission streams containing chlorinated compounds.¹⁰ Inert substrates are

coated with thin layers of these materials to provide maximum surface area for contact with the VOC in the vent stream.

Figure 2-4 is a schematic of a catalytic incineration unit. The vent stream (1) is introduced into a mixing chamber (2) where it is heated to about 316 °C (600 °F) by contact with the hot combustion products from auxiliary burners (3). The heated mixture is then passed through the catalyst bed (4). Oxygen and VOC's migrate to the catalyst surface by gas diffusion and are adsorbed to the surface of the catalyst. Oxidation takes place at these active sites. Reaction products are desorbed from the active sites and transferred by diffusion back into the vent stream.¹¹ The combusted gas may then be passed through a waste heat recovery device (5), such as a cross flow exchanger, to preheat the incoming vent stream or combustion air before being discharged to the atmosphere.

2.1.3.2 Catalytic Incinerator Control Efficiency.

Catalytic incinerators can achieve overall VOC destruction efficiencies up to about 98 percent and HAP destruction efficiencies up to about 95 percent with space velocities in the range 30,000 to 100,000 gas hourly space velocity (GHSV).¹² The efficiency of the catalytic incinerator depends on operating temperature, oxygen concentration, catalyst activity, and the characteristics and concentration of the VOC in the vent stream.

The operating temperatures of combustion catalysts usually range from 316 to 650 °C (600 to 1,200 °F). Temperatures below this range can slow or stop the oxidation reaction. Higher temperatures can result in shortened catalyst life and possible oxidation of the catalyst from the support substrate.

The VOC content of the vent stream must be kept relatively constant and low enough that the catalyst is not overheated and its activity destroyed. To protect the catalyst from overheating, VOC concentrations are usually restricted to 25 to 30 percent of the LEL by insurance company safety requirements. Such concentrations can be achieved by

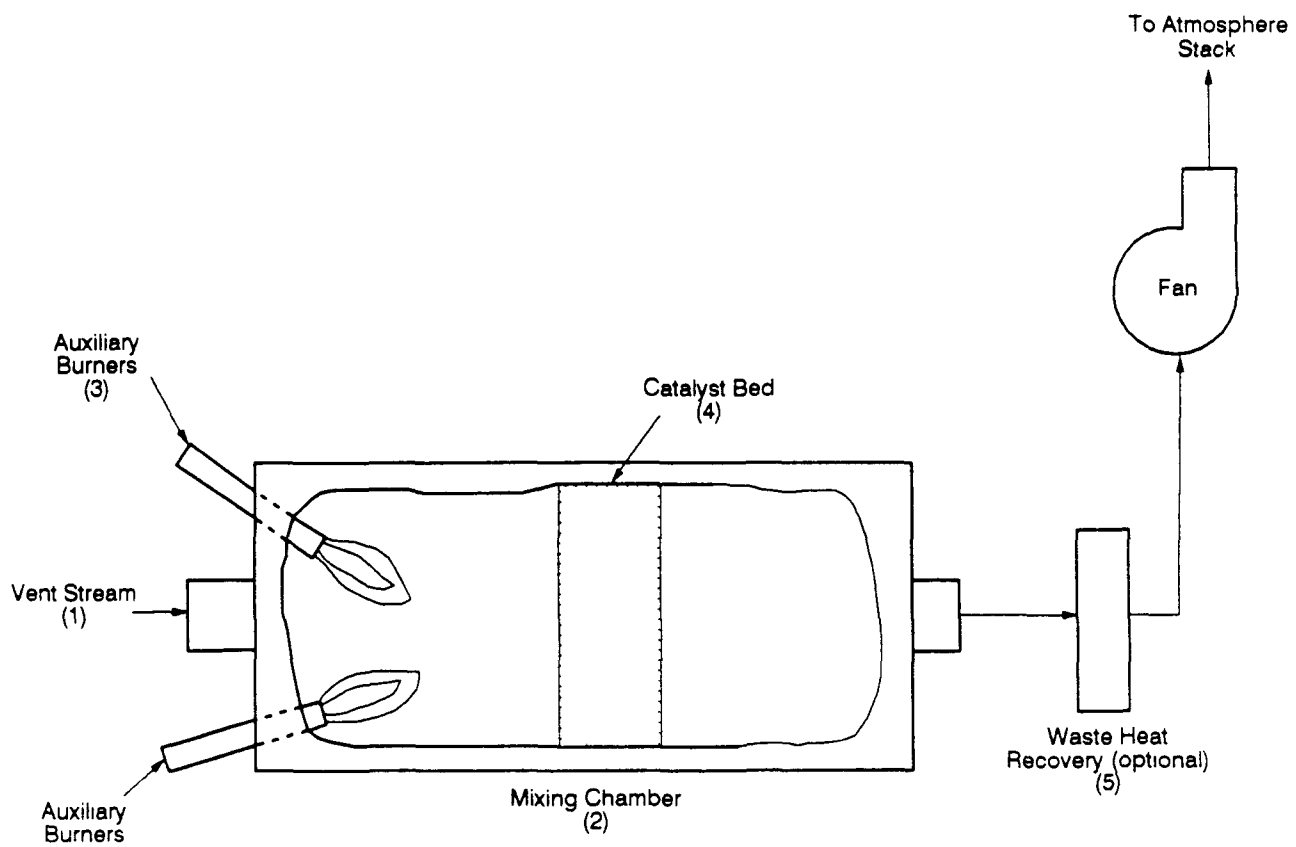


Figure 2-4. Catalytic incinerator.

diluting the vent stream with additional air if the theoretical air to meet the stoichiometric oxygen requirements for combustion is not sufficient.

Any accumulation of particulate matter, condensed VOC, or polymerized hydrocarbons on the catalyst can block the catalyst's effectiveness. Catalysts can also be deactivated by compounds containing sulfur, bismuth, phosphorous, arsenic, antimony, mercury, lead, zinc, tin, or halogens.¹³ If these compounds exist in the vent stream, the VOC will pass through the catalytic incinerator unreacted or be partially oxidized to form compounds (aldehydes, ketones, and organic acids) that are highly reactive atmospheric pollutants and can corrode plant equipment.

2.1.3.3 Applicability of Catalytic Incinerators. The applicability of catalytic incinerators for control of VOC's is limited by the catalyst deactivation sensitivity to the characteristics of the inlet stream. The vent stream to be combusted should not contain materials that can poison the catalyst or deposit on and block the reactive sites on the catalyst surface. In addition, catalytic incinerators are unable to handle high inlet concentrations of VOC or very high flow rates. Catalytic incineration is generally useful for concentrations of 50 to 10,000 ppmv, if the total concentration is less than 25 percent of the LEL and for flow rates of less than 100,000 scfm.¹⁴ Catalytic units are typically used for vent streams with stable flow rates and concentrations.

2.1.4. Industrial Boilers and Process Heaters

Industrial boilers and process heaters can be designed to combust VOC's by incorporating the vent stream into the inlet fuel or by feeding the stream into the boiler or heater through a separate burner. The main distinction between industrial boilers and process heaters is that the former produces steam usually at high temperatures while the latter raises the temperature of process streams.

2.1.4.1 Description of Industrial Boilers. Industrial boilers are combustion units that boil water to produce high

and low pressure steam. Industrial boilers can also be used to combust various vent streams containing VOC's, including vent streams from distillation operations, reactor processes, and other general operations. This description focuses on the use of industrial boilers to reduce emissions of VOC's from vent streams.

The majority of industrial boilers used in the chemical industry are of watertube design, and over half of these boilers use natural gas as a fuel.¹⁵ In a watertube boiler, hot combustion gases contact the outside of heat transfer tubes which contain hot water and steam. These tubes are interconnected by a set of drums that collect and store the heated water and steam. The water tubes have relatively small diameters, 5 cm (2.0 in.), which provide rapid heat transfer, rapid response to steam demands, and relatively high thermal efficiency.¹⁶ Energy transfer from the hot flue gases to the water in the furnace watertube and drum system can be better than 85 percent efficient. Additional energy can be recovered from the flue gas by preheating combustion air in an air preheater or by preheating incoming boiler feed water in an economizer unit.

When firing natural gas, forced- or natural-draft burners thoroughly mix the incoming fuel and combustion air. A VOC-containing vent stream can be added to this mixture or it can be fed into the boiler through a separate burner. In general, burner design depends on the characteristics of the fuel--either the combined VOC-containing vent stream and fuel or the vent stream alone (when a separate burner is used).

2.1.4.2 Description of Process Heaters. A process heater is similar to an industrial boiler in that heat liberated by the combustion of fuels is transferred by radiation and convection to fluids contained in tubular coils. Process heaters are used in many chemical manufacturing operations to drive endothermic reactions. They are also used as feed preheaters and as reboilers for some distillation operations. The fuels used in process heaters include natural gas, refinery offgases, and various grades of fuel oil.

Gaseous fuels account for about 90 percent of the energy consumed by process heaters.¹⁷

There are many variations in the design of process heaters depending on their application. In general, the radiant section consists of the burner(s), the firebox, and tubular coils containing the process fluid. Most heaters also contain a convective section in which heat is recovered from hot combustion gases by convective heat transfer to the process fluid.

2.1.4.3 Factors Affecting Control Efficiency of Industrial Boilers and Process Heaters. The combustion efficiency of boilers and process heaters, like that of incinerators, is determined by the average furnace temperature and residence time. However, when a vent gas is injected as a fuel into the flame zone of a boiler or process heater, the required residence time is reduced because of the relatively high temperature and turbulence of the flame zone.

Furnace residence time and temperature profiles vary for industrial boilers and process heaters depending on the furnace and burner configuration, fuel type, heat input, and excess air level.¹⁸ A mathematical model has been developed that estimates the furnace residence time and temperature profiles for a variety of industrial boilers.¹⁹ This model predicts mean furnace residence times ranging from 0.25 to 0.83 sec for natural gas-fired watertube boilers that range in size from 4.4 to 44 MW (15 to 150 MMBtu/hr). Boilers with a 44-MW capacity or greater generally have residence times and operating temperatures that would ensure a 98 percent VOC destruction efficiency. Furnace exit temperatures for this size of boiler are at least 1,200 °C (2,200 °F), with peak furnace temperatures in excess of 1,540 °C (2,810 °F).

Firebox temperatures for process heaters show relatively wide variations depending on the application. In the chemical industry, firebox temperatures can range from 400 °C (750 °F) for preheaters and reboilers to 1,260 °C (2,300 °F) for pyrolysis furnaces. Tests were conducted by the EPA to determine the benzene destruction efficiency of five process

heaters firing a benzene offgas and natural gas mixture.^{20,21,22} The units tested are representative of process heaters with low temperature fireboxes (reboilers) and medium temperature fireboxes (superheaters). The reboiler and superheater units tested showed greater than 98 percent overall destruction efficiency for C₁ to C₆ hydrocarbons.²¹ Additional tests conducted on a second superheater and a hot oil heater showed that greater than 99 percent overall destruction of C₁ to C₆ hydrocarbons occurred for both units.²²

2.1.4.4 Applicability of Industrial Boilers and Process Heaters. Industrial boilers and process heaters are used throughout the chemical industry to provide steam and heat input essential to chemical processing. In most cases, these industrial boilers and process heaters are of sufficient size to provide the temperature and residence time needed for destruction of VOC's. Further, industrial boilers and process heaters have proved effective in destroying compounds that are difficult to combust (e.g., polychlorinated biphenyls). Therefore, industrial boilers and process heaters can be used to reduce VOC emissions from any vent streams that are certain not to reduce the performance or reliability of the boiler or process heater.

The introduction of a vent stream into the boiler or process heater flame zone can alter the heat transfer characteristics of the furnace. Heat transfer characteristics depend on the flow rate, heat content, and elemental composition of the vent stream and on the size and type of heat-generating unit used. Often, there is no significant alteration of the heat transfer, and in some cases the VOC content of the vent stream can reduce the amount of fuel required to produce the desired heat release rate. In other cases, the change in heat transfer characteristics after introduction of a vent stream may affect the performance of the heat-generating unit and increase fuel requirements. For example, if the vent stream is a large-volume, low-VOC content stream, additional energy may be needed to raise this stream to the operating temperature of the boiler or process heater.

Ducting some vent streams to a boiler or process heater can present potential safety and operating problems. The varying flow rate and organic content of some vent streams can lead to explosive mixtures or flame instability within the furnace.

Because they are corrosive, vent streams with halogenated or sulfur-containing compounds are usually not combusted in boilers or process heaters. When corrosive VOC's are combusted, the flue gas temperature must be maintained above the acid dew point to prevent acid deposition and subsequent corrosion of downstream equipment such as ductwork, heat exchangers, and exhaust stacks.

Boilers and process heaters are most applicable where the potential exists for heat recovery from the combustion of the vent stream. For example, vent streams with a high VOC concentration and high flow rate can provide enough equivalent heat value to act as a substitute for fuel that would otherwise be needed. Because boilers and process heaters cannot tolerate wide fluctuations or interruptions in the fuel supply, they are not widely used to reduce VOC emissions from batch operations or other noncontinuous vent streams.

2.2 COLLECTION SYSTEMS AND RECOVERY DEVICES

This section discusses devices that reduce emissions of VOC by means of product recovery. Because these devices recover material without altering its chemical structure, recovered feedstocks may be returned to the process and recovered product may be collected and sold. A VOC is recovered by separating it from the rest of the stream with a process that is based on physical properties such as vapor pressure or solubility.

The recovery devices discussed in the following five subsections include loading rack vapor collection systems, condensers, steam strippers, carbon adsorbers, and absorbers. The discussion for each device treats its typical design and operation, factors affecting removal efficiency and applicability.

2.2.1 Vapor Collection Systems for Loading Racks

When liquids are transferred into a transport vessel, vapors in the head space of that vessel can be lost to the atmosphere. The principal factors affecting emissions from transfer operations are the vapor pressure of the chemical being transferred and the mode of transfer into the transport vessel. Other factors that may influence emissions from transfer operations include the transfer rate and the purge rate of nitrogen (or other inert gas) through the vessel during transfer.

The vapor pressure of the chemical being transferred has the greatest influence on emissions from transfer operations. For pure materials, the vapor pressure gives a measure of the amount of organic compound lost during transfer. The total potential emissions from any transfer is related to the void volume of the transport vessel and the concentration of the VOC in the head space. The saturation vapor pressure of the pure VOC is used in calculating the maximum concentration possible in the head space of the transport vessel. For low volatility compounds, this vapor pressure is small and therefore the emissions potential is small. For higher volatility compounds, more of the compound evaporates into the vapor space and is lost during transfer. For compounds with vapor pressures above 14.7 psia, the liquid is usually transferred under pressure, which effectively reduces the potential for emissions.

The mode of transfer (loading) is also an important factor in determining emissions from transfer operations. Top splash loading creates the most emissions because it enhances the agitation of the liquid being transferred, creating a higher concentration of the compound in the vapor space. With alternate loading techniques, such as submerged fill or bottom loading, the organic liquid is loaded under the surface of the liquid, which reduces the amount of agitation and suppresses the generation of excess vapor in the head space of the transport vessel.

The transfer rate has a more subtle influence on emissions; its greatest effect is on air quality. Transfer rate will dictate the short-term emission rate of the compound being transferred, thereby influencing exposure to the worker or public.

A nitrogen purge is used to reduce the potential for explosion of some chemicals in air or to keep some chemicals moisture-free. Using an inert gas purge increases the emission rate of VOC lost to the atmosphere because it creates a turnover rate of gas through the transport vessel, increasing the total volume of vapor discharged to the atmosphere. If an inert gas purge is used, the total volume of vapor loss is equal to the sum of the volume of the transport vessel and the volume of the purge gas used. During a transfer cycle, the volume of purge gas could be as great as the volume of the transport vessel, essentially doubling the losses to the atmosphere.

Most vapor collection systems collect the vapors generated during transfer operations and transport them to either a recovery device for return to the process or a combustion device for destruction. In vapor balancing systems, vapors generated during transfer operations are returned directly to the storage facility for the material. Assuming that incompressible fluids are being transferred, negligible or no vapors would be lost to the atmosphere from this "closed system," making no additional controls necessary.

2.2.1.1 Description of Vapor Collection Systems. Vapor collection systems consist of piping that captures and transports to a control device organic compounds in the vapor space of a transport vessel that are displaced when liquids are loaded. These systems may use existing piping normally used to transport liquids under pressure into the transport vessel or piping separate from that for transfer. Collection systems comprise very few pieces of equipment and minimal piping. The principal piece of equipment in a collection system is a vacuum pump or blower, used to induce the flow of

vapors from the transport vessel to the recovery or combustion system.

Effective collection and recovery systems have been used in some sectors of the SOCOMI in response to Occupational Safety and Health Administration (OSHA) regulations and EPA standards. These systems transfer a VOC under pressure into dedicated tank cars. Once the transfer has been completed, the transfer piping is placed under vacuum with vapors being drawn into a compression/refrigeration/condensation system. The recovered liquid is stored for reuse, and the exhaust from the recovery system is combusted in an on-site incinerator. Once the transfer system has been completely evacuated, the transfer piping is isolated by valves at the tank car and at the transfer arm. The only potential emissions from this system are the residual emissions from the combustion device and the connection losses associated with the residual VOC in the spool piece between the transfer arm and the transport vessel. This system represents one of the most efficient collection and recovery systems in the chemical industry.

Blowers can also be used to remove vapors from the head space of the tank car as liquid is transferred into the tank car. Standard recovery techniques such as condensation or refrigeration/condensation systems, or combustion can be applied to the captured vapors. The use of these lower efficiency collection systems would depend upon the physical properties of the chemicals transferred and on the efficiency of the recovery device.

Vapor balancing is another means of collecting vapors and reducing emissions from transfer operations. Vapor balancing is most commonly used where storage facilities are adjacent to the loading facility. As shown in Figure 2-5, an additional line is connected from the transport vessel to the storage tank to return any vapor in the transport vessel displaced by the liquid that is loaded to the vapor space of the storage vessel left by the transferred liquid. Because this is a direct volumetric change, there should be no losses to the atmosphere.

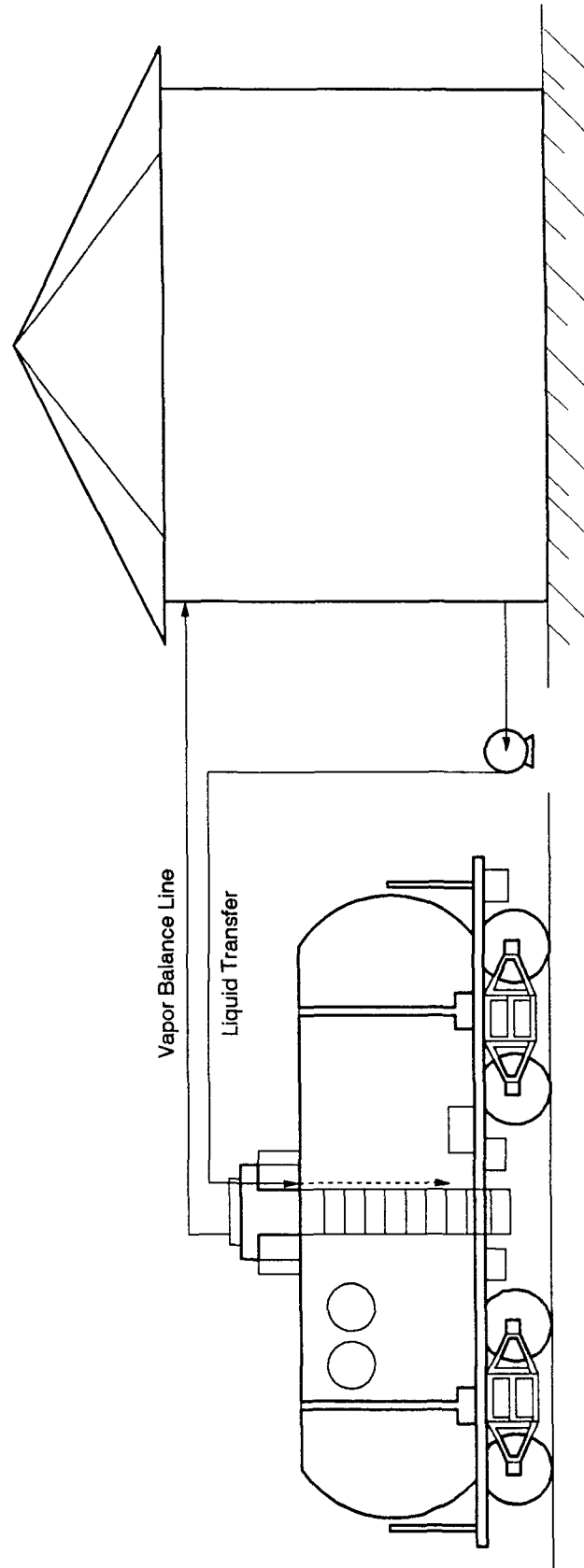


Figure 2-5. Vapor balancing system.

2.2.1.2 Factors Affecting Efficiency of Vapor Collection Systems. The factors affecting the efficiency of a vapor collection system include the following:

- Operating pressure of the collection system;
- Volume of piping between the loading arm and the transport vessel; and
- The efficiency of the ultimate control device.

The operating pressure of the collection system influences the efficiency of collection through the concentration of the VOC remaining in the line(s) after transfer. For systems that operate at lower pressure or under vacuum, the concentration of the VOC is decreased and the total amount of VOC in the piping lowered. This effectively reduces the amount of VOC that may be lost to the atmosphere when disconnecting transfer lines. For systems operating at higher pressures, there would be a larger quantity of VOC remaining in the piping that could be lost.

The volume of piping between the transfer loading arm and transport vessel, and the operating pressure of the collection system establish the quantity of VOC not delivered to the transport vessel and not collected for treatment. If the piping is opened to the atmosphere, this quantity of VOC would be lost, lowering the overall efficiency of the collection and control system. Systems that minimize the piping between the transfer loading arm and the transport vessel are more "efficient" than those with larger piping connections.

The overall efficiency of the collection system and control system is most dependent upon the efficiency of the control system. In the SOCFI, collection systems are generally hard-piped between the transport vessel and the control system. Therefore, there would be no loss of efficiency resulting from the collection system, other than losses associated with connections and disconnections.

2.2.1.3 Applicability of Vapor Collection Systems. The applicability of and decision to use various vapor collection systems depends upon several factors:

- Vapor pressure of the material;
- Value of the product;
- Physical layout of the facility; and
- OSHA considerations.

The vapor pressure of the material being transferred is a principal consideration in the design of the vapor collection system to be used. Materials with higher vapor pressures (greater than 14.7 psia) are stored and loaded under pressure. Loading under pressure eliminates the losses associated with atmospheric transfer operations and limits losses to those associated with connections and disconnections. Materials with very low vapor pressures have relatively little emissions potential and, therefore, little potential for emissions reduction. Costs of collection and recovery systems for these materials may be relatively high when considering the amount and value of the recovered material. Systems such as vapor balancing, however, would be an effective technique for control of emissions from transfer operations of low vapor pressure materials.

Proximity to the storage facility is the principal factor in deciding to use vapor balancing. The viability of installing a vapor balancing system is determined by the cost of piping required to return the vapor to the storage tank. When this distance is minimized, the cost is considered affordable. Because vapor balancing is a simple and cost effective control technique for transfer operations, it is often viewed as a means of achieving reasonably available control technology (RACT) requirements and has been used in many instances as a control measure to meet the emission requirements of State air toxic regulations. For purely economic considerations, expensive products are candidates for more extensive collection and recovery systems.

Additional considerations in the selection of a vapor collection and recovery system are OSHA limitations on work

place exposure to the chemicals being transferred. Some chemical compounds have extremely strict threshold limit values (TLV's) and, therefore, must be more tightly controlled than other, less toxic compounds. The system described above for the control of emissions from existing transfer operations for some VOC's (using vacuum, vapor compression, refrigeration, and combustion) is an example of extensive control measures that limit exposure of individuals to the chemical being transferred. Such systems have been used for highly toxic or carcinogenic compounds.

2.2.2 Condensers

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: (1) by increasing the system pressure at a given temperature, or (2) by lowering the temperature at a constant pressure. This section addresses the latter method.

In a two-component system where one of the components is noncondensable (e.g., air), condensation occurs at dew point (saturation) when the partial pressure of the volatile compound is equal to its vapor pressure. For more volatile compounds (i.e., compounds with lower normal boiling points), a larger amount of the compound remains as vapor at a given temperature; hence, to remove or recover the compound, a lower temperature would be required for saturation and condensation. For such cases, refrigeration can be used to obtain the lower temperatures needed to achieve acceptable removal efficiencies.

2.2.2.1 Description of Condensers. Figure 2-6 depicts a typical configuration for a refrigerated surface condenser system. The basic equipment includes a condenser, refrigeration unit(s), and auxiliary equipment (e.g., precooler, recovery/storage tank, pump/blower, and piping).

The two most commonly used condenser types are surface condensers and direct contact condensers.²³ In surface condensers, the coolant fluid does not contact the vent

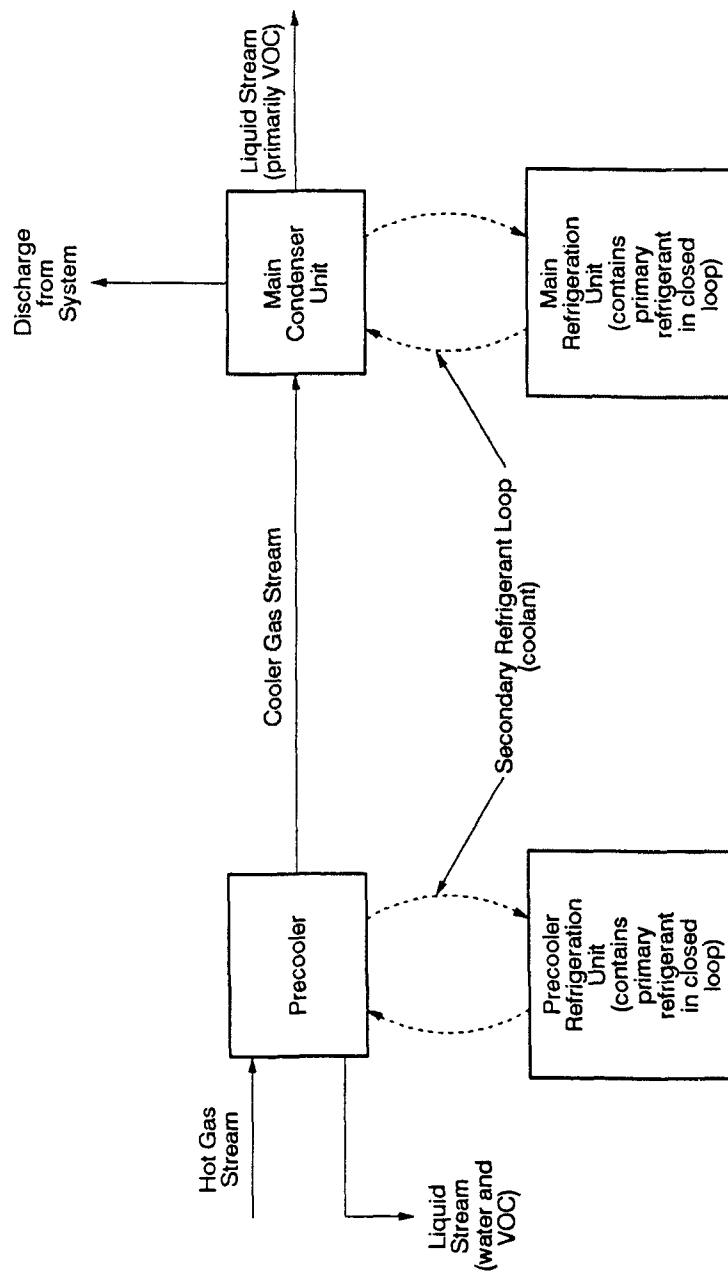


Figure 2-6. Refrigerated condenser system for VOC vapor recovery.

stream; heat transfer occurs through the tubes or plates in the condenser. As the vapor condenses, a film forms on the cooled surface and drains away to a collection tank for storage, reuse, or disposal. Because 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 VOC's from the gas stream.

Most surface condensers in refrigerated systems are the shell-and-tube type (Figure 2-7), which circulates the coolant fluid on the tube side. The VOC's condense on the outside of the tube (the shell side). Plate-type heat exchangers are also used as surface condensers in refrigerated systems. Plate condensers operate under the same principles as the shell-and-tube systems (i.e., no contact between coolant and vent stream), but the two streams are separated by thin, flat plates instead of cylindrical tubes.

In contrast to surface condensers, direct contact condensers cool the vapor stream by spraying a liquid at ambient or lower temperature directly into the vent stream. Spent coolant containing VOC's from direct contact condensers usually cannot be reused directly. Additionally, VOC's in the spent coolant cannot be recovered without further processing. The combined VOC/coolant stream could present a potential waste disposal problem, depending upon the coolant and the specific VOC's.

For many VOC recovery needs, a refrigeration unit generates the low-temperature medium necessary for heat transfer. In refrigerated condenser systems, two kinds of refrigerants are used--primary and secondary. Primary refrigerants such as ammonia (R-717), and chlorofluorocarbons such as chlorodifluoromethane (R-22) or dichlorodifluoromethane (R-12), are those that undergo a phase change from liquid to gas after absorbing heat. Secondary refrigerants or coolants, such as brine solutions, act only as heat carriers and remain in the liquid phase. Conventional systems use a closed primary refrigerant loop that cools the secondary

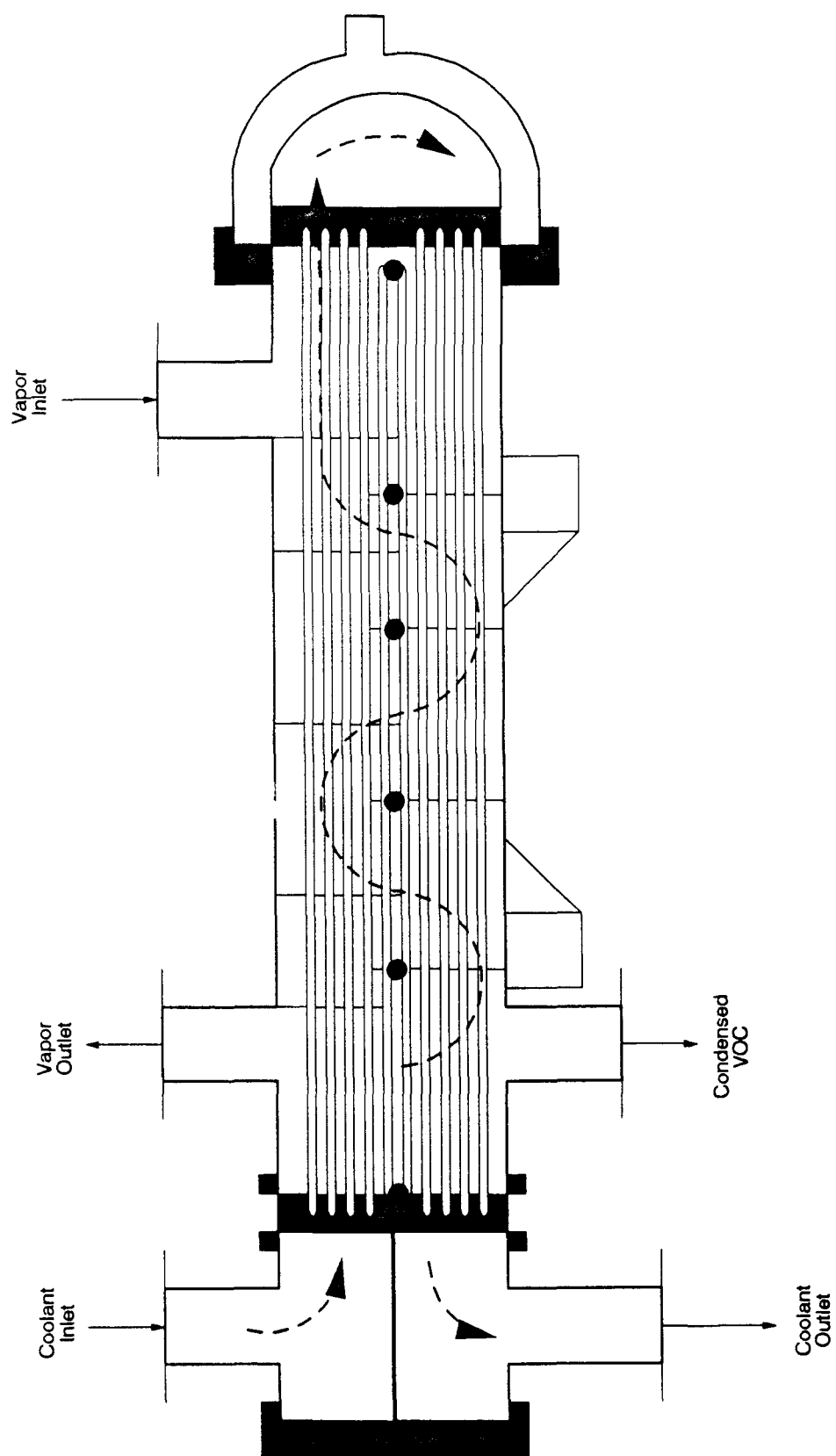


Figure 2-7. Schematic diagram of a shell and tube surface condenser.²⁴

(coolant) loop through heat transfer in the evaporator for the primary refrigerant. The coolant is then pumped to a condenser where it is used to cool the air/VOC vapor stream. In some limited applications, however, the primary refrigerant is used directly to cool the vapor stream.

For applications requiring low temperatures (below about -30 °F), multistage refrigeration systems are frequently employed.²⁵ Multistage systems are of two types--compound and cascade. In compound systems, refrigerant compression stages are connected in series and 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 normally considered for applications requiring temperatures between -50 to -150 °F and allow the use of different refrigerants in each cycle.²⁵ Theoretically, any number of cascaded stages are possible, each stage requiring an additional condenser and an additional stage of compression.

Some applications may require auxiliary equipment such as precoolers, recovery/storage tanks, pumps/blowers, and piping. If the vent stream contains water vapor 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 used to remove the moisture before the vent stream enters the condenser. This precooler would cool the vent stream to approximately 35 to 40 °F, effectively removing the moisture from the vent stream. Alternatively, ice can be melted during an intermittent heating cycle by circulating ambient temperature brine through the condenser or using radiant heating coils. If a system is not operated continuously, ice can be removed by circulating ambient air.

A recovery tank for temporary storage of condensed VOC before its reuse, reprocessing, or transfer to a large storage tank may be necessary in some cases. Pumps and blowers are

typically used to transfer liquid (e.g., coolant and recovered VOC) and gas streams, respectively, within the system.

2.2.2.2 Factors Affecting Condenser Control Efficiency.

The removal efficiency of refrigerated surface condenser systems designed to control vent streams containing air/VOC mixtures depends primarily on the following parameters:

- Volumetric flow rate of the VOC-containing vent stream;
- Inlet temperature of the vent stream;
- Concentrations of the VOC's in the vent stream;
- Absolute pressure of the vent stream;
- Moisture content of the vent stream; and
- Properties of the VOC's in the vent stream:
 - dew points,
 - heats of condensation,
 - heat capacities, and
 - vapor pressures.

Any component of any vapor mixture can be condensed if brought to a low enough temperature and allowed to come to equilibrium, but a condenser cannot lower the VOC 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.

2.2.2.3 Applicability of Condensers. Condensers are widely used as raw material and/or product recovery devices. They may be used to recover VOC's upstream of other control devices or they may be used alone for controlling vent streams containing high VOC concentrations (usually greater than 5,000 ppmv). In these cases, the removal efficiencies of condensers can range from 50 to 95 percent.

The temperature necessary for condensation depends on the properties and concentration of VOC's in the vent stream. Streams having low VOC concentrations and streams containing more volatile (low boiling point) compounds require lower condensation temperatures. Because condenser size and cost

are a function of condensation temperature, condensation may be economically infeasible for some vent streams.

Depending on the type of condenser used, disposal of the spent coolant can be a problem. If cross-media impacts are a concern, surface condensers would be preferable to direct contact condensers.

Condensers used as emission control devices can process flow rates as high as about 2,000 scfm. Condensers for vent streams with flow rates above 2,000 scfm and high concentrations of noncondensables will require significantly larger heat transfer areas. These systems may be costly when compared to the potential recovery of VOC or to other emission reduction techniques.

2.2.3 Steam Strippers

Steam stripping can be used as initial treatment of a process wastewater stream to reduce the VOC loading of that stream before it is sent to the facility-wide wastewater treatment system. A steam stripping system comprises several components (including other devices described for control of VOC's): a feed tank, heat exchanger, steam stripping column, condenser, overheads receiver, and a destruction device (if necessary).

2.2.3.1 Description of Steam Strippers. Steam stripping involves the fractional distillation of wastewater to remove VOC's. The basic operating principle of steam stripping is the direct transfer of heat through contact of steam with wastewater. This transfer of heat more easily vaporizes the more volatile organic compounds. The overhead vapor that contains water and organic compounds is condensed and separated (usually in a decanter) to recover the organic fraction. These recovered organic compounds are usually either recycled for reuse in the process or incinerated in an on-site combustion device for heat recovery.

Steam stripper systems may be operated in batch or continuous mode. Batch steam strippers are more prevalent when the wastewater feed is generated by batch processes, when the characteristics of the feed are highly variable, or when

small volumes of wastewater are generated. Batch strippers may also be used if the wastewater contains relatively high concentrations of solids, resins, or tars. With batch stripping, wastewater to be steam stripped is charged to the receiver, or pot, and brought to the boiling temperature of the mixture. Heat is provided by direct injection of steam or by an external heat exchanger, or reboiler. Solids, tars, resins, and other residues remaining in the pot at the completion of the batch are nonvolatile, heavy compounds that are removed for disposal. Usually, batch steam strippers provide a single equilibrium stage of separation. Therefore, the removal efficiency is essentially determined by the equilibrium coefficients of the pollutants and the fraction of the initial charge distilled overhead. By varying the heat input and fraction of the initial charge boiled overhead, a batch stripper can be used to treat wastewater mixtures with widely varying characteristics.

In contrast to batch strippers, continuous steam strippers are designed to treat wastewater streams with relatively consistent characteristics. Design of the continuous stripper system is based on the flow rate and composition of a specific wastewater feed stream or combination of streams. Multistage, continuous strippers normally achieve greater efficiencies of organic compound removal than batch strippers. Continuous systems may offer other advantages over batch stripping for applications involving wastewater streams with relatively high flows and consistent concentrations. These advantages include more consistent effluent quality, more automated operation, and lower annual operating costs.

Wastewater streams continuously discharged from process equipment are usually relatively consistent in composition. Such wastewater streams would be efficiently treated with a continuous steam stripper system. However, batch wastewater streams can also be controlled by continuous steam strippers by incorporating a feed tank with adequate residence time to provide a consistent outlet composition. Because the system

can be designed to provide a stripper feed of consistent quality, the remaining discussion focuses on continuous steam stripping.

A generic continuous steam stripper system is shown in Figure 2-8. For more effective control in specific cases, alternate feed locations and multiple feed locations are sometimes used. The steam stripper can also be operated under a vacuum or may include a reflux stream where the bottoms stream flows into a reboiler to vaporize, returning a portion of the bottoms stream to the column. In addition, the pH of the feed stream may be altered to change the equilibrium of a low volatility compound, improving its removal through stripping.

The purpose, design, and operation of each of the components of a steam stripping system (a wastewater feed tank, feed/bottoms heat exchanger and steam stripping column, vent lines, condenser system, and ancillary pumps) are discussed below in relation to their functions: collecting and conditioning the wastewater, steam stripping of the wastewater, controlling vents and openings in the system, and recovery of the steam-stripped organic compounds.

2.2.3.1.1 Collecting and conditioning the wastewater.

The controlled sewer system or hard piping from the point of wastewater generation to the feed tank controls emissions before steam stripping. The feed tank, which is covered and vented to an on-site combustion device, collects and conditions the wastewater fed to the steam stripper. The feed tank is usually sized to provide a desired hydraulic retention time of 0.5 to 40 hr for the wastewater feed stream.^{26,27} The desired retention time depends primarily on the variability in wastewater flow rate, characteristics of the inlet wastewater, and the amount of wastewater conditioning needed. Additional surge capacity can provide retention time for wastewater streams with highly variable flow rates to maintain a relatively constant feed rate to the stripper.

If the feed tank is adequately designed, a continuous steam stripper can treat wastewater generated by some batch

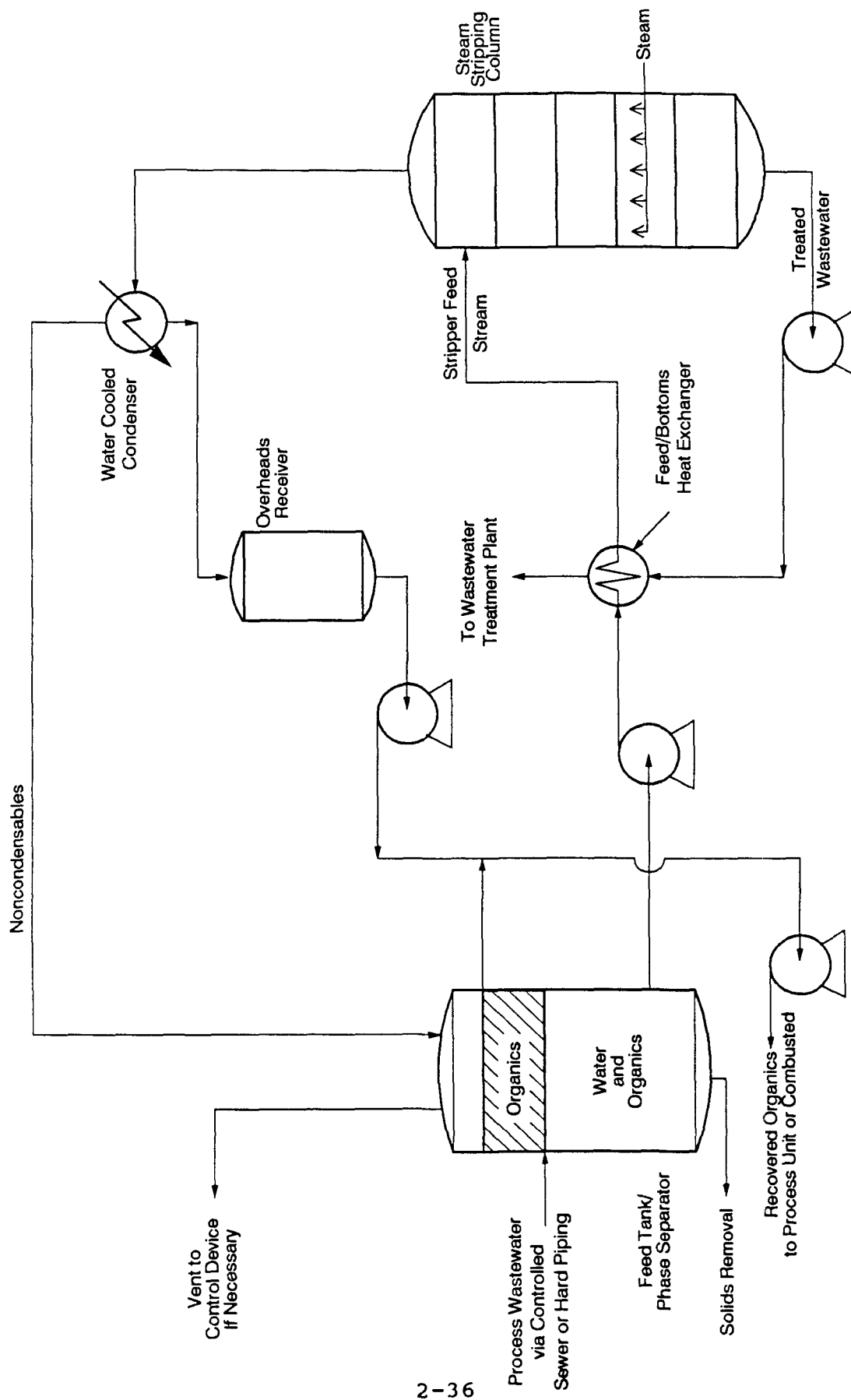


Figure 2-8. Continuous steam stripper system.

processes. In these cases, the feed tank serves as a buffer between the batch process and the continuous steam stripper. During periods of no wastewater flow from the batch process, wastewater stored in the feed tank is fed to the stripper at a relatively constant rate.

The retention time in the feed tank also depends on the degree of conditioning required for the stripper feed stream. Aqueous and organic phases are often present in the stripper feed tank. The feed tank provides the retention time necessary for these phases to separate. Oils and tars normally partition from the water into the organic phase, which is either recycled to the process for recovery of the organic compounds or disposed by incineration. The water phase is fed to the stripper to remove the soluble organic compounds. Solids are also separated in the stripper feed tank; the separation efficiency depends on the density of the solids dissolved in the process wastewater. Some of the less dense solids may remain suspended in the organic or aqueous phases, while the more dense solids settle to the bottom of the tank. These more dense solids are periodically removed from the feed tank and are usually landfilled or landfarmed.

2.2.3.1.2 Steam stripping of the wastewater. After the wastewater is collected and conditioned, it is pumped through the feed/bottoms heat exchanger where it is preheated and then pumped into the steam stripping column. Typically, steam is sparged directly into the stripper at the bottom of the column, and the wastewater feed is introduced into the stripper at the top of the column (Figure 2-8). As the wastewater flows down the column, it contacts the steam flowing countercurrently up the column. Both latent and sensible heat is transferred from the steam to the organic compounds in the wastewater, vaporizing them into the gaseous stream. These constituents flow out the top of the column with any uncondensed steam.

The wastewater effluent leaving the bottom of the steam stripper is pumped through the feed/bottoms heat exchanger which heats the feed stream and cools the bottoms before

discharge. This exchange of waste heat also improves the economy of the steam stripper system. After passing through the heat exchanger, the bottoms stream is usually either routed to an on-site wastewater treatment plant (Figure 2-8) and discharged to a National Pollutant Discharge Elimination System (NPDES)-permitted outfall, or sent to a publicly owned treatment works (POTW).

2.2.3.1.3 Controlling vents and openings in the steam stripper system. In a steam stripper system, vent lines carry gaseous organics, water vapor, and noncondensables to a control device. For the stripper in Figure 2-8, vent lines are placed between the stripper column and primary condenser, between the primary condenser and feed tank, and between the feed tank and existing on-site combustion device. All vent lines are controlled with a combustion device (e.g., thermal and catalytic incinerators, flares, boilers, or process heaters) or a product recovery device (e.g., condensers, carbon adsorbers, or absorbers).

Openings in a steam stripper system that allow for pressure relief, venting, and/or maintenance access are sealed unless in use. Pressure relief valves can be vented to a control device. Because the typical stripper is operated at pressures only slightly greater than atmospheric, pressure relief emissions are typically negligible. Manways and other access areas for maintenance and cleaning are never used during operation and are sealed to prevent emissions during operation.

2.2.3.1.4 Recovery of the steam stripped organic compounds. A condenser system can be used to recover the organic and water vapors in the gaseous overheads stream from the stripping column. If the primary condenser is not sufficient for condensing a large portion of the organic compounds, it may be necessary to install a secondary condenser with brine or a refrigerant. The condensed overheads stream is fed to an overhead receiver. The recovered organic compounds are usually either pumped to storage and recycled to the process unit or combusted for

their fuel value in an incinerator, boiler, or process heater. These combustion devices are discussed in Section 2.1. If an aqueous phase is generated in the overhead receiver, it is returned to the feed tank and recycled through the steam stripper system. Any noncondensable gases and highly volatile organic compounds not recovered by the condenser system are routed to an on-site control device such as a carbon adsorber, boiler, process heater, or an incinerator. Often in practical systems, the vent from the overhead condenser is piped to the feed tank, which is vented to the control device.

2.2.3.2 Factors Affecting Steam Stripper Control Efficiency. Steam stripping achieves emission reductions of 0 to 99 percent, based on the chemical characteristics (e.g., strippability) of the wastewater stream. However, 95 to 99 percent reduction can be achieved for the majority of compounds regulated by the HON. The organic compound removal performance of the steam stripper depends on the degree of contact between the steam and the wastewater. Several factors affecting the degree of contact that occurs in the steam stripper column are: (1) the dimensions of the column (height and diameter); (2) the contacting media in the column (trays or packing); and (3) operating parameters such as the steam-to-feed ratio, column temperatures, and pH of the wastewater.

Increasing the column height increases removal efficiency by increasing the opportunity for contact between the steam and the wastewater. The column height is determined by the number of theoretical stages required to achieve the desired removal efficiency. The number of theoretical stages is a function of the equilibrium coefficient of the pollutants and the efficiency of mass transfer in the column.

The diameter of the column is determined from the required cross-sectional area for liquid and vapor flow through the column. High superficial gas velocities increase turbulence and mixing and result in high column efficiencies. However, the column cross-sectional area must be sufficient to prevent flooding from excessive liquid loading or liquid entrainment due to excessive vapor velocity. The cross-

sectional area also affects the liquid retention time in the column, with higher retention times resulting in higher efficiencies. These opposing factors are considered in selecting the design values of superficial gas and liquid velocities (flow per square unit of tower cross-sectional area) and the tower diameter.

The contacting media in the column play a major role in determining the mass transfer efficiency. Typically, steam stripping columns are equipped with trays (e.g., bubble-cap or sieve) or packing to provide contact between the vapor and liquid phases. Trays are regularly spaced throughout the column and provide staged contact between the two phases; packing provides for continuous contact. Trays are usually more effective for wastewater containing dispersed solids because of the plugging and cleaning problems encountered with packing. Also, tray towers can operate efficiently over a wider range of liquid flow rates than packed towers. Packed towers are often more cost effective to install and operate when treating highly corrosive wastewater streams because corrosion resistant ceramic packing can be used. Also, the pressure drop through packed towers may be less than the pressure drop through tray towers designed for equivalent wastewater loadings. Packed towers are seldom designed with diameters greater than 4 ft, and column heights may be more limited than that of tray towers due to crushing of the packing near the bottom of the column. However, with the use of intermediate packing supports, crushing problems can be reduced.

An increase in the steam-to-feed ratio will increase the ratio of the vapor to liquid flow through the column. This increases the stripping of organic compounds into the vapor phase. Because additional heat is provided when the steam rate is increased, additional water is also vaporized. Therefore, an increase in the steam-to-feed ratio is also accompanied by an increase in the steam rate flowing out of the column in the overheads stream. Steam-to-feed ratios generally range from 0.01 to 1.0 kg/kg.

The steam-to-feed ratio required for high removal efficiency is affected by the temperature of the wastewater fed to the column. If the feed temperature is lower than the operating temperature at the top of the column, part of the steam is required to heat the feed. Good column design provides sufficient steam flow to heat the feed and volatilize the organic constituents. Steam in excess of this flow rate helps to carry the organic compounds out of the top of the column with the overheads stream.

Column operating temperature and wastewater pH also affect organic compound removal performance. Temperature affects the solubility and equilibrium coefficients of the organic compounds. Stripping columns are usually operated at pressures slightly greater than atmospheric, and operating temperatures are usually slightly greater than the normal boiling point of water. Wastewater pH is often controlled by adding caustic to the feed to raise the pH and change the vapor-liquid equilibrium characteristics of the compound in the wastewater matrix and enhance the removal of organic compounds being steam stripped.²⁸

Removal efficiency of a steam stripper decreases with low inlet organic concentrations. Above an inlet concentration threshold of 50 to 100 ppm, the removal efficiency for most compounds is not affected by the concentration of organic compounds in the wastewater; however, at concentrations well below this threshold, steam strippers lose some effectiveness.

Finally, organic compound removal efficiencies are higher for steam strippers treating wastewater containing chlorinated organic compounds. These chlorinated organic compounds are more easily steam stripped than organic compounds such as phenol which are more soluble in water and less volatile.

2.2.3.3 Applicability of Steam Strippers. Based on industry responses to Section 114 information requests, 10 percent of the reported wastewater streams account for greater than 90 percent of the volatile organic HAP compounds by mass. In addition, 20 percent of the total reported wastewater flow accounts for 85 percent of the volatile

organic HAP compounds by mass.²⁹ Therefore, it may be possible to achieve significant VOC emission reduction by controlling a relatively small number of individual wastewater streams and a relatively small volume of wastewater. In many cases, it may be possible to combine two or more of these streams for treatment by the same steam stripper. By hard piping these streams from the point of generation to the steam stripper, emissions from transporting the wastewater are minimized. As streams are combined, the cost of control increases, but the cost of control per stream decreases. This issue is discussed further in the presentation of steam stripper costs in Chapter 3.

Steam stripping is most applicable to treating wastewaters with organic compounds that are highly volatile and have a low solubility in water. The VOC's that have low volatility tend to volatilize less readily and thus are not easily stripped out of the wastewater by the steam. Similarly, VOC's that are very soluble in water tend to remain in the wastewater and also are not easily stripped out by steam. There is also an interactive effect between these two properties; for example, methanol is fairly volatile but has a high solubility in water. Steam stripping wastewaters containing methanol removes approximately one third of the actual methanol content. At least one steam stripper studied operated at an altered pH to change the vapor-liquid equilibrium characteristics of a less volatile compound. The lower-volatility compound was removed efficiently through this pH change.

Oil, grease, and solids content and pH of a wastewater stream also affect the applicability of steam stripping to a particular wastewater. High levels of oil, grease, and solids can cause operating problems for steam strippers. High or low pH may prove to be corrosive to equipment. However, these problems can usually be circumvented by design or wastewater preconditioning techniques.

2.2.4 Carbon Adsorbers

2.2.4.1 Description of Carbon Adsorbers. Adsorption is a mass-transfer operation involving interaction between gaseous- or liquid-phase components and solid-phase components. This section addresses the adsorption of VOC's from vent streams. The gas phase (adsorbate) is captured on the solid-phase (adsorbent) surface by physical or chemical adsorption mechanisms. Physical adsorption takes place when intermolecular (van der Waals) forces attract and hold the gas molecules to the solid surface. Chemisorption occurs when a chemical bond forms between the gaseous- and solid-phase molecules. A physically adsorbed molecule can be removed readily from the adsorbent (under suitable temperature and pressure conditions); the removal of a chemisorbed component is much more difficult.

The most common industrial adsorption systems use activated carbon as the adsorbent. Activated carbon effectively captures certain organic vapors by physical adsorption. The vapors can then be released for recovery by regenerating the adsorption bed with steam or nitrogen. Compared to activated carbon, oxygenated adsorbents such as silica gels, diatomaceous earth, alumina, or synthetic zeolites exhibit a greater selectivity for capturing water vapor than organic gases. Thus, these oxygenated adsorbents would be of little use for the high-moisture vent streams characteristic of some VOC-containing vent streams.³⁰

The design of a carbon adsorption system depends on the chemical characteristics of the VOC being recovered, the physical properties of the inlet stream (temperature, pressure, and volumetric flow rate), and the physical properties of the adsorbent. The mass of VOC that adheres to the adsorbent surface is directly proportional to the difference in VOC concentration between the gas phase and the solid surface. In addition, the quantity of VOC adsorbed depends on the adsorbent bed volume, the surface area of adsorbent available to capture VOC, and the rate of diffusion of VOC through the gas film at the gas- and solid-phase

interface (the mass transfer coefficient). Physical adsorption is an exothermic operation that is most efficient within a narrow range of temperature and pressure.

Figure 2-9 is a schematic diagram of a typical fixed-bed, regenerative carbon adsorption system. Before it enters the carbon bed, the inlet vent stream is usually filtered to prevent bed contamination by soot, resin droplets, and large particles entrained in the vent stream and cooled to maintain the bed at optimum operating temperature and to prevent fires or polymerization of the hydrocarbons (1). The filtered and cooled vapors enter the adsorber stage of the system (2), where they pass through the porous-activated carbon bed.

The dynamics of the process may be illustrated by viewing the carbon bed as a series of layers or mass-transfer zones (3a, b, c). Gases entering the bed are highly adsorbed first in Zone a. Because most of the VOC is adsorbed in Zone a, very little adsorption takes place in Zones b and c. Adsorption in Zone b increases as Zone a reaches equilibrium with organic compounds. The process continues until Zone c is used. When the bed is completely saturated (breakthrough), the incoming VOC-laden stream is routed to an alternate bed while the saturated carbon bed is regenerated.

The carbon bed is regenerated by heating the bed or applying vacuum to draw off the adsorbed gases. Low pressure steam (4) is frequently used as a heat source to strip the adsorbent (carbon) of the VOC collected. After steaming, the carbon bed is cooled and dried, typically by blowing air through it with a fan, and the steam-laden vapors are routed to a condenser (5) and on to a VOC recovery system (6). The regenerated bed is returned to active service while the saturated bed is purged of VOC. The regeneration process may be repeated numerous times; eventually, however, the carbon must be replaced.

2.2.4.2 Factors Affecting Carbon Adsorption Control Efficiency. For well designed and operated carbon adsorber systems, continuous VOC removal efficiencies of more than 95 percent are achievable for a variety of solvents, including

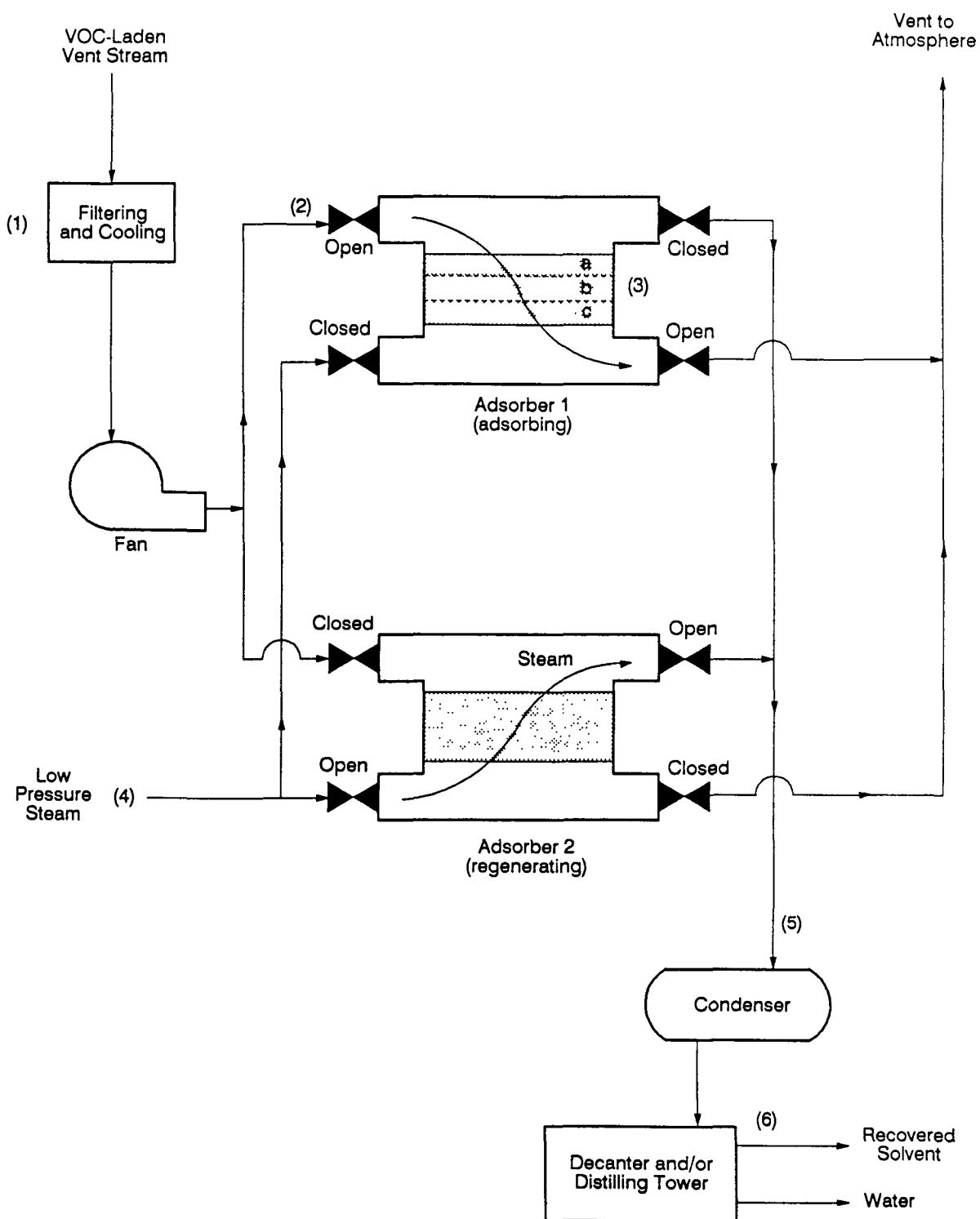


Figure 2-9. Two-stage regenerative adsorption system.

mixtures that contain ketones such as methyl ethyl ketone and cyclohexanone. Several plants have been shown to continuously achieve removal efficiencies of 97 to 99 percent.³¹ The VOC removal efficiency of an adsorption unit depends on the inlet vent stream characteristics, the physical properties of the compounds present in the vent stream, the physical properties of the adsorbent, and the condition of the regenerated carbon bed.

Inlet stream temperature, pressure, and velocity affect adsorption unit efficiency. The adsorption capacity of the carbon and the resulting outlet concentration are dependent upon the temperature of the inlet vent stream. High vent stream temperature increases the kinetic energy of the gas molecules, causing them to overcome van der Waals forces and release from the surface of the carbon. At vent stream temperatures above 100 °F, both adsorption capacity and outlet concentration may be adversely affected. In these cases, inlet stream coolers are typically used.³²

Increasing vent stream pressure improves VOC removal efficiency. Increased stream pressure results in higher concentration of VOC in the vapor phase and an increased driving force for mass transfer to the carbon surface. Conversely, carbon beds are often regenerated by inducing low pressure or a vacuum. Reducing the pressure in the carbon bed effectively lowers the concentration of VOC's in the vapor phase, desorbing the VOC's from the carbon surface to the vapor phase.

The velocity of the vent stream entering the carbon bed must be quite low to allow time for diffusion and adsorption. Linear velocities of 50 to 100 fpm are typically used. At higher velocities the pressure drop across the carbon bed becomes too high for standard blowers. At lower velocities, the bed becomes too large and expensive. If inlet concentrations are low (as is expected in the SOCOMI) the bed area required for the volume of carbon needed usually permits a velocity at the high end of this range.³³

The required depth of the bed for a given compound is directly proportional to the carbon granule size and porosity and to the inlet vent stream velocity. For a given carbon type, bed depth must increase as the vent stream velocity increases. Generally, carbon adsorber bed depths range from 1.5 to 3.0 ft. A bed depth of at least 1.5 ft is used to ensure that the bed is substantially deeper than the mass transfer zone.³⁴

The condition of the carbon bed will change with use. After repeated regeneration, the carbon bed loses activity, resulting in reduced VOC removal efficiency.

2.2.4.3 Applicability of Carbon Adsorption. Although carbon adsorption is an excellent method for recovering some valuable process chemicals, it cannot be used universally for distillation or process vent streams. Carbon adsorption is not recommended under the following conditions, which are common with many VOC-containing vent streams: (1) high VOC concentrations, (2) very high or low molecular weight compounds, (3) mixtures of high and low boiling point VOC's, and (4) high moisture content.

The maximum practical inlet concentration of VOC is typically 10,000 ppmv.³⁵ Adsorbing vent streams with high VOC concentration may result in excessive temperature rise in the carbon bed due to the accumulated heat of adsorption resulting from the VOC loading. If flammable vapors are present, insurance company requirements may limit inlet concentrations to less than 25 percent of the LEL.³⁵ However, vent streams with high VOC concentrations can be diluted with air or inert gases to make a workable adsorption system.

The molecular weight of the compounds to be adsorbed should be in the range of 45 to 130 gm/gm-mole for effective adsorption. High molecular weight compounds that are characterized by low volatility are strongly adsorbed on carbon. The affinity of carbon for these compounds makes it difficult to remove them during regeneration of the carbon bed. Hence, carbon adsorption is generally not applied to such compounds (i.e., boiling point above 400 °F and molecular

weight greater than about 130). Because highly volatile materials (i.e., molecular weight less than about 45) do not adsorb readily on carbon, adsorption is not typically used for controlling emission streams containing such compounds.³⁵

Properly operated adsorption systems can be very effective with homogeneous vent streams but less so with a stream containing a mixture of light and heavy hydrocarbons. The lighter organic compounds tend to be displaced by the heavier (higher boiling) components, greatly reducing system efficiency.³⁶

At adsorbate concentrations above 1,000 ppmv, humidity does not significantly affect working capacity. However, at adsorbate concentrations below 1,000 ppmv, water vapor competes with the VOC's in the vent stream for adsorption sites on the carbon surface.³⁷ In these cases, vent stream humidity levels exceeding 50 percent (relative humidity) are not desirable. Moisture content may be reduced by adding drier dilution air to the vent stream or by passing the stream through a heat exchanger.³⁸

2.2.5 Absorbers

2.2.5.1. Description of Absorbers. Absorption is the selective transfer of one or more components of a gas mixture into a liquid solvent. The transfer consists of solute diffusion and dissolution into a solvent. For any given solvent, solute, and operating conditions, there exists an equilibrium ratio of solute concentration in the gas mixture to solute concentration in the solvent. The driving force for mass transfer at a given point in an operating absorber is the difference between the concentration of solute in the gas and the equilibrium concentration of solute in the liquid. If the solvent contains an additive that will react with the solute, absorption will be enhanced because the equilibrium concentration of solute in the liquid will be zero.

Devices based on absorption principles include spray towers, venturi and wet impingement scrubbers, packed columns, and plate columns. Spray towers require high atomization pressure to obtain solvent droplets 500 to 100 μm in size,

necessary for a sufficiently large surface area for contact between the liquid and gas streams.³⁹ Although they can remove particulate matter effectively, spray towers have the least effective mass transfer capability and are generally restricted to particulate matter removal and control of high-solubility gases such as sulfur dioxide and ammonia.⁴⁰ Venturi scrubbers have a high degree of gas/liquid mixing and provide high particulate matter removal efficiency. They also require high pressure drops (i.e., they are energy intensive) and have relatively short contact times. Therefore, their use is also restricted to high-solubility gases.⁴¹ As a result, VOC control by gas absorption is generally accomplished in packed or plate columns.

Packed towers are vertical columns containing inert packing, manufactured from materials such as porcelain, metal, or plastic, that provides the surface area for contact between the liquid and gas phases in the absorber. Packed columns are used primarily for corrosive materials and liquids with tendencies to foam or plug. They are less expensive than plate columns for small-scale or pilot plant operations where the column diameter is less than 0.6 m (2 ft). They are also suitable where the use of plate columns would result in excessive pressure drops.

In plate or tray columns, contact between the gas and liquid phases takes place in a stepwise fashion on a series of trays. The liquid phase flows down from tray to tray as the gas phase moves up through openings (e.g., perforations or bubble caps) in the tray, passing through the liquid along the way. These columns are preferred for large-scale operations, where internal cooling is desired or where low liquid flow rates would inadequately wet the packing.⁴²

Figure 2-10 is a schematic of a packed tower using countercurrent flow. The vent stream containing VOC's to be absorbed is introduced near the bottom of the tower (1) and allowed to rise through the packing material (2). Solvent flows in from the top of the column, countercurrent to the vapors (3), absorbing the solute from the gas phase and

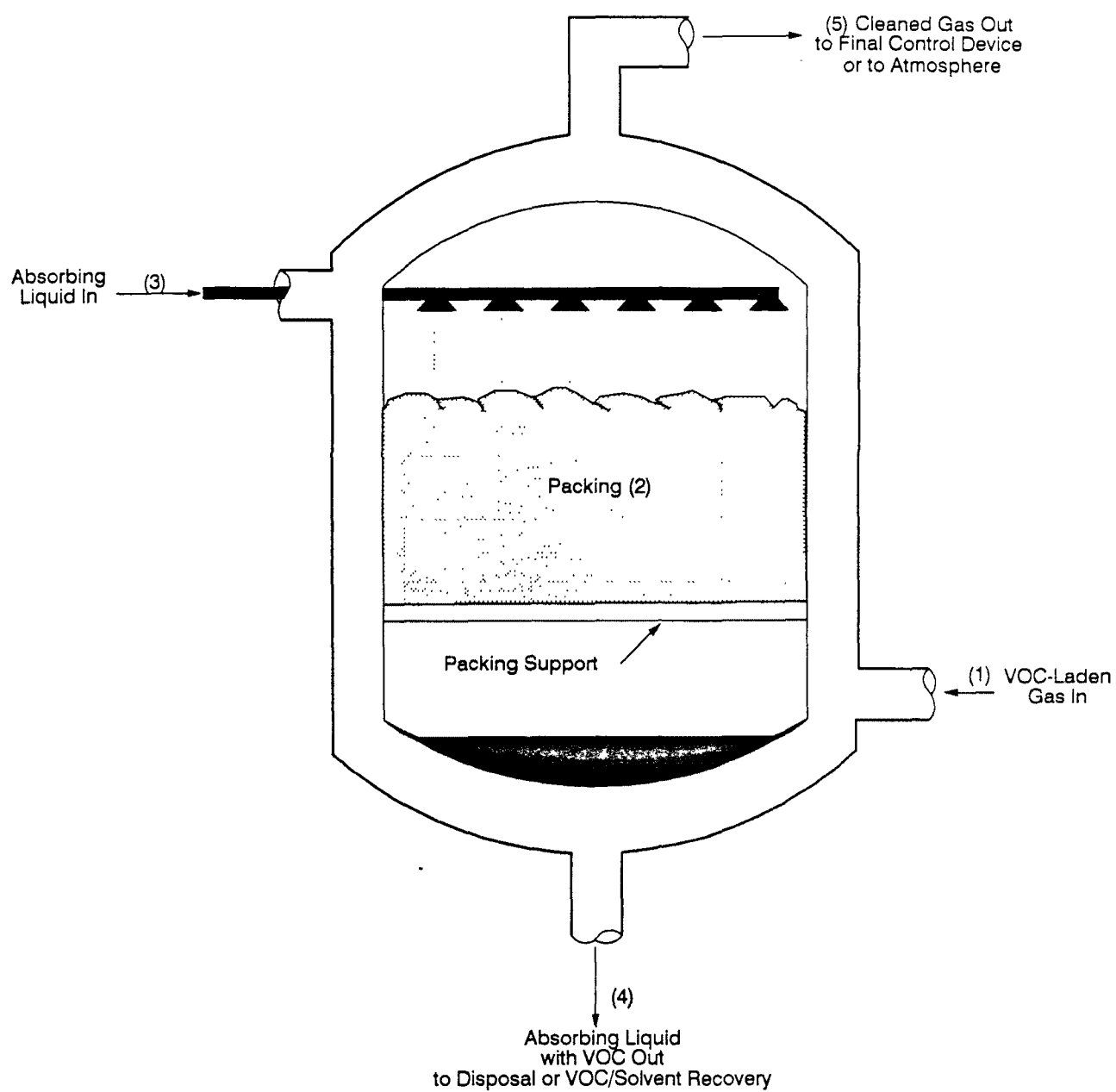


Figure 2-10. Packed tower absorption process.

carrying the dissolved solute out of the tower (4). The vent stream after treatment exits at the top (5) for release to the atmosphere or for further treatment as necessary. The saturated liquid may be sent to a stripping unit where the absorbed VOC is recovered. After the stripping operation the absorbing solution is either recycled to the absorber or sent to a water treatment facility for disposal.

The major tower design parameters to be determined for absorbing any substance are column diameter and height, system pressure drop, and liquid flow rate required. These parameters are derived by considering the solubility, viscosity, density, and concentration of the VOC in the inlet vent stream (all of which depend on column temperature); the total surface area provided by the tower packing material; and the quantity of gases to be treated.

2.2.5.2. Factors Affecting Absorption Control Efficiency. Control efficiencies for absorbers vary from 50 to 95 percent. The VOC removal efficiency of an absorber depends on the solvent selected and on proper design and operation. Absorbing liquids (solvents) are chosen for high solubility for the specific VOC and include liquids such as water, mineral oils, kerosenes, nonvolatile hydrocarbon oils, and aqueous solutions of oxidizing agents, sodium carbonate, and sodium hydroxide.⁴³ For a given solvent and solute, an increase in absorber size (i.e., contact surface area) or a decrease in the operating temperature can increase the VOC removal efficiency of the system. It may be possible in some cases to increase VOC removal efficiency by a change in the solvent.

2.2.5.3. Applicability of Absorption. The applicability of absorption for controlling VOC emissions is dependent primarily upon the availability of a suitable solvent.³⁵ Water may be used for the absorption of organic compounds that have relatively high water solubilities (e.g., most alcohols, organic acids, aldehydes, ketones, amines, and glycols). For organic compounds that have low water solubilities, other

solvents (usually organic liquids with low vapor pressures) are used.⁴⁴

Other factors bearing on the applicability of absorption as a viable emissions control option include the absorptive capacity and strippability of the VOC in the solvent. These factors are especially important to regenerable systems. Absorptive capacity is a measure of the solubility of the VOC in the solvent. The solubility limits the total quantity of VOC that could be absorbed in the system. Strippability relates to the ease with which the VOC can be removed from the solvent. If a VOC cannot be easily desorbed from the solvent, then absorption is a less viable control technique for control of the VOC.⁴⁵

The VOC concentration in the inlet vent stream also determines the applicability of absorption. Absorption is usually considered only when the VOC concentration is above 200 to 300 ppmv.⁴⁴ Below these gas-phase concentrations, the rate of mass transfer of the VOC to the solvent is decreased sufficiently to make reasonable designs infeasible.

2.3 STORAGE TANK IMPROVEMENTS FOR EMISSION REDUCTION

This section discusses two types of vessels commonly used for storage in the chemical manufacturing industry: fixed roof tanks and internal floating roof tanks.

Both types of tanks are cylindrical and have an axis perpendicular to the ground. Each tank has a cone- or dome-shaped roof that is permanently affixed to the tank shell. The internal floating roof tank also has a roof inside the tank that floats on the surface of the stored liquid. A conservation vent, which is a type of pressure- and vacuum-relief valve, is commonly installed on fixed roof tanks to control losses from the tank caused by minor changes of pressure in the vapor space.

The major types of emissions from fixed roof tanks are breathing and working losses. Breathing loss is the expulsion of vapor from a tank vapor space because of diurnal changes in temperature and barometric pressure. Emissions occur in the absence of any change in the liquid level in the tank.

Working losses occur during filling when the liquid level in the tank rises. The vapors are expelled from the tank when the pressure inside the tank exceeds the relief pressure as a result of filling.

Loss of VOC from internal floating roof tanks occurs in one of four ways:

- (1) Through the annular rim space around the perimeter of the floating roof (rim or seal losses);
- (2) Through the openings in the deck required for various types of fittings (fitting losses);
- (3) Through the nonwelded seams formed when joining sections of the deck material (deck seam losses); and
- (4) Through evaporation of liquid left on the tank wall following withdrawal of liquid from the tank (withdrawal loss).

As wind flows over the exterior of an internal floating roof tank, air flows into the enclosed space between the fixed and floating roofs through some of the shell vents and out of the enclosed space through others. Any VOC that has evaporated from the exposed liquid surface and that has not been contained by the floating deck will be swept out of the enclosed space.

Emissions from storage tanks can be reduced by reducing the surface area of liquid that is exposed to air. This can be achieved by: installing a floating roof inside a fixed roof tank, or upgrading an existing internal floating roof. The following sections discuss these tank improvements as emission control techniques. Their potential for emission reduction is compared to fixed roof tanks, which are considered the minimum acceptable equipment currently used for the storage of volatile organic liquids (VOL's).

2.3.1 Description of Tank Improvements

The presence of a floating roof (deck) inside a fixed roof tank significantly reduces the surface area of exposed liquid. The deck serves as a physical barrier between the VOL and the air that enters the tank through vents.

To retrofit a fixed roof tank with an internal floating deck, the tank must be cleaned and degassed and vents must be cut in the tank roof to minimize the possibility of hydrocarbon vapors accumulating in flammable concentrations. Installing sections of a welded deck in the tank will require cutting an opening in the tank shell larger than any standard manway.

Upgrading an existing floating roof involves cleaning and degassing the tank and replacing some parts of the deck (e.g., rim seals and deck fittings) with parts that allow fewer emissions. For example, uncontrolled deck fittings (Table 2-1) are replaced with controlled fittings. Liquid-mounted rim seals may be substituted for vapor-mounted seals. Also, a secondary rim seal may be added to supplement the control achieved by the primary seal. Access to the tank for upgrading an existing internal floating roof is usually made through an existing manway.

2.3.2 Factors Affecting Control Efficiency

Because evaporation is the primary emission mechanism associated with storage tanks, emissions from both fixed roof and floating roof tanks vary with the vapor pressure of the stored liquid. Thus, the control efficiency of retrofitting a fixed roof tank with an internal floating deck depends upon the material being stored.

Other factors affecting emissions, and therefore control efficiency, are tank size, number of turnovers, and the type of deck and seal system selected. Installing an internal floating roof can reduce emissions by 76 to 98 percent.⁴⁶ The relative effectiveness of one internal floating roof design over another is a function of how well the deck can be sealed.

To simplify the discussion, an example tank is used as a common basis for evaluating effectiveness. The example tank has the following characteristics:

- Tank diameter--85 ft;
- Tank height--47 ft;
- Tank capacity--2,000,000 gal;
- Vapor pressure of the stored liquid--0.71 psia;

TABLE 2-1. CONTROLLED AND UNCONTROLLED INTERNAL
FLOATING ROOF DECK FITTINGS

Deck Fitting Type	Equipment Description	
	Uncontrolled	Controlled
1. Access hatch	Unbolted, ungasketed cover ^a ; or unbolted, gasketed cover	Bolted, gasketed cover
2. Automatic gauge float well	Unbolted, ungasketed cover ^a ; or unbolted, gasketed cover	Bolted, gasketed cover
3. Column well	Built-up column-sliding cover, ungasketed ^a ; pipe column-sliding cover, ungasketed	Flexible fabric sleeve seal or a gasketed sliding cover
4. Ladder well	Ungasketed sliding cover ^a	Gasketed sliding cover
5. Sample pipe or well	Sample well with slit fabric seal, 10% open area ^a	Sample well with slit fabric seal, 10% open area ^a
6. Vacuum breaker	Weighted mechanical actuation, ungasketed ^a	Weighted mechanical actuation, gasketed

^aFittings assumed in the uncontrolled case for estimating the effectiveness of fittings controls. This scenario is representative of no single tank, but rather is the composite of what is estimated based on a survey of users and manufacturers to be typical of fittings on the majority of tanks currently in service. Note that the sample well with split fabric seal was used in the uncontrolled case for calculating emissions because it is in common use. It was also used in the controlled case because it is the lowest emitting fitting type.

- Density of stored liquid--7.22 lb/gal;
- Molecular weight of the product vapor--92.13 lb/lb-mole; and
- Turnover rate--seven per year.

Probably the most typical internal floating roof design is the noncontact, bolted, aluminum internal floating roof with a single vapor-mounted wiper seal and uncontrolled fittings. As discussed above, there are four types of losses from this roof design. Contributions made by the four types of losses to the total loss from the example tank are estimated as follows:

- (1) Rim or seal losses: 33 percent;
- (2) Fitting losses: 35 percent;
- (3) Deck seam losses: 30 percent; and
- (4) Withdrawal losses: 2 percent.

With the exception of withdrawal losses, which are inherent in all internal floating roof designs, the losses listed above can be reduced by employing roofs with alternative design features. Table 2-2 lists alternative floating roof equipment designs and the emission rate associated with each type of equipment installed on the example tank. The following sections, organized according to types of losses, elaborate on the alternative equipment that can be employed on internal floating roofs.

2.3.2.1 Control Of Seal losses. Internal floating roof seal losses can be minimized by employing liquid-mounted primary seals instead of vapor-mounted seals and/or by employing secondary wiper seals in addition to primary seals.

Available emissions test data suggest that the location of the seal (i.e., vapor- or liquid-mounted) and the presence of a secondary seal are the major factors affecting seal losses. A liquid-mounted primary seal has a lower emissions rate, and thus a higher control efficiency, than a vapor-mounted seal. Equipping the example tank with an internal floating roof having a liquid-mounted rim seal results in an emission reduction of 96.0 percent over a fixed roof tank; whereas a deck with a vapor-mounted seal achieves only

TABLE 2-2. EFFECTIVENESS OF INTERNAL FLOATING ROOFS
ON AN EXAMPLE TANK^a

Example Fixed Roof Tank	Case	Equipment Type ^{b,c}	Internal Floating Roof Tank	
			Total Emission Rate ^d (Mg/yr)	Reduction Over Fixed Roof Tank Emission Rate (%)
Total emission rate 17.78 Mg/yr (working loss = 9.90) (breathing loss = 7.88)	1	bIFR _{vm}	0.88	95.1
	2	bIFR _{vm,cf}	0.76	95.7
	3	bIFR _{vm,cf,ss}	0.58	96.7
	4	bIFR _{lm}	0.71	96.0
	5	bIFR _{lm,ss}	0.65	96.3
	6	bIFR _{lm,cf}	0.60	96.6
	7	bIFR _{lm,cf,ss}	0.54	97.0
	8	wIFR _{lm,cf,ss}	0.29	98.4

^aExample tank is 2,000,000-gal capacity; 85 ft in diameter, 47 ft in height, 0.71 psia vapor pressure, 92.13 lb/lb-mole molecular weight of product and condensed product vapor, and seven turnovers per year.

^bKey:

b = bolted roof deck; w = welded roof deck; IFR = internal floating roof; v = vapor-mounted primary seal; m = metallic shoe primary seal; l = liquid-mounted primary seal; cf = controlled fittings; ss = rim-mounted secondary seal.

^cUncontrolled fittings (as designated by absence of cf):

- (1) access hatch, with ungasketed, unbolted cover;
- (2) automatic gauge float well, with ungasketed, unbolted cover;
- (3) built-up column well, with ungasketed sliding cover;
- (4) ladder well, with ungasketed sliding cover;
- (5) adjustable roof legs;
- (6) sample well with slit fabric seal (10 percent open area);
- (7) 1-in. diameter stub drains; and
- (8) vacuum breaker with gasketed weighted mechanical actuation.

TABLE 2-2. EFFECTIVENESS OF INTERNAL FLOATING ROOFS
ON AN EXAMPLE TANK^a (CONCLUDED)

Controlled fittings (cf): (1) access hatch, with gasketed, bolted cover; (2) automatic gauge float well, with gasketed, bolted cover; (3) built-up column well, with gasketed, sliding cover; (4) ladder well, with gasketed sliding cover; (5) adjustable roof legs; (6) sample well with slit fabric seal (10 percent open area); (7) 1-in. diameter stub drains; and (8) vacuum breaker, with gasketed weighted mechanical actuation.

^dReference 47.⁴⁷

95.1 percent reduction (Table 2-2). A secondary seal, with either a liquid- or a vapor-mounted primary seal, provides an additional level of control.

The type of seal used plays a less significant role in determining the emissions rate.⁴⁸ The type of seal is important only to the extent that the seal must be suitable for the particular application. For instance, an elastomeric wiper seal is commonly employed as a vapor-mounted primary seal or as a secondary seal for an internal floating roof. Because of its shape, this seal is not suitable for use as a liquid-mounted primary seal. Resilient filled seals, on the other hand, can be used as both liquid- and vapor-mounted seals.

The effectiveness of alternate internal floating roof seal systems can be evaluated by inspecting the rim seal loss factors (K_R). These factors have been developed based on test data for estimating losses for various seal systems and are listed in Table 2-3 for seals with average gaps.⁴⁹ Estimates of control efficiency and incremental control efficiency are also listed in the table. The control efficiency estimates indicate the effectiveness of the various seal systems at reducing emissions over the level achieved by a vapor-mounted primary seal. In Table 2-3, the vapor-mounted primary seal is assumed to be the baseline control level. The incremental control efficiency estimates demonstrate the effectiveness of each seal system relative to the next less stringent seal system (i.e., the next higher emitting seal system). These efficiencies are calculated directly from the K_R values.

As shown in the table, application of a liquid-mounted primary and secondary seal system in place of a vapor-mounted primary seal would reduce seal losses an estimated 76 percent. On the example tank, where these seal losses represent roughly one-third of the total loss from the tank (i.e., $b_{IFR_{VM}}$ case), this 76 percent reduction in seal losses translates to a 25 percent reduction in the total loss from the floating roof tank. Relative to fixed roof tank emissions, the additional control provided by the liquid-mounted primary and secondary

TABLE 2-3. INTERNAL FLOATING ROOF RIM SEAL SYSTEMS,
SEAL LOSS FACTORS, AND CONTROL EFFICIENCIES

Seal System	K_r^a (lb-mole/ft-yr)	Seal Loss Control Efficiency Related To Baseline	Incremental Control Efficiency
Vapor-mounted primary seal only	6.7	IFR ^b baseline (0%)	--
Liquid-mounted primary seal only	3.0	55%	55%
Vapor-mounted primary and secondary seals	2.5	63%	17%
Liquid-mounted primary and secondary seals	1.6	76%	36%

^a K_r = Rim seal loss factors.

^bIFR = Internal floating roof type.

seal system over the vapor-mounted primary seal system increases the effectiveness of the internal floating roof from 95.1 percent to 96.3 percent for the example tank. (See Case 1 vs. Case 5 in Table 2-2.)

2.3.2.2 Control Of Fitting Losses. The numerous fittings that penetrate or are attached to an internal floating roof include access hatches, column wells, roof legs, sample pipes, ladder wells, vacuum breakers, and automatic gauge float wells. Fitting losses occur when VOC's leak around these fittings. Fitting losses can be controlled with gasketing and sealing techniques or by the substitution of fittings that are designed to leak less. Table 2-1 lists the fitting types considered representative of uncontrolled and controlled fittings.

The effectiveness of fitting controls at reducing the overall emission rate is a function of the number of fittings of each type employed on a given tank. On the example tank, use of controlled fittings reduces the total fitting loss by about 36 percent. Because fitting losses are about 35 percent of the total emissions from the example internal floating roof tank (i.e., for an IFR_{vm}), the controlled fittings reduce the overall emissions by about 13 percent over a similar tank without fitting controls. The additional emission reduction achieved by installing controlled fittings increases the control efficiency of the example internal floating roof tank from 95.1 percent to 95.7 percent over a fixed roof tank as the base case.

2.3.2.3 Control Of Deck Seam Losses. Depending on the type of floating roof used, deck seam losses can contribute to the total loss from an internal floating roof. For the example tank used as a basis for comparison throughout this section (i.e., $bIFR_{vm}$), deck seam losses are 30 percent of the total loss. When seal losses and fitting losses are controlled, the relative contribution to the total loss from deck seams increases. In the case of a bolted, noncontact, internal floating roof with liquid-mounted primary seals, controlled deck fittings, and secondary seals ($bIFR_{lm,cf,ss}$),

deck seam losses contribute about 47 percent of the total loss.

Deck seam losses are inherent in several floating roof types. Any roof constructed of sheets or panels fastened by mechanical fasteners (e.g., bolts) is expected to have deck seam losses. Two roof types were tested to determine deck seam losses.⁴⁹ The first was a bolted, aluminum, noncontact roof and the second was a bolted, aluminum panel-type, contact roof. The design of the mechanical fasteners used on these two roof types varies significantly. In addition, one roof type floats above the liquid surface while the other floats in contact with the liquid surface. Despite these differences, the seams on these two roof types were found to emit at roughly the same rate per meter of seam. Deck seam losses, therefore, are considered to be a function of the length of the seams and not the type of mechanical fastener or the position of the deck relative to the liquid surface.

Deck seam losses are controlled by selecting a roof type with vapor-tight deck seams. The welded deck seams on steel pan roofs are vapor tight. Fiberglass-lapped seams of a glass fiber reinforced polyester roof may be vapor tight as long as there is negligible permeability of the liquid through the seam-lapping materials. Some manufacturers provide gaskets for bolted metal deck seams. One vendor of contact, aluminum bolted decks has provided bench-scale test results that substantiate emissions reduction due to gasketing deck seams.⁵⁰

Selecting a welded roof (rather than a bolted roof) will eliminate deck seam losses. On an internal floating roof with liquid-mounted primary seal, secondary seal, and controlled fittings, the elimination of deck seam losses improves the overall effectiveness for the example tank from a 97.0 to 98.4 percent control efficiency (see Case 7 vs. Case 8 in Table 2-2), relative to a fixed roof tank.

2.3.3 Applicability of Storage Tank Improvements

The applicability of storage tank improvements as a technology to reduce VOC emissions is dependent upon

characteristics of the particular VOC. Because floating decks are often constructed of aluminum panels, this control technology may not be applicable to tanks storing halogenated compounds, pesticides, or other compounds that are incompatible with aluminum. Contact between these compounds and an aluminum deck could corrode the deck and cause product contamination.

In addition, chemical vapor pressure may affect the selection of tank improvements as an applicable control technology. For chemicals with very low vapor pressure, fixed roof tank emissions will already be so low that installing an internal floating roof may not significantly reduce emissions further. (This lower vapor pressure limit cannot be specified because the equations for calculating emissions from internal floating roof tanks are only applicable for vapor pressures above 0.1 psia). For chemicals with medium vapor pressures (up to about 9 psia), emission reductions of 95 percent and above are achievable with this technology. However, for chemicals with higher vapor pressures, achievable emission reduction starts to decrease with increasing vapor pressure. Thus, tank improvements may not have acceptable emission reduction efficiency for chemicals with high vapor pressures.

2.4 EQUIPMENT LEAK EMISSION SOURCES AND EMISSION CONTROL TECHNIQUES

Equipment leak emissions in the chemical manufacturing industries occur when process fluid (either liquid or gaseous) is released through the sealing mechanisms of equipment in the plant. This section discusses the sources of equipment leak emissions and control techniques that can be applied to reduce emissions from equipment leaks, including the applicability of each control technique and its associated effectiveness in reducing emissions.

There are many potential sources of equipment leak emissions in an organic chemical processing unit. The following sources of equipment leaks are considered: pumps, compressors, agitators, pressure relief devices, open-ended

lines, sampling connections, process valves, connectors, instrumentation systems, and product accumulator vessels.

Emissions from equipment leaks have been the subject of intense study by the EPA and industry. Emissions from the populations of equipment leak sources have been found to be random occurrences, when considering individual pieces of equipment. Further, no relationship between emissions and process variables, such as line temperatures and pressure, has been established. Precise quantitation of emission reduction effectiveness for the techniques described in this section has not been made on a unit-by-unit basis, but approximate quantitation has been made on an industry-wide basis for certain emission reduction techniques. Many of these were described in detail in the EPA's additional information document on emissions, emissions reductions, and costs of controlling emissions from equipment leaks.⁵¹

The techniques for reducing emissions from equipment leaks are as diverse as the types of sources. These techniques can be classified in the following major categories:

- Equipment (modifications and leakless equipment);
- Closed vent systems; and
- Work practices.

The selection of a control technique and its effectiveness in reducing emissions depends on a number of factors, including:

- Type of equipment (valve, pump, compressor, etc.);
- Equipment service (gas, light liquid, heavy liquid);
- Process variables that limit equipment selection (temperature, pressure);
- Process stream composition; and
- Costs.

The various control techniques that can be applied to the different types of equipment components and their estimated potential emission reductions are presented in the following sections. Section 2.4.1 describes the various equipment components that can be sources of emissions and the equipment modifications and alternative equipment that can be applied to

reduce emissions from these equipment components. Closed vent systems and work practices are described in Sections 2.4.2 and 2.4.3, respectively. Both closed vent systems and certain work practices can be applied to a number of different equipment types and therefore are described in more general terms.

2.4.1 Equipment Description and Specifications

Equipment leaks may be reduced by retrofitting with equipment designed to reduce or prevent leakage. Equipment modifications and leakless equipment for each equipment type are described below. Some of the modifications may be applicable to more than one type of equipment.

2.4.1.1 Pumps. Pumps are used extensively in the chemical industries for the movement of organic liquids.⁵² The centrifugal pump is the most widely used pump in the chemical industries; however, other types, such as the positive displacement (reciprocating) pump, are also used. Chemicals transferred by pump can leak at the point of contact between the moving shaft and the stationary casing. Consequently, all pumps except the sealless type, such as canned-motor, magnetic drive, and diaphragm pumps, require a seal at the point where the shaft penetrates the housing in order to isolate the pumped fluid from the environment.

2.4.1.1.1 Seals for pumps. Two generic types of seals, packed and mechanical, are used on pumps. Packed seals can be used on both reciprocating and rotary action (centrifugal) pumps. A packed seal consists of a cavity ("stuffing box") in the pump casing filled with packing material that is compressed with a packing gland to form a seal around the shaft. Coolant is required to remove the frictional heat between the packing and shaft. The necessary lubrication is provided by a coolant that flows between the packing and the shaft.⁵² Deterioration of the packing can result in leakage of process liquid.

Mechanical seals are limited in application to pumps with rotating shafts. There are single and double mechanical seals, with many variations to their basic design, but all

have a lapped seal face between a stationary element and a rotating seal ring.⁵³ The rotating seal ring and stationary element faces are lapped to a very high degree of flatness to maintain contact throughout the entire mutual surface area. In a single mechanical seal, the faces are held together by the pressure applied by a spring on the drive and by the pump pressure transmitted through the pumped fluid on the pump end. An elastomer O-ring seals the rotating face to the shaft. The stationary face is sealed to the stuffing box with another elastomer O-ring or gasket.

For double mechanical seals, two seals are arranged back-to-back, in tandem, or face to face. In the back-to-back arrangement, a closed cavity is created between the two seals. A seal liquid, such as water or seal oil, is circulated through the cavity. Because the seal liquid surrounds the two seals, it can be used to control the temperature in the stuffing box. For the seal to function properly, the pressure of the seal liquid must be greater than the operating pressure of the pump. In this manner, any leakage would occur across the seal faces into the process or the environment. Therefore, the seal liquid must be compatible with the process fluid and the environment.⁵⁴

In a tandem dual mechanical seal arrangement, the seals face the same direction. The secondary seal backs up the primary seal. The cavity between the two seals is filled with a barrier fluid, which may serve to control temperature of the seals. In the tandem arrangement, the barrier fluid may be maintained at a pressure lower than that in the stuffing box, allowing any leakage through the seal from the process into the barrier fluid. To reduce emissions from such an arrangement, a barrier fluid reservoir system should be used. At the reservoir, the process liquid can vaporize (i.e., degas) and be released to the atmosphere, or to a control device for proper recovery or destruction.⁵⁴

In the face-to-face arrangement, two rotating faces are mated with a common stationary element. The barrier fluid in the cavity between the seals may be maintained at higher or

lower pressures than the stuffing box. As in the tandem arrangement, leakage from the process fluid to the barrier fluid can occur if the barrier fluid is maintained at a pressure lower than that in the stuffing box. Preventing emissions from the barrier fluid requires (1) operation of the barrier fluid at a pressure higher than that of the stuffing box or (2) use of a degassing reservoirs vented to a control device.

The actual emission reduction achievable through use of double mechanical seals depends on the frequency of seal failure. Failure of both the inner and outer seals can result in relatively large emissions at the seal area of the pump. Pressure monitoring of the barrier fluid may be used to detect failure of the seals.⁵⁵ Visual inspection of the seal area can also detect failure of the outer seals. Seal failure would require the leaking pump to be shut down for maintenance.

Double mechanical seals are used in many process applications; however, there are some conditions that preclude their use. Their maximum service temperature is usually less than 260 °C, and mechanical seals can rarely be used successfully on pumps with reciprocating shaft motion. Further, double mechanical seals cannot be used where the process fluid contains slurries, polymeric, or undissolved solids.

2.4.1.1.2 Sealless pumps. Another type of pump used in the chemical industry is the sealless pump (i.e., canned-motor, diaphragm, and magnetic drive pumps). Sealless pumps are used primarily in processes where the pumped fluid is hazardous, highly toxic, or very expensive and where every effort must be made to prevent all possible leakage of the fluid. Canned-motor pumps have interconnected cavity housings, motor rotors, and pump casings. As a result, the motor bearings run in the process liquid and all seals are eliminated. Because the process liquid is the bearing lubricant, abrasive solids in the process lines cannot be tolerated. Canned-motor pumps are widely used for handling

organic solvents, organic heat transfer liquids, and light oils.

Diaphragm pumps perform similarly to piston and plunger pumps. However, the driving member is a flexible diaphragm of metal, rubber, or plastic. The primary advantage of this arrangement is the elimination of all packing and seals exposed to the process liquid provided the diaphragm's integrity is maintained. This is important when handling hazardous or toxic liquids. Emissions from diaphragm pumps can be large, however, if the diaphragm fails.

In magnetic-drive pumps, no seals contact the process fluid. An externally-mounted magnet coupled to the pump motor drives the impeller in the pump casing.

2.4.1.2 Compressors. Compressors provide motive force for transporting gases through a process unit in much the same way that pumps transport liquids. Compressors are typically driven with rotating or reciprocating shafts. Thus, the sealing mechanisms for compressors are similar to those for pumps, that is, packed and mechanical seals. Emissions from this source type may be reduced by improving the seals' performance or by collecting and controlling the emissions from the seal. Emissions from mechanical contact seals depend on the type of seal or control device used and the frequency of seal failure.

Shaft seals for compressors are of several different types: labyrinth, restrictive carbon rings, mechanical contact, and liquid film. All of these seal types restrict leaks, although none of them completely eliminates leakage. Compressors can be equipped with ports in the seal area to evacuate collected gases, which could then be controlled.

The labyrinth compressor seal is composed of a series of close-tolerance, interlocking "teeth," which restrict the flow of gas along the shaft. Labyrinth seals have the largest leak potential of the different seal types. Many variations in "tooth" design and materials of construction are available to improve compressor seal performance. Properly applied "tooth"

configuration and shape can reduce leakage by up to 40 percent over a straight pass labyrinth seal.⁵⁶

Restrictive carbon ring seals consist of multiple stationary carbon rings with close shaft clearances. This type of seal may be operated dry, with a sealing fluid, or with a buffer gas. Restrictive ring seals can achieve lower leak rates than labyrinth seals.⁵⁶

Mechanical contact seals are similar to the mechanical seals described for pumps. Clearance between the rotating and stationary elements is reduced to essentially zero. Oil or another suitable lubricant is supplied to the seal faces. Mechanical contact seals can achieve the lowest emission rates of the types of compressor seals described here. Like mechanical seals for pumps, however, they are not suitable for all processing conditions.⁵⁶

A buffer or barrier fluid may be used with these mechanical seals to form a buffer between the compressed gas and the environment, similar to barrier fluids in pumps. This system requires a clean, external gas supply that is compatible with the gas being compressed. Barrier gas can become contaminated and must be disposed of properly, for example by venting to a control device.

Centrifugal compressors also can be equipped with liquid film seals. The seal is formed by a film of oil between the rotating shaft and stationary gland. Process gas can be released to the atmosphere when the circulating oil is returned to the oil reservoir.⁵⁷ To eliminate release of volatile air pollutants from the seal oil system, the reservoir can be vented to a control device.

2.4.1.3 Agitators. Agitators are used in the chemical industries to stir or blend chemicals. As with pumps and compressors, emissions from agitators can occur at the interface of a moving shaft and a stationary casing. Emissions from this source type may be reduced by improving the seal or by collecting and controlling emissions. Four seal arrangements are commonly used with agitators: packed seals, mechanical seals, hydraulic seals, and lip seals.

Packed seals for agitators are very similar in design and application to the packed seals for pumps (Section 2.4.1.1).

Although mechanical seals are more costly than other seal arrangements, they provide better leakage rate reduction. Also, the maintenance frequency of properly installed and maintained mechanical seals is one-half to one-fourth that of packed seals.⁵⁸ In fact, at pressures greater than 1,140 kPa (165 psia), packed seals are rarely used because the performance of mechanical seals is superior.⁵⁹ Mechanical seals can be designed specifically for these higher pressure applications. As with packed seals, the mechanical seals for agitators are similar to the design and application of mechanical seals for pumps.

The hydraulic seal is the simplest and least-used agitator shaft seal. In this type of seal, an annular cup attached to the process vessel contains a liquid that contacts an inverted cup attached to the rotating agitator shaft. The primary advantage of this seal is that it is a noncontact seal. However, this seal is limited to low temperatures and pressures and can only handle very small pressure fluctuations. Process chemicals may contaminate the seal liquid and then be released into the atmosphere as equipment leak emissions.

Lip seals, which are relatively inexpensive and easy to install, can be used on a top-entering agitator as a dust or vapor seal. The sealing element in the lip seal is a spring-loaded elastomer. Once the seal has been installed, the agitator shaft rotates in continuous contact with the lip seal. Pressure limitations of the seal are 14 to 21 kPa (2 to 3 psi) because it operates without lubrication. Operating temperatures are limited by the characteristics of the selected elastomer. Emissions can be released through this seal when it wears excessively or when the operating pressure surpasses the pressure limitation of the seal.

2.4.1.4 Pressure Relief Devices. Insurance, safety, and engineering codes require that pressure relief devices or systems be used in applications where the process pressure may

exceed the maximum allowable working pressure of the process equipment. Pressure relief devices include rupture disks and safety/relief valves. The most common pressure relief device is a spring-loaded valve designed to open when the operating pressure of a piece of process equipment exceeds a set pressure. In this example, the pressure relief valve is constructed so that it will close or reseal after the operating pressure has decreased to a level below the set pressure. Equipment leak emissions from spring-loaded relief valves may be caused by failure of the valve seat or valve stem, improper reseating after overpressure relief, or process operation near the relief valve set pressure which may cause the relief valve to frequently open and close or "simmer."

Manufacturers of relief valves state that resilient seat or "O-ring" relief valves provide better reseal qualities after an overpressure relief compared to standard relief valves. The applicability of resilient seat technology is limited by material compatibility and operating conditions. This technology would have no effect on emissions from overpressure episodes, emissions due to failure of the seat surfaces, or "simmering."

Rupture disks are designed to burst at overpressure to allow the process gas to vent directly to the atmosphere. Rupture disks allow no emissions as long as the integrity of the disk is maintained. The rupture disk must be replaced after each pressure relief episode to restore the process to an operating pressure condition. Although rupture disks can be used alone, they are sometimes installed upstream of a relief valve to prevent emissions through the relief valve seat.

Rupture disk/relief valve combinations require certain design constraints and criteria to avoid potential safety hazards. For example, appropriate piping changes must be made to prevent disk fragments from lodging in and damaging the relief valve when relieving overpressure. A block valve upstream of the rupture disk can be used to isolate the rupture disk/relief valve combination and permit in-service

replacement of the disk after it bursts. Otherwise, emissions could result through the relief valve.

A dual-pressure relief system with a three-way valve can be used to maintain operation while having continuous overpressure protection. The three-way valve can be open to either one of the dual rupture disk (RD)/pressure relief valves (PRVs) so that one RD/PRV combination is always in service. If one of the rupture disks fails, it can be made available for repair by switching the three-way valve to the other RD/PRV combination. Rupture disk/relief valve combinations must have some provision for testing the integrity of the disk. The American Petroleum Institute (API) has provided guidance on the design of instrumentation to indicate failure of the rupture disk. According to API guidelines, the area between the rupture disk and relief valve should be connected to a pressure indicator, recorder, or alarm.⁶⁰ The control efficiency of the disk/valve combination is assumed to be 100 percent when operated and maintained properly. If disk integrity is not maintained or if the disk is not replaced after overpressure relief, the control efficiency is lowered, but this reduction in efficiency has not been quantified.

2.4.1.5 Open-Ended Lines. Emissions from open-ended lines are caused by leakage through the seat of an upstream valve in the open-ended line. Emissions that occur through the stem and gland of the valve are not considered "open-ended" emissions and are addressed in Section 2.4.1.7. Emissions from open-ended lines can be controlled by installing a cap, plug, flange, or second valve to the open end. The control efficiency of these control measures is assumed to be 100 percent. In the case of a second valve, the upstream valve should always be closed first after use of the valves to prevent the trapping of fluids between the valves. Each time the cap, plug, flange, or second valve is opened, any VOC that has leaked through the first valve seal will be released.

2.4.1.6 Sampling Connections. Emissions from sampling connections occur as a result of purging the sampling line to obtain a representative sample of the process fluid. Based on survey data on the SOCFI, approximately 25 percent of the valves on open-ended lines are used for sampling connections.⁶¹ Emissions from sampling connections can be reduced by using a closed loop sampling system or disposing of the purged process fluid in a control device. The closed loop sampling system is designed to return the purged fluid to the process at a point of lower pressure. A throttle valve or other device is used to induce the pressure drop across the sample loop. Closed loop sampling is assumed to be 100 percent effective for controlling emissions from sample purge. The purged fluid could also be directed to a control device such as an incinerator, in which case the control efficiency would depend on the efficiency of the incinerator in removing the VOC. Because some pressure drop is required to purge sample through the loop, low pressure processes or storage tanks may not be amenable to closed loop sampling. Also, safety requirements may prohibit closed loop sampling in some instances.

2.4.1.7 Process Valves. Valves are the most common and numerous process equipment type found in the chemical industry. Valves are available in many designs, most of which contain a valve stem that operates to restrict or allow fluid flow. Typically, the stem is sealed by a packing gland or O-ring to prevent leakage of process fluid to the atmosphere. Emissions from valves occur at the stem or gland area of the valve body when the packing or O-ring in the valve fails.

2.4.1.7.1 Seals for valves. Valves that require the stem to move in and out or turn must utilize a packing gland. A variety of packing materials are suitable for conventional packing glands. The most common packing materials are the various types of braided asbestos that contain lubricants; other packing materials include graphite, graphite-impregnated fibers, and tetrafluorethylene. The choice of packing material depends on the valve application and configuration.⁶²

Conventional packing glands can be used over a wide range of operating temperatures. At high pressures, these glands must be quite tight to attain a good seal.⁶³

Elastomeric O-rings are also used for sealing process valves, but are not suitable where there is sliding motion through the packing gland. These seals are rarely used in high-pressure service, and operating temperatures are limited by the seal material.⁶⁴

2.4.1.7.2 Sealless valves. Emissions from process valves can be eliminated if the valve stem can be isolated from the process fluid. Two types of leakless or sealless valves are available: diaphragm valves and sealed bellows valves.

Diaphragm valves isolate the valve stem from the process fluid using a flexible elastomer or metal diaphragm. The position of the diaphragm is regulated by a plunger, which is controlled by the stem. The stem may be actuated manually or automatically by standard hydraulic, pneumatic, or electric actuators. In this arrangement, the stem/plunger pushes the diaphragm toward the valve bottom, throttling the flow of the process fluid. The diaphragm and stem/plunger are connected so that it is impossible for them to be separated under normal working conditions. When the diaphragm reaches the bottom of the valve, it is seated firmly, forming a leak-proof seal. This configuration is recommended for fluids containing solid particles and for medium-pressure service. Depending on the diaphragm material, this type of valve can be used at temperatures as high as 205 °C and in strong acid service. If the diaphragm fails, the valve can become a relatively larger source of emissions.⁶⁵

Diaphragm valves are very corrosion resistant and have reportedly performed well as control valves with minimum maintenance. The design problems associated with these valves are the temperature and pressure limitations of the elastomer used for the diaphragm. Use at temperatures beyond the operating limits of the material tends to damage or destroy the diaphragm. Also, constraints on operating pressure may

limit the application of diaphragm valves to medium- and low-pressure operations.

Sealed bellows valves are another alternative leakless design for isolating the valve stem from the process fluid. In this type of valve, metal bellows are welded to the bonnet and disk of the valve, thereby isolating the stem from the process. These valves can be designed to withstand high temperatures and pressures and can provide leak-free service at operating conditions beyond the limits of diaphragm valves. However, they are usually dedicated to highly toxic services and the nuclear industry. Further, they are only available in globe and gate valve configurations, and the bellows alloy must be carefully selected to prevent corrosion of the crevices of the bellows under severe conditions.

The control effectiveness of both diaphragm and sealed bellows valves is virtually 100 percent, although a failure of the diaphragm or bellows could cause temporary emissions much larger than those from other types of valves.

2.4.1.8 Connectors. Flanges, threaded fittings, and other fittings used to join sections of piping and equipment are connectors. They are used wherever pipe or other equipment (such as vessels, pumps, valves, and heat exchangers) require isolation or removal.

Flanges are bolted, gasket-sealed connectors. Normally, flanges are used for pipes with diameters of 50 mm or greater and are classified by pressure rating and face type. The primary causes of flange leakage are poor installation and thermal stress, which results in the deformation of the seal between the flange faces.⁶⁶ Flanges can become sources of emissions when they are assembled poorly or when replacement gaskets are improperly chosen.

Threaded fittings are made by cutting threads into the outside end of one piece (male) and the inside end of another piece (female). These male and female parts are then screwed together like a nut and bolt. Threaded fittings are normally used to connect piping and equipment having diameters of 50 mm or less. Seals for threaded fittings are made by coating the

male threads with a sealant (e.g., pipe dope) before joining it to the female piece. The sealant may be a polymeric tape, brush-on paste, or other spreadable material that acts like glue in the joint. These sealants typically need to be replaced each time the joint is broken. Emissions can occur as the sealant ages and eventually cracks. Leakage can also occur as the result of poor assembly or application of the sealant, and thermal stress of the piping and fittings.

Emissions from connectors can be controlled by regularly scheduled maintenance. Potential emissions can be reduced by replacing the gasket or sealant materials. Because most connectors cannot be isolated for maintenance or replacement during process operation, any maintenance must occur during shut-downs. In cases where connectors are not required for process modification or periodic equipment removal, emissions from connectors can be eliminated by welding the connectors together.

2.4.1.9 Instrumentation Systems. An instrumentation system is a group of equipment components used to condition and convey a sample of process fluid to analyzers and instruments for the purpose of determining process operating conditions (e.g., composition, pressure, and flow rate). Valves and connectors are the predominant types of equipment used in instrumentation systems, although other equipment may be included. For the purposes of identification, only valves nominally 12.7 mm and smaller and connectors nominally 19 mm and smaller are typically considered instrumentation systems. Emissions resulting from the components in the instrumentation system are controlled as they are for the same component in the process system.

2.4.1.10 Product Accumulator Vessels. Product accumulator vessels are small, primarily fixed roof storage tanks designed to regulate material flow through a process. They include the overhead and bottoms receiver vessels used with fractionation columns and the product separator vessels used in series with reactor vessels to separate reaction products. Under normal operating conditions, the amount of

material entering and exiting the vessel is the same and the net flux in the vessel is zero. Emissions occur when VOC's are vented to the atmosphere either directly or through a blowdown drum or vacuum system. Control of emissions from product accumulator vessels require the capture and transportation of VOC's through a closed vent system to a control device.

2.4.2 Closed Vent Systems and Control Devices

Emissions from equipment leaks may be controlled by installing a closed vent system around the leaking equipment and venting the emissions to a control device. This method of control is only applicable to certain equipment types (e.g., pumps, compressors, agitators, pressure relief valves, and product accumulator vessels). Because of the many valves, connectors, and open-ended lines typically found in chemical facilities, it is not practical to use this technique for reducing emissions from all of these potential sources for an entire process unit. However, a closed vent system can be used to control emissions from a limited number of components, which could be enclosed and maintained under negative pressure and vented to a control device.

For pumps, compressors, agitators, and product accumulator vessels, a closed vent system would require some type of flow-inducing device to transport the emissions from the seal area of the piece of equipment to the control device. The seal area must be enclosed to collect the emissions, and a vacuum eductor or a compressor could be used to remove vapors from the seal area. However, normal operating practices for some of this equipment may require frequent visual inspection or mechanical adjustment in the seal area. Also, explosive mixtures may be created by enclosing some seal areas. Therefore, safety and operating practices limit the use of closed vent streams in some applications.

2.4.2.1 Combustion Control Devices. Combustion control devices that can be used to control VOC emissions collected in closed vent systems include incinerators, industrial boilers, process heaters, and flares. Control efficiencies of the

various techniques depend on operating characteristics and the properties of the VOC being controlled.

Enclosed combustion devices include incinerators, boilers, and process heaters. All three are capable of achieving greater than 95 percent combustion efficiency depending on the operating parameters of the individual combustion device and the vent stream being controlled. Thermal incinerators, catalytic incinerators, and boilers and process heaters are described in detail, including discussion of their efficiencies and applicabilities, in Sections 2.1.2, 2.1.3, and 2.1.4, respectively.

Flares are commonly used in chemical plants but have limited application for controlling the low flow rate and low concentration streams associated with equipment leaks. Flares designed to handle large volumes of vapors associated with overpressure releases may also be used to handle low volumes of equipment leak emissions. However, optimum mixing is not achieved in these cases because the vent stream exit velocity is low and large flares usually cannot properly inject steam into low volume streams. Flares are described in Section 2.1.1, which includes a discussion of flare efficiency and applicability.

2.4.2.2 Vapor Recovery Systems. Vapor recovery systems collect VOC without destroying them. Condensers, adsorbers, and absorbers are all examples of vapor recovery systems and are described in Sections 2.2.2, 2.2.4, and 2.2.5, respectively. These systems all require careful design and operating practices that depend on the VOC species present. A design or operating procedure that works well for one VOC may not work well for other compounds.

2.4.3 Work Practices

Leak detection and repair methods are used to identify equipment components that are emitting significant amounts of VOC and to reduce these emissions. The emission reduction potential for LDAR as a control technique is highly variable and depends on several factors, the most important of which are the frequency of monitoring and the techniques used to

identify leaks. Monthly monitoring is typically more effective in reducing emissions than quarterly monitoring because leaks can be identified and repaired more quickly. Repair of leaking components is required only when the equipment leak emissions reach a set level--the leak definition level. A low leak definition will instigate repair at lower levels, resulting in a lower overall emission rate. The leak occurrence rate, leak recurrence rate, and repair effectiveness of individual sources also affect emission reductions.

Emissions from leaking sources can be reduced by repair, modification, or replacement of the source. The following sections describe various leak detection programs and repair methods to reduce emissions from the leaking equipment.

2.4.3.1 Leak Detection Methods. Leak detection methods include individual component surveys, area (walk-through) surveys, and fixed point monitors. Individual component surveys form a part of the other methods.

2.4.3.1.1 Individual component survey. Each source of equipment leak emissions (pump, valve, compressor, etc.) can be checked for leakage of VOC by visual, audible, olfactory, soap bubble, or instrument techniques. Visual methods are good for locating liquid leaks, especially large pump seal failures. A visible leak does not necessarily indicate VOC emissions, however, because the leaking material may be non-VOC. High-pressure leaks may be detected by the sound of escaping vapors, and leaks of odorous materials may be detected by smell.

Spraying soap on equipment components can be used to survey individual components in certain applications. If the soap solution forms bubbles or blows away, a leak is indicated. If bubbles do not form in the soap solution and the soap solution is not blown away, no leak is detected. Disadvantages of this method are that (1) it does not distinguish leaks of hazardous VOC's from leaks of nonhazardous VOC's; (2) it is only semiquantitative, because it requires the observer to determine subjectively the rate of

leakage based on behavior of the soap bubbles; and (3) it is limited to "cool" sources, because temperatures above 100 °C would cause the water in the soap solution to evaporate. This method is also not suited for moving shafts on pumps or compressors, because the motion of the shaft may interfere with the motion of the bubbles caused by a leak.

Using a portable hydrocarbon detection instrument is the best method for identifying leaks of VOC from equipment components. The instrument is used to sample and analyze the air close to the potential leak surface by traversing the sampling probe tip over the entire area where leaks may occur. This sampling traverse is called "monitoring." The hydrocarbon concentration of the sampled air is displayed on the instrument meter and is an indication of the VOC emission rate from the component. Components that have indicated concentrations higher than a specified "action level" are marked for repair.

2.4.3.1.2 Area survey. An area survey (or walk-through survey) requires the use of a portable hydrocarbon detector and a strip chart recorder. The procedure involves carrying the instrument within 1 m of the upwind and downwind sides of process equipment and associated sources of equipment leak emissions. The instrument is then used for an individual component survey in a suspected leak area. The efficiency of this method for locating leaks is not well established. The time and labor requirements for the walk-through are much lower than for an individual component monitoring survey, but this method will not detect leaks from sources such as overhead valves or relief valves. Leaks from adjacent units and adverse meteorological conditions can affect the results of the walk-through survey. Consequently, the walk-through survey is best for locating only large leaks at small expense.

2.4.3.1.3 Fixed point monitors. This method consists of placing several automatic hydrocarbon sampling and analysis instruments at various locations in the process unit. The instruments may sample the ambient air intermittently or continuously. Elevated hydrocarbon concentrations indicate a

leaking component. As in the walk-through method, an individual component survey with a portable hydrocarbon detector is required to identify the specific leaking component in the area. Leaks from adjacent units and meteorological conditions may affect the results. The efficiency of this method is not well established, but fixed point monitoring of VOC's is not as effective as a complete individual component survey.⁶⁷ This monitoring method is expensive because fixed-point monitors are expensive, multiple units may be required, and the portable instrument is also needed to locate the particular leaking component. Calibration and maintenance costs may be high. Fixed-point monitors are used successfully to detect emissions of hazardous or toxic substances (such as vinyl chloride) as well as potentially explosive conditions. Fixed-point monitors can provide an increased detection efficiency by selecting a particular compound as the sampling criterion.

2.4.3.2 Repair Methods. The following descriptions of repair methods include only those features of each equipment emission source (pumps, valves, etc.) that need to be considered. They are not intended to be complete repair procedures.

Many pumps have in-line or parallel spares that can be used while the leaking pump is being repaired. Leaks from packed seals may be reduced by tightening the packing gland. The packing may deteriorate to the point where further tightening has no effect or even increases emissions from the seal. At this point, the packing can be replaced with the pump out of service. With mechanical seals, the pump must be dismantled to repair or replace the leaking seal. Dismantling pumps can result in spillage of some process fluid. If the seal leak is small, evaporative emissions of VOC from such spillage may be greater than the continued leak from the seal. Precautions must be taken to prevent or reduce these emissions.

Leakage from compressors with packed seals may be reduced by tightening the packing gland, as described for pumps.

Repair of compressors with mechanical seals requires the compressor be removed from service. Because compressors typically do not have spares, immediate repair may not be practical or possible without a process unit shutdown.

Like pumps and compressors, agitators can leak organic chemicals at the point where the shaft penetrates the casing, and seals are required to minimize equipment leak emissions. Leaks from packed seals may be reduced by the repair procedure described for pumps. Repair of other types of seals require the agitator to be out of service, which often requires shutdown of the process or isolation of the particular agitator being repaired. If the leak is small, temporary emissions resulting from a shutdown may be greater than the emissions from the leaking seal; precautions must be taken to prevent or reduce these emissions.

Relief valves that leak usually must be removed for repair. In some cases of improper reseating, manual release of the valve may improve the seat seal. To remove the relief valve without shutting down the process, a block valve may be required upstream of the relief valve. A spare relief valve should be attached while the faulty valve is repaired and tested. The repair and replacement of a relief valve does not guarantee that the next overpressure relief will not result in another leak.

A rupture disk can be installed upstream from a pressure relief valve to eliminate leaks until an overpressure release occurs. Once a release occurs, the rupture disk must be replaced to prevent further leaks. A block valve is required to isolate the rupture disk for replacement.

Most valves have a packing gland that can be tightened while in service. Although this procedure should decrease the emissions from the valve, it can actually increase the emission rate if the packing is old and brittle or has been over-tightened. Plug-type valves can be lubricated with grease to reduce emissions around the plug. Some types of valves have no means of in-service repair and must be isolated from the process and removed for repair or replacement. Other

valves, such as control valves, may be excluded from in-service repair by operating or safety procedures. Most control valves have a manual bypass loop that allows them to be isolated and removed. Most block valves cannot be isolated easily, although temporary changes in process operation may allow isolation in some cases. The emissions resulting from the shutdown of a process unit to isolate a leaking valve may be greater than those occurring from the valve before the next process change (when the valve could be more easily repaired). Depending on site-specific factors, it may be possible to repair process valves by injecting a sealing fluid into the source. This type of repair may affect the operability of the valve and necessitate replacing the valve shortly after its repair. Injection of sealing fluid has been used successfully to repair leaks from valves in petroleum refineries in California.⁶⁸

In some cases, leaks from connectors can be reduced by replacing the connector gaskets, but most connectors cannot be isolated to permit gasket replacement. Tightening of connector bolts also may reduce emissions from connectors. Where connectors are not required for process modification or periodic equipment removal, emissions from connectors can be eliminated by welding them.

2.5 REFERENCES

1. Kalcevic, V. (IT Envirosience). Control Device Evaluation: Flares and the Use of Emissions as Fuels. In: Organic Chemical Manufacturing, Volume 4: Combustion Control Devices. Report 4. Prepared for U.S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-450/3-80-026. December 1980.
2. Klett, M.G. and J.B. Galeski (Lockheed Missiles and Space Co., Inc.). Flare Systems Study. Prepared for U.S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-600/2-76-079. March 1976.
3. U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards. Evaluation of the Efficiency of Industrial Flares: Background--Experimental Design--Facility. EPA-600/2-83-070. Research Triangle Park, NC. August 1983.
4. U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards. Reactor Processes in Synthetic Organic Chemical Manufacturing Industry--Background Information for Proposed Standards. EPA 450/3-90-016a. Research Triangle Park, NC. June 1990.
5. Letter from Matey, J.S., Chemical Manufacturers Association, to Beck, D., EPA/CPB. November 25, 1981. Response to questionnaire on distillation technical issues.
6. Reed, R.J. North American Combustion Handbook. Cleveland, OH. North American Manufacturing Company. 1978. p. 269.
7. U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards. OAQPS Control Cost Manual. Fourth Edition. EPA-450/3-90-006. Research Triangle Park, NC. January 1990. p. 3-43.
8. Telecon. Stone, D.K., Radian Corporation, with Dowd, E., ARI Technology. January 18, 1990. Incinerator sizes and turndown.
9. Memorandum and attachments from Farmer, J.R., EPA/ESD, to Ajax, B. et al. August 22, 1980. Thermal incinerators and flares.

10. Kosusko, M., and G. Ramsey. Destruction of Air Emissions Using Catalytic Oxidation. U.S. Environmental Protection Agency, Air and Energy Engineering Research Laboratory. Research Triangle Park, NC. Publication No. EPA/600-D-88/107. May 1988. p. 3.
11. U.S. Environmental Protection Agency, Office of Air and Waste Management. Control Techniques for Volatile Organic Emissions from Stationary Sources. EPA-450/2-78-002. Research Triangle Park, NC. May 1978. p. 32.
12. Ref. 10, p. 5.
13. Kenson, R.E. A Guide to the Control of Volatile Organic Emissions. MetPro Corp., Systems Division. Harleysville, PA. S104A.5. August 1981.
14. Ref. 10, p. 2.
15. Devitt, T., P. Spaite, and L. Gibbs (PEDCo Environmental, Inc.). Population and Characteristics of Industrial/Commercial Boilers in the U.S. Prepared for U.S. Environmental Protection Agency. Washington, DC. Publication No. EPA 600/7-79-178a. August 1979.
16. U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards. Fossil Fuel Fired Industrial Boilers - Background Information, Volume 1: Chapters 1 - 9. EPA-450/3-82-006a. Research Triangle Park, NC. March 1982. p. 3-27.
17. Hunter, S.C. and S.S. Cherry (KVB). NO_x Emissions from Petroleum Industry Operations. Prepared for the American Petroleum Institute. Washington, DC. API Publication No. 4311. October 1979. p.83.
18. Castaldini, C., H.K. Willard, D. Wolbach, and L. Waterland (Acurex Corporation). A Technical Overview of the Concept of Disposing of Hazardous Wastes in Industrial Boilers. Prepared for the U.S. Environmental Protection Agency. Cincinnati, OH. EPA Contract No. 68-03-2567. January 1981. p. 44.
19. Ref. 18, p. 73.
20. U.S. Environmental Protection Agency. Benzene--Organic Chemical Manufacturing. Emission Test Report, Ethylbenzene/Styrene, Amoco Chemicals Company (Texas City, Texas). EMB Report No. 79-OCM-13. Research Triangle Park, NC. August 1979.

21. U.S. Environmental Protection Agency. Benzene--Organic Chemical Manufacturing, Ethylbenzene Styrene. Emission Test Report, El Paso Products Company (Odessa, Texas). EMB Report No. 79-OCM-15. Research Triangle Park, NC. April 1981.
22. U.S. Environmental Protection Agency. Benzene--Organic Chemical Manufacturing, Ethylbenzene Styrene. Emission Test Report, USS Chemicals (Houston, Texas). EMB Report No. 80-OCM-19. Research Triangle Park, NC. August 1980.
23. U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards. OAQPS Control Cost Manual. Fourth Edition. Supplement 1. Chapter 8. EPA-450/3-90-006a. Research Triangle Park, NC. January 1992.
24. Erikson, D.G. (IT Envirosience). Control Device Evaluation: Condensation. In: Organic Chemical Manufacturing, Volume 5: Adsorption, Condensation, and Absorption Devices. Prepared for U.S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-450/3-80-027. December 1980. Report 2. p. II-2.
25. Green, D.W., ed. Perry's Chemical Engineers' Handbook, Sixth Edition. New York, McGraw Hill Book Company. 1984. pp. 12-29, 12-30.
26. Letter from Plant A to Farmer, J., EPA/ESD. October 1986. Confidential Section 114 response.
27. Letter from Plant B to Farmer, J., EPA/ESD. November 1986. Confidential Section 114 response.
28. Trip Report. Howle, R.H., and M. A. Vancil, Radian Corporation, to file. 7 p. Report of May 12, 1987, visit to Allied Fibers.
29. Memorandum from Zukor, C.J., Radian Corporation to Lassiter, P., EPA/CPB. January 27, 1992. Development of National Emission Impacts from Responses to the March 1990 Section 114 Wastewater Questionnaire: A Summary of the Section 114 Database.
30. Stern, A.C. Air Pollution. Third Edition. Volume IV. New York, Academic Press. 1977. p. 336.
31. Barnett, K.W., P.A. May, and J.A. Elliott (Radian Corporation). Carbon Adsorption for Control of VOC Emissions: Theory and Full Scale System Performance. Prepared for U.S. Environmental Protection Agency. Research Triangle Park, NC. EPA Contract No. 68-02-4378. June 6, 1988. p. 2-2.

32. Basdekis, H.S., and C.S. Parmele (IT Enviroscience). Control Device Evaluation: Carbon Adsorption. In: Organic Chemical Manufacturing, Volume 5: Adsorption, Condensation, and Absorption Devices. Report 1. U.S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-450/3-80-027. December 1980. p. II-1.
33. Ref. 32, p. II-7
34. Ref. 31, p. 3-16.
35. U.S. Environmental Protection Agency, Air and Energy Engineering Research Laboratory. Handbook--Control Technologies for Hazardous Air Pollutants. Publication No. EPA-625/6-86-014. Research Triangle Park, NC. September 1986. p. 26-27.
36. Research and Education Association. Modern Pollution Control Technology, Volume I. New York. 1978. pp. 22-23.
37. Ref. 31, pp. 3-34 and 3-38.
38. Ref. 35, p. 65.
39. Ref. 30, p. 24.
40. Ref. 11, p. 72.
41. Ref. 24, p. II-1.
42. Ref. 25, p. 14-1.
43. Ref. 11, p. 76.
44. Standifer, R.L. (IT Enviroscience). Control Device Evaluation: Gas Absorption. In: Organic Chemical Manufacturing, Volume 5: Adsorption, Condensation, and Absorption Devices. Report 3. U.S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-450/3-80-027. December 1980. p. I-1.
45. Ref. 44, p. I-7.
46. Memorandum from Probert, J.A., Radian Corporation, to HON project file. August 7, 1991. Achievable emission reduction for internal floating roofs.
47. Memorandum from Zukor, C.J., Radian Corporation, to HON project file. July 29, 1991. Emission rates from example internal floating roof tanks with various combinations of rim seals, deck fittings, and deck seams.

48. U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards. VOC Emissions from Volatile Organic Liquid Storage Tanks--Background Information for Proposed Standards. EPA-450/3-81-003a. Research Triangle Park, NC. p. 4-15. July 1984.
49. Ref. 48, Appendix C.
50. Letter and attachments from Blumquist, F.L., PETREX, Inc., to McDonald, R.J., EPA/CPB. April 26, 1991. Submittal of test data on gasketing deck seams.
51. U.S. Environmental Protection Agency. Fugitive Emission Sources of Organic Compounds--Additional Information on Emissions, Emission Reductions, and Costs. Section 5. EPA-450/3-82-010. Research Triangle Park, NC. April 1982.
52. Erikson, D. G. and V. Kalcevic. (IT Envirosience, Inc.) Fugitive Emissions. In: Organic Chemical Manufacturing, Volume 3: Storage, Fugitive, and Secondary Sources. Report 2. U.S. Environmental Protection Agency. Research Triangle Park, NC. EPA-450/3-80-025. December 1980. p. II-2.
53. Ramsden, J. H. How to Choose and Install Mechanical Seals. Chem. Eng. 85(22):97-102. October 9, 1978.
54. Ref. 53, p. 99.
55. Ref. 52, p. IV-I.
56. Nelson, W. E. Compressor Seal Fundamentals. Hydrocarbon Pro. 56(12):91-95. December 1977.
57. Ref. 52, p. II-6.
58. Ramsey, W. D. and G. C. Zoller. How the Design of Shafts, Seals, and Impellers Affects Agitator Performance. Chem. Eng. 83 (18): 101-108. August 30, 1976.
59. Ref. 58, p. 105.
60. American Petroleum Institute. Guide for Pressure Relief and Depressuring Systems. API RP 521. Washington, DC. September 1969.
61. Ref. 52, p. II-13.
62. Lyons, J. L. and C. L. Askland. Lyons' Encyclopedia of Valves. New York, Van Nostrand Reinhold Company. 1975. 290 p.

63. Templeton, H. C. Valve Installation, Operation and Maintenance. Chem. Eng. 78(23):141-149. October 11, 1971.
64. Ref. 63, p. 147-148.
65. Pikulik, A. Manually Operated Valves. Chem. Eng. 85(7):121. April 3, 1978.
66. McFarland, I. Preventing Flange Fires. Chem. Eng Progr. 65(8):59-61. August 1969.
67. Hustvedt, K. C. and R. C. Weber. (U. S. Environmental Protection Agency.) Detection of Volatile Organic Compound Emissions from Equipment Leaks. Presented at the 71st Annual Meeting of the Air Pollution Control Association. Houston, TX. June 25-30, 1978.
68. Teller, J. H. Advantages Found in On-Line Leak Sealing. Oil Gas J. 77(29): 54-59. July 16, 1979.

3.0 COST ANALYSIS

This chapter presents methodologies for costing selected technologies described in Chapter 2 for control of VOC emissions from the five source types in the organic chemical manufacturing industry. The presentation of these methods includes discussion of the design aspects of these technologies that influence costs.

3.1 COST METHODOLOGY FOR COMBUSTION SYSTEMS

This section presents the costing methodologies for two control technologies that reduce VOC emissions by destruction of the organic compounds through combustion. Equipment associated with these systems, design assumptions, and costing equations are provided for flares and incinerators. The incinerator cost includes a packed tower scrubber system that removes acidic vapors from the incinerator flue gas for streams containing halogenated VOC's.

3.1.1 Cost Methodology for Flare Systems

Flare design and costs are based on Chapter 7 of the OAQPS Control Cost Manual (OCCM).¹

3.1.1.1 Flare Design Considerations Affecting Cost. The flare costed in this section is an elevated, steam-assisted, smokeless flare. Elements of the flare system include the knockout drum, liquid seal, stack, gas seal, burner tip, pilot burners, and steam jets. To size the flare system, correlations were developed relating process vent stream flow rate and heat content to the flare height and tip diameter. The general design specifications used in developing these correlations are presented in Table 3-1.

Flare height and tip diameter are the primary design parameters used to determine the capital cost of a flare. The tip diameter selected is a function of the combined flow rates

TABLE 3-1. FLARE GENERAL DESIGN SPECIFICATIONS

Item	Specification										
Emission control efficiency	98 percent destruction										
General flare design	Elevated, steam-assisted, smokeless flare										
- minimum net heating value	300 Btu/scf of gas being combusted										
- minimum flare tip diameter	2.5 cm (1.0 in.)										
- minimum flare height	9.1 m (30 ft)										
- maximum ground-level heat intensity ^a	1,576 W/m ² (500 Btu/hr/ft ²)										
- flare tip velocities ^b	$HV \leq 11.2$ (300): $V_{max} = 18.3$ m/sec (60 ft/sec) + natural gas to 2 MJ/Nm^3 (300 Btu/scf) 11.2 (300) < $HV < 37.3$ (1,000): $\log(V_{max}) = (HV + 1,214)/852$ $HV > 37.3$ (1,000): $V_{max} = 122$ m/sec (400 ft/sec)										
- fraction of heat radiated	0.2										
- fraction of heat transmitted	1.0										
- number of pilots	<table> <tr> <th>Number of Pilots</th><th>Tip Diameter^c</th></tr> <tr> <td>1</td><td>$D \leq 25$ (0 ≤ 10)</td></tr> <tr> <td>2</td><td>$30 < D \leq 61$ (12 < D ≤ 24)</td></tr> <tr> <td>3</td><td>$75 < D \leq 152$ (30 < D ≤ 60)</td></tr> <tr> <td>4</td><td>$152 < D$ (60 < D)</td></tr> </table>	Number of Pilots	Tip Diameter ^c	1	$D \leq 25$ (0 ≤ 10)	2	$30 < D \leq 61$ (12 < D ≤ 24)	3	$75 < D \leq 152$ (30 < D ≤ 60)	4	$152 < D$ (60 < D)
Number of Pilots	Tip Diameter ^c										
1	$D \leq 25$ (0 ≤ 10)										
2	$30 < D \leq 61$ (12 < D ≤ 24)										
3	$75 < D \leq 152$ (30 < D ≤ 60)										
4	$152 < D$ (60 < D)										
- pilot gas requirement	2.0 m ³ /hr (70 scf/hr) of natural gas per pilot										
- steam requirement	0.4 kg steam/kg vent gas										
- purge gas requirement	Purge gas added to maintain a minimum flare tip velocity of 0.01 m/sec (0.04 ft/sec)										
Supplemental fuel requirement	Natural gas required to maintain vent stream HV of 11.2 MJ/m ³ (300 Btu/scf)										

^aExcluding solar radiation of 300 Btu/hr ft².

^bHV = Heat content value of process vent stream, MJ/m³ (Btu/scf). V_{max} = maximum vent stream flow rate. A flare tip velocity equal to 80 percent of the maximum smokeless velocity [18.3 m/sec (60 ft/sec)] is used in the costing equations.

^cD = Tip diameter, in cm (in.).

of the vent stream and supplemental fuel, and the assumed tip velocity. Supplemental fuel requirements and tip velocity values are listed in Table 3-1. The maximum allowed velocity ranges from 18.3 to 121.9 m/sec (60 to 400 ft/sec), as required by rules defined in the Federal Register.² The flare tip diameter is sized to allow a flare tip velocity of 80 percent of the maximum design velocity. Determination of flare height is based on worker safety requirements and is selected so the maximum heat intensity at ground level including solar radiation is 1,576 W/m² (500 Btu/hr ft²). Vendor contacts indicate the smallest elevated flare commercially available is 2.54 cm (1 in.) in diameter and 9.14 m (30 ft) high.

After the flare tip diameter, D, and flare height, H, are determined, the natural gas required for pilots and purge and the mass flow rate of steam required are calculated. Pilot gas consumption is a function of the number of pilots, which is selected based on the tip diameter (Table 3-1). Pilot gas consumption is calculated based on an energy-efficient model of 2 scm/hr (70 scf/hr) per pilot burner. The purge gas requirement is also a function of the tip diameter and the minimum design purge gas velocity of 0.012 m/sec (0.04 ft/sec) at the tip (Table 3-1). Steam is required to maintain a steam-to-flare-gas ratio of 0.4 kg steam/kg flare gas.

3.1.1.2 Development of Flare Capital Costs. The capital cost of a flare is based on vendor-supplied information as correlated in the OCCM cost equations, regressed from the combined data set over a range of tip diameters and flare heights. Flare equipment costs, C_F, are calculated based on stack height, H (ft), and tip diameter, D (in.), according to type of support for the flare:

- Self supported:

$$C_F \text{ (July 1989 dollars)} = [78.0 + 9.14(D) + 0.749(H)]^2$$

- Guy supported:

$$C_F \text{ (July 1989 dollars)} = [103 + 8.68(D) + 0.470(H)]^2$$

- Derrick supported:

$$C_F \text{ (July 1989 dollars)} = [76.4 + 2.72(D) + 1.64(H)]^2$$

The total cost for flare equipment includes the costs of the flare tower (stack) and support, burner tip, pilots, piping for utilities, piping from base, meters and controls for utilities, liquid seal, gas seal, and galvanized caged ladders and platforms as required. The material of construction is carbon steel, except for the upper 4 ft and burner tip, which is 310 stainless steel.

Vent stream piping costs are a function of flare tip diameter and length of piping. The pipe diameter is assumed to be equal to the flare tip diameter. Pipe costs are calculated using the following equation:

$$C_p \text{ (July 1989 dollars)} = 1.27 * D^{1.21} * 400 \text{ for } D < 24$$

$$C_p \text{ (July 1989 dollars)} = 1.39 * D^{1.07} * 400 \text{ for } D \geq 24$$

where:

C_p = vent stream piping costs; and

D = diameter (in.)

These costs include 400 ft of Schedule 40, carbon steel pipe.

Knockout drum costs are a function of drum diameter (in.) and drum wall thickness (in.):

$$C_K = 14.2 \{ (d) * (t) * [h + 0.812 * (d)] \}^{0.737}$$

where:

C_K = knockout drum costs;

d = drum diameter; and

t = drum wall thickness

The cost of collection fan, C_{Fan} , is a function of total stream flow rate, Q_{tot} (scfm):

$$C_{fan} = 96.96 * Q_{tot}^{0.5472}$$

Total flare system equipment cost is the sum of flare, piping, knockout drum, and fan costs:

$$EC = C_F + C_K + C_p + C_{Fan}$$

where:

EC = equipment costs;

C_F = flare cost;

C_K = knockout drum cost; and

C_p = piping costs.

Purchased equipment cost equals equipment cost plus factors for auxiliary equipment (i.e., instrumentation at 0.10, sales tax at 0.03, and freight at 0.05). Installation cost is estimated as a percentage of purchased equipment cost. The total capital investment is obtained by multiplying the purchased equipment cost by an installation factor of 1.92. Further details about flare costs are presented in Chapter 7 of the OCCM.¹

3.1.1.3 Development of Flare Total Annual Cost. The annual costs for a flare system include direct operating and maintenance costs, and annualized capital charges. The assumptions used to determine annual costs are presented in Table 3-2, and are given in July 1989 dollars. Direct operating and maintenance costs include operating and maintenance labor, maintenance materials, and utilities.

3.1.1.3.1 Labor costs. The operating labor requirements are 630 hr/yr for typical flare systems. Maintenance labor requirements are .5 hr per 8-hr shift. Supervisory labor is estimated to be 15 percent of the operating labor cost. The hourly rate for maintenance labor is assumed to be 10 percent higher than that for operating labor.

3.1.1.3.2 Capital charges. Capital recovery costs are estimated using the following equation:

$$CRC = CRF * TCI$$

where:

$$\begin{aligned} CRC &= \text{capital recovery charges;} \\ CRF &= \text{capital recovery factor; and} \\ TCI &= \text{total capital investment.} \end{aligned}$$

The capital recovery factor (0.1314) is based on a 10 percent interest rate and a 15-year life for the equipment. Taxes, insurance, and administrative costs are assumed to be 4 percent of the total capital investment. Also, overhead costs are assumed to be 60 percent of all labor and maintenance costs.

3.1.1.3.3 Utility costs. The utilities considered in the annual cost estimates include steam, natural gas, and electricity. Cost of steam to eliminate smoking is equal to

TABLE 3-2. BASES AND FACTORS FOR ANNUAL COSTS
FOR FLARES

<u>Basis for Direct Annual Costs</u>	<u>Factor</u>
Operating labor (man-hours/8-hr shift)	0.575
Maintenance labor (man-hours/8-hr shift)	0.5
Labor rates (July 1989 \$/hr) ^a	
Operating labor	13.20
Maintenance labor	14.50
Supervisory labor cost (percent of operating labor cost)	15
Maintenance materials cost (percent of maintenance labor cost)	100
Utilities (July 1989 \$) ^a	
Electricity (\$/1,000 KW-hr)	50.9
Natural Gas (\$/1,000 ft ³)	3.03
Steam (\$/Mg)	7.68
<u>Basis for Indirect Annual Cost</u>	
Equipment life (years)	15
Interest rate (percent)	10
Capital recovery factor	0.1314
Taxes, insurance, administration (percent of total capital investment)	4
Overhead (percent of total labor and maintenance costs)	60

^aBased on Reference 3. ³

the annual steam consumption multiplied by the unit cost. Electrical costs for operating a fan are proportional to total stream flow rate. Flares use natural gas in three ways: for pilot burners, as an auxiliary fuel for low Btu vent streams, and as a purge gas. The procedures for estimating electricity, steam, and supplemental fuel requirements are described in Section 3.1.1.1 and in Chapter 7 of the OCCM.¹

3.1.1.3.4 Maintenance costs. Maintenance material costs are assumed to equal maintenance labor costs, which are discussed in Section 3.1.1.3.1.

3.1.2 Cost Methodology for Incinerator Systems

Incinerator costs were developed using Chapter 3 of the OCCM.¹ Scrubber costs were developed using the procedure outlined in EPA's Handbook on Control Technologies for Hazardous Air Pollutants (HAP Manual),⁴ with equipment costs updated from recent technical journal information.^{5,6}

3.1.2.1 Thermal Incinerator Design Considerations Affecting Cost. The thermal incinerator system consists of the following equipment: combustion chamber, instrumentation, recuperative heat exchanger, blower, collection fan and ductwork, quench/scrubber system (if applicable), and stack. The OCCM contains further discussion of incinerator control system design. Control system elements and design assumptions specific to SOCM vent streams are discussed below. General incinerator design specifications are presented in Table 3-3.

3.1.2.1.1 Combustion air requirements. The amount of oxygen in the vent stream or bound in the VOC establishes the supplemental combustion air requirement, which has an impact on both the capital and operating costs of the thermal incinerator. Supplemental combustion air may be required to support combustion. This cost analysis assumes that the vent stream is a mixture of VOC, oxygen, and an inert gas such as nitrogen. After combustion, the excess oxygen content in the incinerator flue gas is designed to be at least 3 mole percent, which is based on commonly accepted operating practice.

TABLE 3-3. INCINERATOR GENERAL DESIGN SPECIFICATIONS

Item	Specification
Emission control efficiency	98 percent destruction
Minimum incinerator capacity ^a	500 scfm
Maximum incinerator capacity	50,000 scfm
Incinerator temperature	
- Nonhalogenated vent streams	870 °C (1,600 °F)
- Halogenated vent streams ^b	1,100 °C (2,000 °F)
Chamber residence times	
- Nonhalogenated vent streams	0.75 sec
- Halogenated vent streams ^b	1.00 sec
Supplemental fuel requirement	Natural gas required to maintain incinerator temperature with 3 mole percent excess oxygen in flue gas
Scrubber system	Used when halogenated VOC is present to remove corrosive combustion by-products
- Type	Packed tower
- Packing type	2-in. rings, ceramic
- Scrubbing liquid	Water
- Scrubber gas temperature	25 °C (77 °F)

^aFive hundred scfm is the minimum incinerator size used to determine capital cost. A 10:1 turndown ratio is assumed and a minimum flow rate of 50 scfm is used to determine operating costs.

^bUsed when halogenated VOC are present due to the difficulty of achieving complete combustion of halogenated VOC at lower temperatures.

To calculate the amount of combustion air required to ensure a flue gas O₂ concentration of 3 mole percent, a complete stoichiometric equation must be balanced for each compound present in the vent stream. In many cases, the complete chemical composition of the vent stream is not known. Thus, to cost incinerator systems for typical vent streams encountered, a design molecule approach was used. The design molecule was based on a survey of typical values for carbon, hydrogen, oxygen, sulfur, and chloride ratios for a group of 219 organic compounds.^{8,9} For the analysis, the average VOC molecular composition of 68.3 percent carbon, 11.4 percent hydrogen, and 20.3 percent oxygen was used to calculate combustion air requirements. These weight ratios correspond to a molecular formula of C_{2.85}H_{5.70}O_{0.63}. If zero percent O₂ in the waste stream is assumed, a dilution ratio (mole of air per mole of VOC) of approximately 18:1 is required to achieve 3 percent O₂ in the incinerator flue gas. If sufficient oxygen is present to ensure 3 percent O₂ in the flue gas, no combustion air is added.

3.1.2.1.2 Dilution air requirements. After the required combustion air is calculated and added to the total vent stream flow, the overall heat content (Btu/scf) of the stream is recalculated. Adding combustion air will effectively dilute the stream and lower the heat content of the combined stream fed to the incinerator. However, if the heat content of the vent stream is still greater than 98 Btu/scf for nonhalogenated streams or greater than 95 Btu/scf for halogenated streams, additional dilution air must be added to reduce the heat content to these acceptable levels. The imposition of a maximum heat content level prevents the temperature in the incinerator from exceeding the design specifications.

The minimum flow rate to the incinerator is 50 scfm. It is assumed that vent streams smaller than 50 scfm will be mixed with air to achieve this minimum flow rate. The maximum incinerator flow rate is 50,000 scfm. Flow rates greater than

50,000 scfm are assumed to be handled by multiple incinerators.

3.1.2.1.3 Recuperative heat recovery. Halogenated vent streams are not considered candidates for heat recovery systems and are costed assuming zero percent heat recovery. This conservative design assumption is imposed because of the potential for corrosion in the heat exchanger and incinerator. If the temperature of the flue gas leaving the heat exchanger, T_{fo} , were to drop below the acid dew point temperature, acid gases would condense. Significant corrosion can lead to shortened equipment life, higher maintenance costs, and potentially unsafe working conditions.

Nonhalogenated vent streams are considered candidates for recuperative heat recovery. The extent of heat recovery depends on the heat content of the vent stream after dilution. Four different heat recovery scenarios are evaluated for nonhalogenated streams. The cost algorithm includes systems with 0, 35, 50, and 70 percent heat recovery. The extent of heat exchange to be used is decided by an economic optimization procedure with the following restrictions. No heat recovery is allowed for vent streams with a heat value greater than 25 percent of the LEL because of the possibility of explosion or damaging temperature excursions within the heat exchanger. This limit typically corresponds to a heat content of 13 Btu/scf. Therefore, if the heat content of the total vent stream is still greater than 13 Btu/scf, even after addition of required combustion and dilution air, then the vent stream cannot be preheated. For streams with a heat content less than 13 Btu/scf, the entire stream is preheated in the recuperative heat exchanger, thus allowing for maximum energy recovery. However, for streams with a heat content greater than 13 Btu/scf, only the combustion/dilution air stream can be preheated. In this case, the cost optimization procedure evaluates the option of preheating only the air stream and combines the VOC stream with the preheated air stream in the incinerator.

All allowable heat recovery percentages are evaluated, and the calculated total capital and annual costs are based on the most cost effective configuration. The trade-off between the capital cost of the equipment and the operating cost (fuel) of the system determines the optimum level of energy recovery.

3.1.2.1.4 Incinerator design temperature. The destruction of VOC is a function of incinerator temperature and residence time in the combustion chamber. The design VOC destruction efficiency is 98 percent, which can be met by well-designed and well-operated thermal incinerator systems. Previous EPA studies show that 98 percent 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 sec.¹⁰ Thermal oxidation of halogenated VOC requires higher temperature oxidation to convert the combustion product to a form that can be more readily removed by flue gas scrubbing. For instance, chloride-containing vent streams are burned at high temperature to convert the chlorine to HCl instead of to Cl₂, because HCl is more easily scrubbed. Available data indicate that a temperature of 2,000 °F and residence time of 1 sec are necessary to achieve 98 percent VOC destruction efficiency for halogen-containing vent streams.¹⁰

3.1.2.1.5 Acid gas scrubber design consideration. Acid gas scrubbers are packed towers designed to promote absorption of acidic combustion by-products. Scrubbers consist of the following major equipment: quench chamber, absorption column and platform, packing, piping, ductwork and blowers, and the release stack. The HAP Manual contains a full discussion of scrubber design procedures. System elements and design assumptions specific to this analysis are discussed below. General scrubber design specifications are presented in Table 3-4. Column diameter and height are the primary design parameters in determining the capital cost of the scrubber. These design parameters establish the column shell geometry and amount of packing required.

TABLE 3-4. SCRUBBER GENERAL DESIGN SPECIFICATIONS

Item	Specification
Emission control efficiency ^a	98 percent recovery
Column type ^a	Packed tower
Quench chamber temperature ^a	25 °C
Scrubbing liquid ^a	Water
Packing type ^a	2-in. ceramic rings
Packing constant ^{b,c} :	
a	28
b	3.82
c	0.41
d	0.45
e	0.74
s	0.22
Y	0.0125
g	11.13
r	0.00295
Absorption factor ^a	19
Schmidt number for acid streams (gas) ^d :	
Hydrogen chloride	0.809
Hydrogen bromide	1.175
Hydrogen fluoride	0.78
Schmidt number for acid streams liquid):	
Hydrogen chloride ^b	381
Hydrogen bromide ^d	517.1
Hydrogen fluoride ^d	300.6

^aFrom Reference 4.^bFrom Reference 11. ¹¹^cFrom Reference 12. ¹²^dFrom Reference 13. ¹³

The design procedure assumes no heat effects are associated with the absorption process and that both the gas and liquid streams are dilute solutions. The system flow rates (gas and liquid) are assumed to be constant across the scrubber. Vendor data for suggested liquid-to-vapor flow rate ratios throughout scrubber columns were used to develop an average of 17 gpm of liquid per 1000 scfm of vapor.¹⁴ The equilibrium curve can be approximated as a straight line with a worst case slope of 0.1.¹⁵ This value reflects the highly soluble nature of hydrogen chloride, hydrogen fluoride, and hydrogen bromide, which results in a large driving force for absorption.

Column diameter determination. The column diameter is estimated based on the characteristics of the vent stream, the absorption liquid, the packing material, and an assumed column flooding condition. The column operating range is assumed to be 60 percent of the flooding rate. Typically, design manuals obtain the diameter of the column from graphical information presented as a correlation curve for flooding rate in randomly packed absorption towers. The abscissa for the column diameter curves is defined by the following expression:

$$\text{abscissa} = (L/G)(\rho_G/\rho_L)^{0.5}$$

where:

$$\begin{aligned} L &= \text{liquid mass flow rate (lb/hr);} \\ G &= \text{gas stream mass flow rate (lb/hr);} \\ \rho_G &= \text{gas stream density (lb/ft}^3\text{); and} \\ \rho_L &= \text{liquid stream density (lb/ft}^3\text{).} \end{aligned}$$

After substituting the appropriate parameters, the abscissa value can be determined. Graphically, estimating the column diameter would require locating the abscissa point and proceeding up to the flooding line and selecting the corresponding ordinate. The ordinate is defined as:

$$\text{ordinate} = (G_{\text{area}})^2 (a/\epsilon^3) (\mu_L^{0.2}) / (g_c \rho_L \rho_G)$$

where:

$$\begin{aligned} G_{\text{area}} &= \text{gas stream flux rate at flooding conditions (lb/ft}^2 \text{ sec);} \\ a, \epsilon &= \text{packing constants;} \end{aligned}$$

g_c = gravitational constant (ft/sec²); and
 μ_L = viscosity of solvent (centipoise).

The flux rate is estimated by simple substitution of the system constants, and the column diameter is calculated based on the estimated flux rate.

To solve for G_{area} algebraically, the flooding line is approximated mathematically. The abscissa is calculated as described above. The ordinate value can be approximated by the following equation:

$$\text{ordinate} = -0.9809 * (\text{ABS})^{[-0.0065 * \ln(\text{ABS})]} + \text{ABS} - 0.0219$$

where:

ABS = abscissa.

The flux rate is then estimated by the expression:

$$G_{area} = 0.6 * [\text{ordinate} * \rho_G * \rho_L * g_c / (a/\epsilon^3) * \mu_L^{0.2}]^{0.5}$$

This approximation method provides a reasonable estimate of the column flux rate without performing graphical interpretations.

Once the flux rate has been estimated, the column cross-sectional area and corresponding diameter are determined. The column area and diameter are calculated by:

$$\begin{aligned} A_{\text{column}} &= MW * g / (3600 * G_{area}) \\ D_{\text{column}} &= 1.13 (A_{\text{column}})^{0.5} \end{aligned}$$

where:

A_{column} = column area (ft²);
 g = gas molar flow (lb-moles/hr);
 MW = Molecular weight of entire gas stream; and
 D_{column} = column diameter (ft).

Column height determination. The column must be tall enough to ensure the required removal efficiency. The height of a packed column is calculated by determining the required number of theoretical transfer units and multiplying by the height of a transfer unit. A transfer unit is a measure of the difficulty of the mass transfer operation and is a function of the solubility and concentration of the solute in

the gas and liquid streams. The number of transfer units are expressed as NOG (number of gas transfer units) or NOL (number of liquid transfer units), depending on whether the gas film or liquid film resistance controls the absorption rate.

In emission control applications, gas film resistance typically is controlling; therefore, NOG will be used in the following calculations. The expression for packing height is:

$$H_{t\text{packing}} = \text{NOG} * \text{HOG}$$

where:

$$\begin{aligned} H_{t\text{packing}} &= \text{packing height (ft);} \\ \text{NOG} &= \text{number of gas transfer units; and} \\ \text{HOG} &= \text{height of an overall gas transfer unit (ft).} \end{aligned}$$

Although determining the number of gas transfer units is not usually complicated when dilute solutions are involved, NOG can be calculated using the following equation:

$$\text{NOG} = \frac{\ln\{(\text{VOCe}/\text{VOC}_0) \times [1 - (1/\text{AF})] + (1/\text{AF})\}}{[1 - (1/\text{AF})]}$$

where:

$$\begin{aligned} \text{VOCe} &= \text{VOC inlet concentration (ppmv);} \\ \text{VOC}_0 &= \text{VOC exit concentration from column (ppmv);} \\ &\text{and} \\ \text{AF} &= \text{VOC absorption factor.} \end{aligned}$$

To calculate VOC_0 , the inlet VOC concentration must be calculated and then reduced by the scrubber control efficiency. The absorption factor used in this analysis¹⁵ is 19.16, based upon typical absorption conditions for strong acids.

The calculation of the height of an overall transfer unit is performed based upon the following equation:

$$\text{HOG} = H_G + (1/\text{AF}) * H_L$$

where:

$$\begin{aligned} \text{HOG} &= \text{height of an overall transfer unit (ft);} \\ H_G &= \text{height of a gas transfer unit (ft); and} \\ H_L &= \text{height of a liquid transfer unit (ft).} \end{aligned}$$

Generalized correlations are available to calculate H_G and H_L ;

these are based on the type of packing and the gas and solvent flow rates. The correlations for H_G and H_L are as follows:

$$H_G = [b \cdot (3,600 G_{\text{area}})^c / (L'')^d] \cdot (Sc_G)^{0.5}$$

$$H_L = Y \cdot (L'' / \mu_L)^s \cdot (Sc_L)^{0.5}$$

where:

$$b, c, d, Y, \text{ and } s = \text{packing constants};$$

$$L'' = \text{liquid flow rate (lb/hr-ft}^2\text{)};$$

$$\mu_L = \text{liquid viscosity (lb/hr-ft)};$$

$$Sc_G = \text{gas stream Schmidt number; and}$$

$$Sc_L = \text{liquid stream Schmidt number.}$$

In this calculation, the effect of temperature on the Schmidt numbers is assumed to be negligible. The value for the variable L'' in this equation is defined as follows:

$$L'' = L / A_{\text{column}}$$

Once the height of the column is determined, the total system height is calculated based upon the following expression:

$$H_{\text{total}} = H_{\text{packing}} + 2 + 0.25 \cdot D_{\text{column}}$$

Estimation of the total system height provides the remaining design factor required for costing purposes.

3.1.2.2 Development of Thermal Incinerator and Scrubber Capital Costs. The costing analysis follows the methodology outlined in the OCCM.¹ Equipment cost correlations are based on data provided by various vendors; each correlation is valid for incinerators in the 500 to 50,000 scfm range.¹ Thus, the smallest incinerator size used for determining equipment costs is 500 scfm; for flow rates above 50,000 scfm, additional incinerators are costed.

Equipment costs for thermal incinerators are given as a function of total volumetric throughput. Four equations were used in the costing analysis, each pertaining to a different level of heat recovery. After converting to July 1989 dollars, the equations are:

$EC = 10772 Q_{\text{tot}}^{0.2355}$	HR = 0 percent
$EC = 13760 Q_{\text{tot}}^{0.2609}$	HR = 35 percent
$EC = 17848 Q_{\text{tot}}^{0.2502}$	HR = 50 percent
$EC = 22333 Q_{\text{tot}}^{0.2500}$	HR = 70 percent

where:

EC = equipment costs (July 1989 dollars)
Q_{tot} = total volumetric throughput (scfm); and
HR = heat recovery.

The cost of ductwork (not included in equipment costs) is based on a 24-in. diameter duct made of 1/8-in. thick carbon steel with two elbows per 100 ft. The following equation is that used by Vataavuk:⁶

$$\text{Pipe Cost} = [(210 * 24^{0.839}) + (2 * 4.52 * 24^{1.43})] * 3$$

The length of duct is assumed to be 300 ft. Collection fan costs are calculated using methods developed by Richardson Engineering Services and are discussed in Section 3.2.12. The duct and fan costs are added to the equipment costs.

Purchased equipment cost is calculated by accounting for instrumentation, taxes, and freight, which are assumed to be 18 percent of total equipment cost (including auxiliary equipment).

Installation costs are estimated as a percentage of purchased equipment costs. Table 3-5 lists the values of direct and indirect installation factors for thermal incinerators. Installation costs can be as high as 61 percent of the purchased equipment cost. However, the OCCM suggests that installation costs would be approximately 20 to 25 percent of the purchased equipment cost for incinerator units handling flow rates less than 20,000 scfm.¹

The costing methodology for the scrubber follows the procedure outlined in the HAP manual.⁴ Equipment costs for scrubbers are given as a function of column weight, WT_{col}, in pounds. The following were used in the costing analysis:

$$EC = \left[1.900604 * \left(\frac{wt_{col}}{1000} \right)^{0.9389} \right] * 1000 * \frac{CE \text{ index}}{298.2}$$

$$wt_{col} = (48 * D_{col} * HT_{col}) + 39 * (D_{col})^2$$

where:

$$wt_{col} = \text{weight of the column (lb);}$$

TABLE 3-5. CAPITAL COST FACTORS FOR THERMAL INCINERATORS^a

Cost Item	Factor
<u>Direct Costs</u>	
Purchased equipment costs	
Incinerator + auxiliary equipment ^b	As estimated, A ^c
Instrumentation ^d	0.10 A
Sales taxes	0.03 A
Freight	0.05 A
Purchased equipment cost	<u>B = 1.18 A</u>
Direct installation costs	
Foundations and supports	0.08 B
Handling and erection	0.14 B
Electrical	0.04 B
Piping	0.02 B
Insulation for ductwork	0.01 B
Painting	0.01 B
Direct installation cost	<u>0.30 B</u>
Site preparation	As required, SP
Buildings	<u>As required, Bldg.</u>
Total direct costs (DC)	1.30 B + SP + Bldg.
<u>Indirect Costs (Installation)</u>	
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 cost (IC)	<u>0.31 B</u>
Total capital investment = DC + IC	<u>1.61 B + SP + Bldg.</u>

^aFrom Reference 1.^bDuctwork and any other equipment normally not included with unit furnished by incinerator vendor.^cThe factor "A" is the total equipment cost for the incinerator and auxiliary equipment.^dInstrumentation controls often furnished with the incinerator, and thus often included in the equipment costs.

D_{col} = diameter of the column (ft); and
 HT_{col} = height of the column (ft).

3.1.2.3 Development of Thermal Incinerator Total Annual Cost. Annual costs for the thermal incinerator system include direct operating and maintenance costs, as well as annualized capital charges. An assumed incinerator turn-down ratio of 10:1 was used in this cost analysis.⁷ Consequently, the minimum flow rate for determining operating costs is assumed to be 50 scfm. Additional dilution air is added where necessary to raise the total flow rate of the fuel, vent stream, and air to 50 scfm. The bases for determining thermal incinerator annual costs are presented in Table 3-6. Each cost parameter is reviewed below.

3.1.2.3.1 Labor costs. The operating labor requirements vary depending on the components of the overall system. Continuously operating incinerator systems without a scrubber require the least amount of operating labor (548 hr/yr or 0.5 hr per 8-hr shift). Systems employing a scrubber require an additional 548 hr/yr operating labor. Maintenance labor requirements are assumed to be identical to operating labor requirements, that is, 548 hr/yr for the incinerator and 548 hr/yr for the scrubber. Supervisory cost is estimated to be 15 percent of the operating labor cost. The maintenance labor hourly rate is assumed to be 10 percent higher than the operating labor hourly rate.

3.1.2.3.2 Capital recovery charges. The capital recovery factor (0.163) is based on a 10 percent interest rate and a 10-year life for the equipment. Taxes, insurance, and administrative costs are assumed to be 4 percent of the total capital investment. Overhead is estimated to be 60 percent of the total labor and maintenance costs.

3.1.2.3.3 Utility costs. The utilities considered in the annual cost estimates include natural gas and electricity. The procedures for estimating electricity and supplemental fuel requirements are described in Chapter 3 of the OCCM.¹

TABLE 3-6. BASES AND FACTORS FOR ANNUAL COSTS
FOR THERMAL INCINERATORS

<u>Basis for Direct Annual Costs</u>	<u>Factor</u>
Operating labor (man-hours/8-hr shift)	0.5
Maintenance labor (man-hours/8-hr shift)	0.5
Labor rates (July 1989 \$/hr) ^a	
Operating labor	13.20
Maintenance labor	14.50
Supervisory labor cost (percent of operating labor cost)	15
Maintenance materials cost (percent of maintenance labor cost)	100
Utilities (July 1989 \$) ^a	
Electricity (\$/1,000 KW-hr)	50.9
Natural Gas (\$/1,000 ft ³)	3.03
<u>Basis for Indirect Annual Cost</u>	
Equipment life (years)	10
Interest rate (percent)	10
Capital recovery factor	0.163
Taxes, insurance, administration (percent of total capital investment)	4
Overhead (percent of total labor and maintenance costs)	60

^aBased on Reference 3.

3.1.2.3.4 Maintenance costs. Maintenance material costs are assumed to be equal to maintenance labor costs, which are discussed in Section 3.1.2.3.1.

3.1.2.4 Scrubber Annual Cost. Annual costs for the acid gas scrubber include such direct operating costs as scrubber utility costs, operating labor, maintenance labor and materials, supervisory labor, and capital recovery charges. All labor costs are calculated using the same methodology as that used for thermal incinerators (Table 3-6). Utility costs include electricity and water costs.

Capital recovery charges are estimated based on a 10-year equipment life and an interest rate for depreciation of 10 percent. Taxes, insurance, and administrative costs are assumed to be 4 percent of the total capital investment. Overhead is estimated as 60 percent of the total labor and maintenance costs.

3.2 COST METHODOLOGY FOR COLLECTION SYSTEMS AND RECOVERY DEVICES

3.2.1 Cost Methodology for Vapor Collection Systems for Loading Racks

This section discusses the capital and annual costs for retrofitting a transfer loading rack with a vapor collection system, retrofitting tank trucks and tank cars to be vapor tight and compatible with loading rack vapor collection systems, and incorporating a nitrogen blanketing system for the rack and vehicles. The costs were obtained from technical work performed in support of the New Source Performance Standards (NSPS) for bulk gasoline terminals and the NESHAP for benzene emissions from benzene transfer operations.

3.2.1.1 Design Considerations Affecting Costs. Several factors affect the design and the costs of retrofitting the loading racks and vehicles. These parameters include the following:

- Transfer loading throughput;
- Chemical vapor pressure;
- Transport vehicle type;
- Transport vehicle fleet size;

- Transport vehicle vapor tightness; and
- Nitrogen blanketing.

The loading throughput affects the size of the loading rack and the costs to retrofit the rack. The loading throughput also affects the number of transport vehicles in a fleet, which, in turn, affects the costs to make the vehicles vapor tight and the costs to retrofit the fleet with the proper vapor collection equipment.

The vapor pressure of a compound affects the level of control necessary to reduce product loss. Transferring a compound with a low vapor pressure might not require a vapor collection system because only a small quantity of the material would actually be vaporized and emitted to the atmosphere.

A nitrogen blanketing system may be used to reduce the potential for VOC vapors to accumulate in concentrations approaching the explosive range. Nitrogen blanketing may also be used during transfer of some organic compounds to eliminate the moisture in the ambient air. This is usually done to ensure that extremely hydrophilic compounds remain moisture-free during transfer.

3.2.1.2 Development of Capital Costs. The total capital investment includes the costs for vapor collection equipment, retrofitting the loading rack, retrofitting the vehicle fleet for vapor collection and vapor tightness, nitrogen blanketing, and installation.

Vapor collection equipment includes piping, fans, and motors. The amount of piping needed is a function of the distance between the transfer rack and the control device. It is assumed that the cost of this piping will be included in the cost of the control device.

A fan is used to transport the VOC vapors from the loading rack to the control device. The capital cost of the fan and motor may be calculated from the following equation:¹⁶

$$\begin{aligned} &\text{Fan Costs} \\ &(\text{Including Motor}) = 100.79 * (\text{Flow Rate, cfm})^{0.5472} \\ &(\text{July 1989 \$}) \end{aligned}$$

Based on information provided in the BID for the bulk gasoline terminals NSPS, the installed cost in July 1989 dollars for converting transport vehicles from top splash loading to bottom loading is \$8,080 per tank truck and \$15,700 per tank car. The installed cost for converting a loading rack from top loading to bottom loading is \$252,400 per rack. These values include the cost of equipment for connecting the vehicles and racks to a vapor recovery system.^{17,18} Approximately 97 percent of the SOCFI has already converted its vehicles and, where necessary, loading racks for submerged fill or bottom loading.¹⁹ Such facilities will not incur the retrofit costs described above.

3.2.1.3 Development of Annual Costs. Direct annual cost includes the electricity to run the vapor collection system fan and the toll for the N₂ blanketing system. The electricity cost for the motor is a function of the vent stream flow rate, the hours of operation, and the pressure drop through the piping from the loading rack to the control device. Because transfer operations are discrete events, not continuous, the fan will not operate continuously. However, it will operate during loading and for a period after each loading to clear the lines of VOC vapors from residual liquid in the lines. As a conservative estimate, it is assumed that the fan will operate for a period twice that of loading. Pressure drop will be a function of flow rate and pipe length. The following equation can be used to estimate the annual electricity costs for the fan:²⁰

Annual Fan
Electricity

$$\text{Costs (July 1989 \$)} = 9.926 * 10^{-6} * (\text{Flow Rate}) (\Delta P) (\text{Loading Hours} * 2)$$

where:

Flow rate = vent stream flow rate in cfm; and

ΔP = pressure drop in inches of water.

For the large volume of nitrogen needed to blanket the vehicles during loading, a facility is assumed to use a nearby air separation plant. The cost is based on the volume of

nitrogen needed, which would equal the throughput transferred at the facility. The following equation can be used to estimate this cost:²¹

$$\begin{array}{l} \text{Annual Cost for} \\ \text{N}_2 \text{ Blanketing} \\ \text{(July 1989\$)} \end{array} = 274 * (\text{Annual Throughput, MMgal/yr})$$

Indirect annual cost includes capital recovery, vehicle vapor-tightness testing, and maintenance for the vapor collection system. These costs are summarized in Table 3-7. Capital recovery is calculated as:

$$\text{Capital recovery} = 0.163 * \text{TCI}$$

where the capital recovery factor (0.163) is based on an equipment life of 10 years and an interest rate of 10 percent. The annual cost for vehicle testing to ensure vapor tightness is \$366 per vehicle in July 1989 dollars, and the maintenance charge for the vapor collection system²² is \$244 per rack in July 1989 dollars.

The costs for vapor collection and transport are combined with the costs derived for the control device (flare or incinerator) to calculate the overall total capital investment and annual costs. The costs for the combustion control devices are discussed in Section 3.1.

3.2.2 Cost Methodology for Condensers

This section discusses capital and annual costs for packaged refrigerated surface condenser systems. These costs were derived from vendor data.²³

3.2.2.1 Condenser Design Considerations Affecting Cost.

The design parameters of a refrigerated condenser system that affect cost are the same ones that affect removal efficiency because they determine the condensation temperature and heat load (refrigeration tonnage) of the system. These design parameters include the following:

- Volumetric flow rate of the VOC-containing vent stream;
- Inlet temperature of the vent stream;
- Concentrations of the VOC's in the vent stream;
- Absolute pressure of the vent stream;

TABLE 3-7. TOTAL ANNUAL COST FOR LOADING RACK
VAPOR COLLECTION SYSTEMS

Cost Item	Cost (July 1989 \$)
Direct Annual Cost	
Toll charge for N ₂ blanketing ^a	284 * (Throughput, MMgal/yr)
Electricity for fan/motor ^b	See equation in test
Indirect Annual Cost	
Capital recovery ^c	0.163 * TCI
Vehicle vapor-tight testing ^d	\$366 per vehicle
Maintenance for vapor collection system ^d	\$244 per rack

^aFrom Reference 21.

^bFrom Reference 20.

^cCapital recovery is based on an equipment life of 10 years and an interest rate of 10 percent.

^dFrom Reference 22.

- Required removal efficiency of the VOC;
- Moisture content of the vent stream; and
- Properties of the VOC's in the vent stream:
 - heats of condensation,
 - heat capacities, and
 - vapor pressures.

The cost equations in the following sections are all functions of condensation temperature and heat load.

3.2.2.2 Development of Capital Costs. This section presents the procedures and data necessary for estimating capital costs for packaged refrigerated surface condenser systems. Total capital investment includes equipment costs for the entire refrigerated condenser system, auxiliary equipment costs, taxes, freight charges, instrumentation, and installation costs. Taxes and freight charges are based on factors described in the OCCM.¹ Instrumentation is included in the equipment costs for packaged systems.

Vendors of refrigerated surface condenser systems were asked to provide cost estimates for a wide range of applications. These vendor quotes were used to develop the costs presented here. Table 3-8 lists equipment cost equations for 13 condensation temperatures.

The cost estimates apply to skid-mounted systems designed for hydrocarbon vapor recovery. The systems are operated intermittently, allowing 30 to 60 min for defrosting by circulation of warm brine. The achievable VOC removal efficiencies for these systems are 90 to 95 percent.

The packaged systems include the refrigeration unit with the necessary pumps, compressors, condensers/evaporators, and 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 costs in Table 3-8 are based on a new plant installation; no retrofit cost considerations are included because retrofit cost factors are very site specific.

Purchased equipment costs include the packaged equipment cost and factors for sales tax (3 percent) and freight

TABLE 3-8. EQUIPMENT COST EQUATIONS FOR PACKAGED
REFRIGERATED CONDENSER SYSTEMS^a

Condensation Temperature (°F)	Equipment Costs (July 1989 dollars)	Applicable Tonnage
$40 \leq T$	$1,813 R^b + 13,521$	$R > 0.85$
$30 \leq T < 40$	$2,274 R + 13,830$	$R > 0.63$
$20 \leq T < 30$	$2,926 R + 13,776$	$R > 0.71$
$10 \leq T < 20$	$3,997 R + 17,465$	$R > 0.44$
$0 \leq T < 10$	$5,016 R + 18,034$	$R > 0.32$
$-10 \leq T < 0$	$6,977 R + 16,789$	$10 > R > 0.21$
	$3,600 R + 53,834$	$R > 10$
$-20 \leq T < -10$	$9,450 R + 16,813$	$5 > R > 0.13$
	$4,475 R + 51,083$	$R > 5$
$-30 \leq T < -20$	$5,417 R + 50,111$	$R > 2.42$
$-40 \leq T < -30$	$6,824 R + 50,103$	$R > 1.92$
$-50 \leq T < -40$	$8,380 R + 49,991$	$R > 1.58$
$-60 \leq T < -50$	$8,940 R + 54,549$	$R > 1.25$
$-70 \leq T < -60$	$11,172 R + 52,141$	$R > 1.33$
$-100 \leq T < -70$	$22,248 R + 58,633$	$R > 0.67$

^aFrom References 23 and 24. ²⁴

^bR= Condenser heat load in tons of refrigeration.

(5 percent). Instrumentation and controls are included in equipment costs for the packaged units. Thus,

$$PEC (\$) = EC (1 + 0.03 + 0.05) = 1.08 * EC$$

where:

PEC = purchased equipment costs; and

EC = equipment costs.

The total capital investment for packaged systems is obtained by multiplying the purchased equipment cost by the total installation factor²⁵ (1.15):

$$TCI (\$) = 1.15 * PEC$$

where:

TCI = total capital investment; and

PEC = purchased equipment costs.

Depending on the site conditions, the installation costs for a given system could deviate significantly from costs generated by these average factors. Guidelines are available for adjusting these average installation factors.^{25,26} If an existing condenser is removed so that a more efficient condenser may be put in its place, the cost of the demolition is estimated to be \$284 based on 8 man-hours of labor.²⁷

3.2.2.3 Development of 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 3-9.

Direct annual costs (DAC) include labor (operating and supervisory), maintenance (labor and materials), and electricity. Operating labor is estimated at 1/2-hr per 8-hr shift.²⁵ Supervisory labor is assumed to be 15 percent of the operating labor cost. Maintenance labor is estimated at 1/2-hr per 8-hr shift. Maintenance material costs are assumed to equal maintenance labor costs.

Utility costs for refrigerated condenser systems include electricity requirements for the refrigeration unit and any pumps and blowers. The power required by the pumps and blowers is negligible compared with the power requirements of

TABLE 3-9. BASES AND FACTORS FOR ANNUAL COSTS
FOR REFRIGERATED CONDENSER SYSTEMS

<u>Basis for Direct Annual Costs</u>	<u>Factor</u>
Operating labor (man-hours/8-hr shift)	0.5
Maintenance labor (man-hours/8-hr shift)	0.5
Labor rates (July 1989 \$/hr) ^a	
Operating labor	13.20
Maintenance labor	14.50
Supervisory labor cost (percent of operating labor cost)	15
Maintenance materials cost (percent of maintenance labor cost)	100
Utilities (July 1989 \$) ^a	
Electricity (\$/1,000 kW-hr)	50.9
Natural Gas (\$/1,000 ft ³)	3.03
Product recovery credit	Emission Reduction * Chemical Market Price
<u>Basis for Indirect Annual Cost</u>	
Equipment life (years)	15
Interest rate (percent)	10
Capital recovery factor	0.1314
Taxes, insurance, administration (percent of total capital investment)	4
Overhead (percent of total labor and maintenance costs)	60

^aBased on Reference 3.

the refrigeration unit. Electricity requirements for refrigerated condenser systems are summarized below:

<u>Electricity (E)</u>	<u>Temperature</u>
1.3 kW/ton	$40\text{ }^{\circ}\text{F} \leq T$
2.2 kW/ton	$20\text{ }^{\circ}\text{F} \leq T < 40\text{ }^{\circ}\text{F}$
4.7 kW/ton	$-20\text{ }^{\circ}\text{F} \leq T < 20\text{ }^{\circ}\text{F}$
5.0 kW/ton	$-50\text{ }^{\circ}\text{F} \leq T < -20\text{ }^{\circ}\text{F}$
11.7 kW/ton	$-100\text{ }^{\circ}\text{F} \leq T < -50\text{ }^{\circ}\text{F}$

These estimates were developed from product literature obtained from one vendor.²³ Electricity costs can then be calculated from the following expression:

$$C_e (\$/\text{yr}) = [R(\text{tons})/0.85] * E (\text{kW}/\text{ton}) * (\text{hr}/\text{yr}) * (\$/\text{kW-hr})$$

where C_e = annual electricity costs;
 R = refrigeration requirements;
 E = electricity requirements;

and the factor 0.85 accounts for the mechanical efficiency of the compressor.

Indirect annual costs are the sum of capital recovery costs plus general and administrative (G&A), overhead, taxes, and insurance costs. Overhead is assumed to equal 60 percent of the sum of maintenance materials and operating, supervisory, and maintenance labor.

The system capital recovery cost is based on an estimated 15-year equipment life. For a 15-year life and an interest rate of 10 percent, the capital recovery factor is 0.1315. Thus,

$$C_C (\$/\text{yr}) = 0.1315 * \text{TCI}$$

where:

C_C = capital recovery cost; and
 TCI = total capital investment.

Administrative charges, taxes, and insurance are factored from TCI, typically 2, 1, and 1 percent, respectively.

If the condensed VOC can be directly reused or sold without further treatment, the credits from this operation are incorporated in the total annual cost estimates. The following equation can be used to estimate the VOC recovery credits:

$RC (\$/yr) = \text{Emission Reduction (lb/yr)} * \text{Market Price } (\$/lb)$
where:

$RC = \text{recovery credit } (\$/yr).$

Data on market prices are available from sources such as the Chemical Marketing Reporter.

Total annual costs are calculated as the sum of direct annual costs and indirect annual costs, minus recovery credits:

$$TAC = DAC + IAC - RC$$

where:

$TAC = \text{total annual costs;}$

$DAC = \text{direct annual costs;}$

$IAC = \text{indirect annual costs; and}$

$RC = \text{recovery credits.}$

3.2.3 Cost Methodology for Steam Stripping

This section discusses steam stripper design considerations affecting cost and the general methodology for developing capital and annual costs for steam strippers. A detailed example of the application of this methodology to cost a steam stripper for an example wastewater stream is given in Appendix D of this volume.

3.2.3.1 Steam Stripper Design Considerations Affecting Cost. As discussed in Chapter 2, a number of factors are involved in the design of a steam stripper: stripper configuration (direction of flow as well as tray vs. packed bed design); wastewater flow rate; steam flow rate or steam-to-feed ratio; column height and diameter; wastewater feed temperature and pH; and the vapor-liquid partitioning of compounds to be removed (expressed by Henry's Law constant). This discussion is limited to sieve tray column steam strippers with countercurrent steam flow and noncorrosive wastewater with a typical feed temperature of 35 °C (95 °F).

Factors affecting the steam stripper capital costs include the column diameter, column height, and the size of auxiliary equipment (feed tanks, feed preheater, condenser, decanter, flame arrestor, and pumps). The column diameter and the size of the auxiliary equipment are a function of the

design wastewater feed rate. The column must be wide enough to provide a desired pressure drop and liquid retention time in the column using correlations developed to prevent column flooding. As the wastewater feed rate increases, the column diameter will increase proportionally. With increases in the column diameter and the size of the auxiliary equipment, the capital costs of the system increase. The height of the steam stripper column depends upon the desired compound removal efficiency. The column height is designed to accommodate the height of trays necessary to achieve a desired compound removal efficiency. Typically, higher removal efficiencies require more trays, thereby increasing the capital cost of the system.

The steam stripper annual costs are affected most by the annual steam requirement, which is a function of the steam-to-feed ratio and the wastewater feed rate. The steam-to-feed ratio is selected to obtain the desired compound removal efficiency. Higher compound removal efficiencies generally require a greater steam-to-feed ratio and, therefore, result in higher steam costs. The steam requirements for the stripper are also a direct function of the wastewater feed rate.

3.2.3.2 Development of Steam Stripper Capital Costs.

The total capital investment for a steam stripper system includes purchased equipment cost, direct installation cost, and indirect installation cost.²⁸ The purchased equipment cost comprises the basic equipment cost, auxiliary piping and equipment cost, instrumentation, freight, and sales tax. The basic equipment cost is the sum of the price of each component of the steam stripper system and is estimated using engineering cost estimation techniques.^{5,26,29,30,31,32,33} Table 3-10 presents equations for the costs of the various components of the steam stripper system. All costs are for carbon steel construction except for sieve trays and pumps. It was assumed that these components would be constructed of stainless steel because they are subject to the greatest wear and are exposed to the harshest conditions.

TABLE 3-10. EQUIPMENT COST EQUATIONS FOR A STEAM STRIPPING UNIT

Equipment Component	Costing Equation ^a	Cost Index ^b	Qualifier ^c	Cost Reference
Feed tanks	$\exp[2.331+1.3673*\ln(v)-0.063088*\ln(v)^2]$ $\exp[11.362-0.6104*\ln(v)+0.045355*\ln(v)^2]$	230.9	1,300 gal < V ≤ 21,000 gal 21,000 gal < V < 11,000,000 gal	29 29
Feed preheater	$\exp[8.551-0.30863*\ln(A)+0.06811*\ln(A)^2]$ $(\exp[-1.1156+0.0906*\ln(A)])$	230.9	150 ft ² < A < 12,000 ft ²	29
Steam stripper	$(1A + 1B + 1C + 2A + 2B + 2C + 2D)/2$			
1A. Column shell, skirts, nozzles and manholes	$\exp[6.823+0.14178*\ln(W_S)+0.02468*\ln(W_S)^2]$	230.9	9020 lb < W _S < 2,470,000 lb	31
1B. Platforms and ladders	$151.81*D_i(0.63316)*L_t(0.80161)$	230.9	3 ft < D _i < 24 ft; 57.5 ft < L _t < 170 ft	31
1C. Sieve trays-stainless steel	$0.85*(1.189+0.0577*D)^*$ (No. of Trays)*278.38*exp[0.1739*D]	230.9	2 ft < D < 16 ft	31
2A. Column shell	83.69*W _S (0.6124)	225.9	1,000 lb < W _S < 70,000 lb	30
2B. Manholes	(No. of Manholes)*18*(53.83+40.71*TS)	225.9	0.375 inch < T _S < 2 inch	30
2C. Nozzles	each nozzle length (No. of nozzles)*(length of i=1 nozzles, inch)*[24.57+35.94*T _S]	225.9	0.375 inch < T _S < 2 inch	30
2D. Sieve trays-stainless steel	(no. of Trays)*214.54*exp[0.2075*D _i]	225.9	2 ft < D _i < 12.5 ft	30
Primary condenser	2228.8*exp[0.00411*A]	343	37.5 ft ² < A < 240 ft ²	5
	5328.8*exp[0.0008762*A]	343	240 ft ² < A < 1500 ft ²	5

TABLE 3-10. EQUIPMENT COSTS EQUATIONS FOR A STEAM STRIPPING UNIT
(CONCLUDED)

Equipment Component	Costing Equation ^a	Cost Index ^b	Qualifier ^c	Cost Reference
Overhead collection decanter	$74.55(V)^{0.5662}$	225.9	100 gal < V < 100,000 gal	32
Flame arrestor	\$100/arrestor	319	NA	33
Pumps - stainless steel	$8740.7 * W^{(0.4207)}$; W in hp	347.8	10 gpm < Q < 150 gpm	5
	$13783.4 * W^{(0.2890)}$; W in hp	347.8	30 gpm < Q < 900 gpm	5

^aAll cost equations are based on components constructed of carbon steel except the equations for sieve trays and pumps. Cost equations for these two components are based on construction with stainless steel.

^bThe July 1989 Chemical Engineering fabricated equipment cost index is 356.0. To calculate costs in July 1989 dollars, multiply the equations in the table by the ratio of 356.0 to the cost index given in the table.

^cA = surface area; D = steam stripper outside column diameter; D_i = steam stripper inside column diameter; L_t = steam stripper tangent-to-tangent length; Q = flow rate through component; T_S = steam stripper column wall thickness; V = volume of component; W = work; and W_S = steam stripper column weight.

The total capital investment required to install a new steam stripper unit is calculated as a direct function of the basic equipment cost value. The purchased equipment cost is calculated by multiplying the basic equipment cost by an appropriate percentage value. This percentage value and the other multipliers discussed below are selected from ranges recommended in cost estimation reference documents.

Piping costs are implicitly included in the direct installation cost; however, auxiliary piping (i.e., additional piping for the combination of wastewater streams and vapor vent lines for storage tanks) are accounted for separately in the purchased equipment cost. The purchased equipment cost is used to estimate the steam stripper system direct installation costs and indirect installation costs. Each of these costs is calculated by multiplying the purchased equipment cost by an appropriate percentage value. The direct installation cost includes items such as electrical wiring, insulation, equipment support and erection, and painting of equipment. The indirect installation cost includes engineering, construction and field expense, construction fee, start-up and testing, and contingency. The sum of purchased equipment cost, direct installation cost, and indirect installation cost yields the total capital investment. The total capital investment can also include costs for buildings, off-site facilities, land, working capital, and yard improvements; however, these costs are not typically included in the purchased equipment cost for a steam stripper system.

Table 3-11 summarizes the equations used in estimating total capital investment for a steam stripper system. The total capital investment for installing a new steam stripper system is presented in Figure 3-1 as a function of the system's design wastewater feed rate. The total capital investment costs for this graph were calculated using the cost equations in Table 3-11 and are based on a steam stripper system design which is sized for each wastewater flow rate. The stripper system design was developed using the ASPEN computer program. Figure 3-1 presents total capital

TABLE 3-11. COST METHODOLOGY FOR ESTIMATING TOTAL CAPITAL INVESTMENT
FOR A STEAM STRIPPING SYSTEM

Cost Component	Cost Equation	Cost Reference
<u>Direct Equipment Costs</u>		
Base Equipment Cost (BEC)	Sum of costs for feed tanks, feed preheater, steam stripping column, primary condenser, overhead collection decanter, flame arrestor, pumps. (See Table 3-10 for individual equations)	5, 26, 29, 30, 31, 32, 33
Piping (Pipe)	$\$36.83/\text{m}$ ($\$11.23/\text{ft}$)	34
Instrumentation	$0.1 * [\text{BEC} + \text{Pipe}]$	26
Sales Tax and Freight	$0.08 * [\text{BEC} + \text{Pipe}]$	26
Purchased Equipment Cost (PEC)	(BEC, Pipe, Instrumentation, Tax + Freight)	
<u>Direct Installation Costs</u>		
Foundations and Supports	12% of PEC	26
Electrical	1% of PEC	26
Erection and Handling	40% of PEC	26
Painting	1% of PEC	26
Insulation	1% of PEC	26
Total Direct Installation Cost	(Foundations and supports, electrical, erection and handling, painting, insulation)	
<u>Indirect Installation Costs</u>		
Engineering and Supervision	10% of PEC	26
Construction and Field Expense	10% of PEC	26
Construction Fee	10% of PEC	26
Startup and Testing	1% of PEC	26
Contingency	3% of PEC	26
Total Indirect Installation Cost	(Engineering and supervision, construction and field expense, construction fee, startup and testing, contingency)	
TOTAL CAPITAL INVESTMENT (TCI)	(PEC, Total Direct Installation, Total Indirect Installation)	

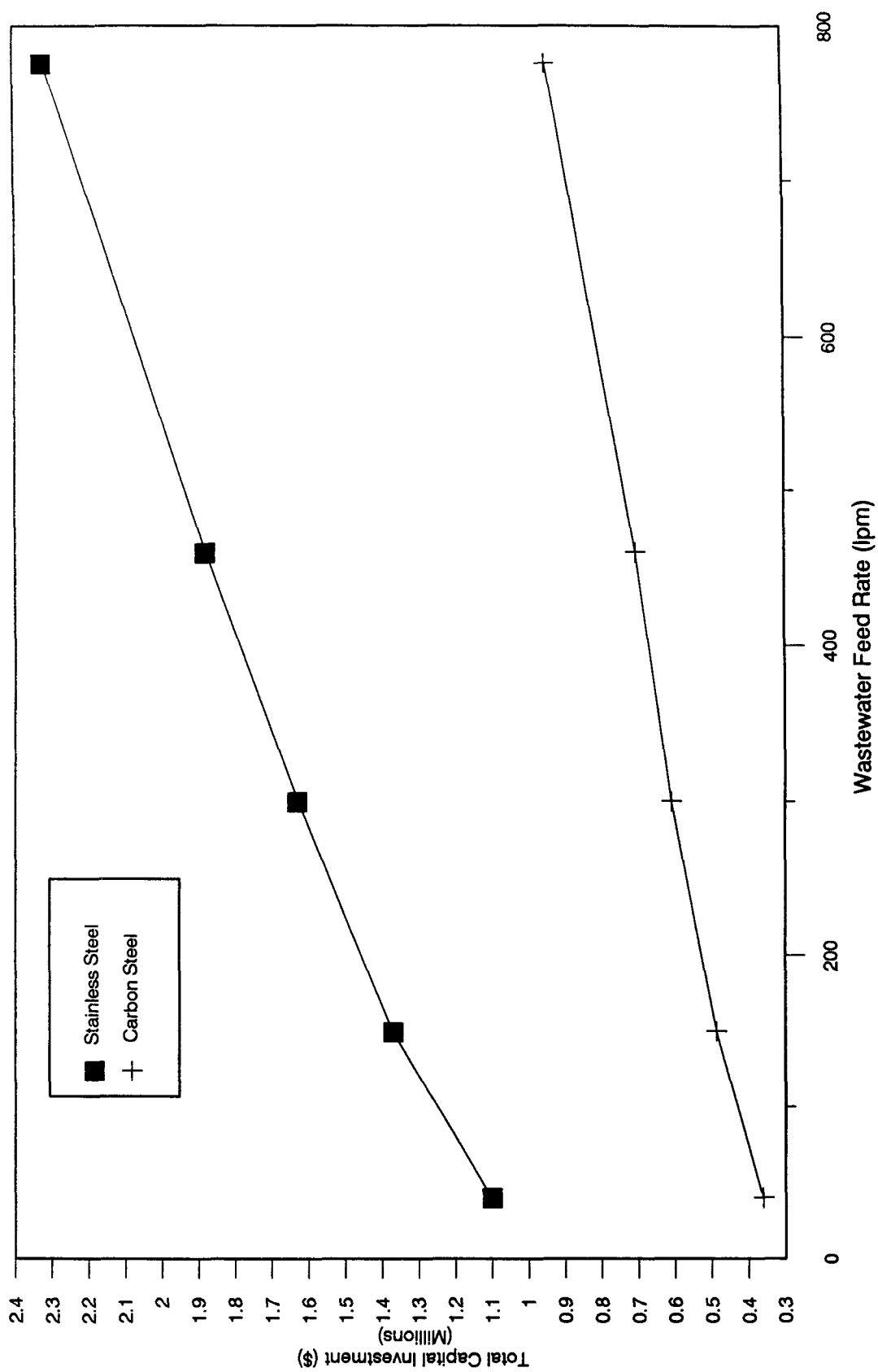


Figure 3-1. Summary of total capital investment versus wastewater feed rate for steam stripping unit.

investment for both carbon and stainless steel construction. Stainless steel construction costs are included for comparison because streams with corrosive wastewater (i.e., very high or low pH) will require a steam stripper system constructed of a corrosion resistant material. Equipment costs for stainless steel were developed from the same information sources as for carbon steel. Generally, a factor for material of construction was used for conversion of carbon steel cost to 304 stainless steel cost. As shown in Figure 3-1, the total capital investment is a direct function of the wastewater flow rate to the steam stripper unit, with stainless steel construction being approximately two to seven times more costly than carbon steel.

3.2.3.3 Development of Steam Stripper Annual Costs. The total annual cost is the total of all costs incurred to operate the steam stripper system throughout the year. The annual operating costs comprise direct and indirect charges. Direct annual costs comprise expenses incurred during normal operation of the steam stripper process, including utilities, labor, and maintenance activities.

Three types of utilities are used to operate the steam stripper process unit: electricity, steam, and cooling water. Electricity is needed to operate pumps and other electrical components in the system. The electricity required for the pumps is calculated using design flow rates for each pump and assuming a developed head of approximately 37 m (120 ft) of water and a pump efficiency of 64 percent. The steam costs are estimated using the design steam loading of 0.096 kg steam/l (0.80 lb/gal) of wastewater feed. The cooling water cost is calculated using water requirements necessary for the overhead primary condenser.

Other direct costs include labor and maintenance. Labor cost was estimated using the number of hours required to operate a steam stripper process unit (0.5 hr/shift) and a labor rate of \$13.20/hr.³ The supervisory and administrative costs are estimated as 15 percent of operating labor. The maintenance costs consist of labor and materials. The

maintenance labor cost assumes 0.5 hr/shift operation and a \$14.50/hr labor rate. The maintenance materials cost is assumed to equal 100 percent of maintenance labor cost.

Indirect operating expenses are incurred regardless of the operating status of the steam stripper system. The cost of overhead is estimated to be 60 percent of all labor and maintenance costs. The remaining components of the indirect annual costs are taken as percentages of the total capital investment. The capital recovery for the steam stripper system is based on a 15-year equipment life at an interest rate of 10 percent.

Another element of the total annual cost that is included as a benefit in this estimate is the credit for the recovered chemicals. This factor accounts for any cost credits which result from the organic chemicals being recovered from the overheads stream. There are several alternatives for handling the recovered organic chemicals. If steam is produced on-site, the recovered organic chemicals can be used as fuel for the existing boiler. The money saved by not having to purchase conventional fuels (i.e., fuel oil or natural gas) is the recovery credit. In this situation, the value of the recovered chemicals is equal to the fuel value only. Another option is to reuse the recovered chemicals in the manufacturing process. In some cases the organic chemicals can be recycled directly to the process; in other cases the organic chemicals must be separated by distillation before reuse. The costs saved in reducing the purchase of raw materials is the recovery credit and is valued at the cost of the chemicals; however, this cost savings may be offset by the cost of distilling the recovered organic chemicals. Another option is to sell the recovered organics to a chemical manufacturer who will recover the individual components in the waste organic stream. However, in cases where a cost-effective use for the recovered organics does not exist, the plant will have to pay for disposal of the collected organic chemicals. There will be no cost savings in this case; in fact, an additional cost for disposal may be incurred.

This cost estimation methodology assumes that the recovered organic chemicals are used as fuel for an existing boiler rather than recycled to the process or sold to a chemical manufacturer. To calculate the recovery credit, the heat content of the recovered chemicals was estimated based on their composition.

Table 3-12 summarizes the equations discussed above which are used to estimate total annual cost for a steam stripper system. The annual unit operating cost (\$/ℓ) for the steam stripper is calculated by dividing the total annual cost (\$/yr) by the annual wastewater throughput (ℓ/yr). The annual unit operating cost for installing and operating a steam stripper system is compared to the system's design wastewater feed rate in Figure 3-2. These annual unit operating costs were calculated using the cost equations in Table 3-11 and are based on a stripper system design which is sized for each wastewater flow rate. The stripper system design was developed using the ASPEN computer program. The costs for both carbon and stainless steel construction are presented. As shown in Figure 3-2, the annual unit operating cost is an indirect function of the wastewater feed rate to the steam stripper unit, with stainless steel construction being approximately 1.5 times more costly than carbon steel.

3.3 COST METHODOLOGY FOR STORAGE TANK IMPROVEMENTS

3.3.1 Design Considerations Affecting Cost

As described in Chapter 2, there are many parameters affecting the emission control efficiency of internal floating roof tanks. Some of the parameters are chemical specific while others are related to the floating deck. The deck design considerations that affect cost are:

- The number and type of rim seals;
- The type of deck fittings (i.e., controlled or uncontrolled); and
- The type of deck seams (i.e., bolted or welded).

For this analysis, when a new floating roof is installed on a fixed roof tank, it is assumed that the deck is bolted and has a liquid-mounted primary seal and controlled fittings.

TABLE 3-12. COST METHODOLOGY FOR ESTIMATION OF TOTAL ANNUAL COST
FOR A STEAM STRIPPING SYSTEM

Cost Component	Cost Equation ^a	Cost Reference
<u>Direct Annual Costs</u>		
Utilities		
Electricity	\$ 0.0509/kWhr	35
Steam	\$ 7.68/Mg (\$7.77/ton)	35
Water	\$ 0.058/1,000 l (\$0.22/1,000 gal)	36
Labor		
Operating labor	\$ 13.20/hr	26
Supervision and administration	15 percent of operating labor	26
Maintenance		
Labor	\$ 14.50/hr	26
Materials	100 percent of maintenance labor	26
Total Direct Annual Costs	(Utilities, labor, maintenance)	
<u>Indirect Annual Costs</u>		
Overhead	60 percent of all labor and materials	26
Property taxes	1 percent of TCI	26
Insurance	1 percent of TCI	26
Administrative charges	2 percent of TCI	26
Capital recovery	10 percent at 15 years	26
Total indirect annual costs	(Overhead, property taxes, insurance, administrative charges, capital recovery)	
Recovery credit	Estimate based on what can be done with recovered organics	
TOTAL ANNUAL COST	TDAC + TIAC - RC	

^aTCI = total capital investment; TDAC = total direct annual costs; TIAC = total indirect annual costs; and RC = recovery credit.

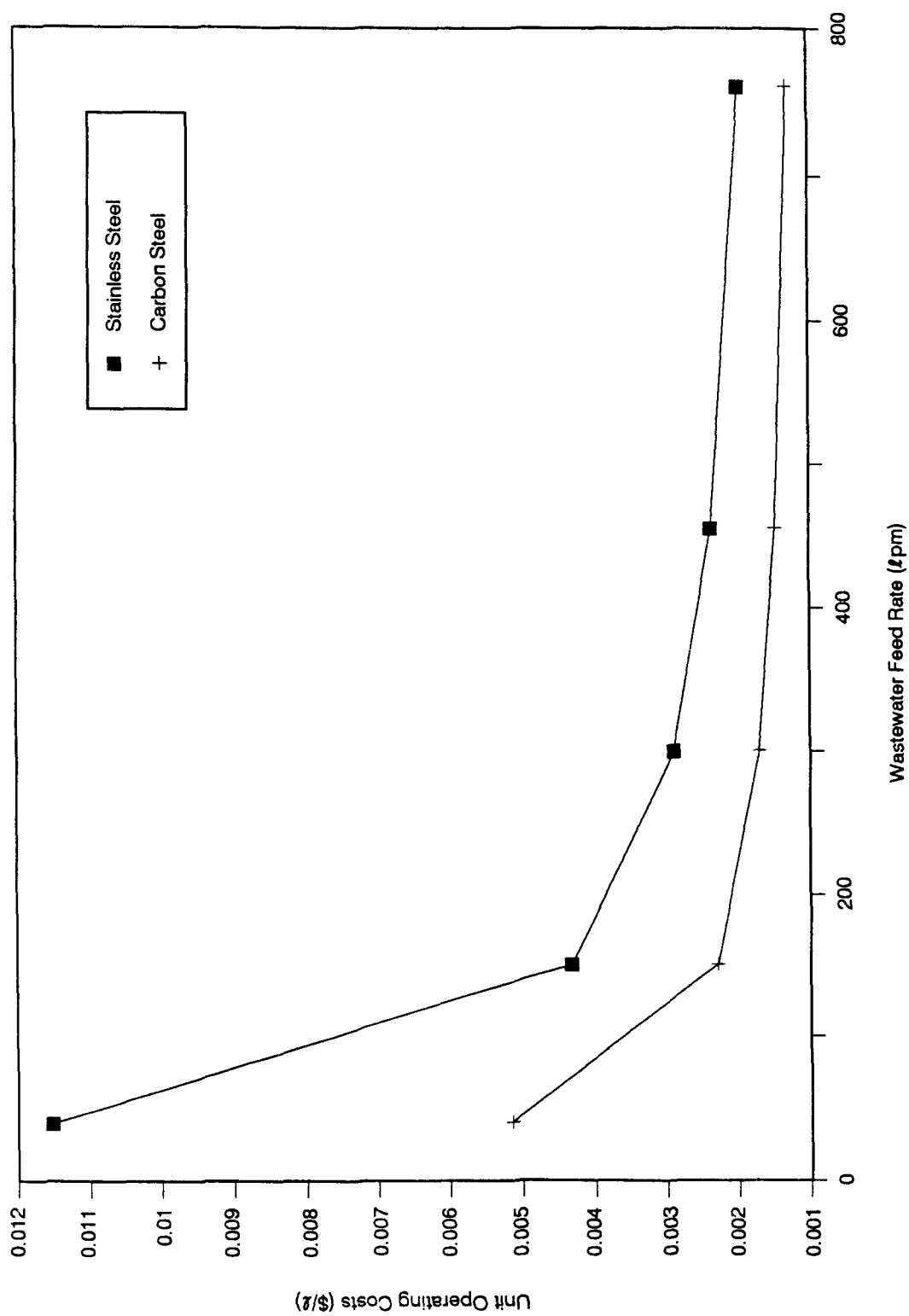


Figure 3-2. Unit operating costs versus wastewater feed rate for steam stripping unit.

Typical existing internal floating roofs have vapor-mounted primary seals and uncontrolled deck fittings. Some tanks also have secondary seals. These existing internal floating roof tanks are assumed to be upgraded by the addition of a secondary seal (when necessary) and controlled deck fittings.

The costs presented in this section were taken from the EPA report "Control of Volatile Organic Compound Emissions from Volatile Organic Liquid Storage in Floating and Fixed Roof Tanks,"³⁷ which contains costs and cost equations for several different deck designs. For consistency with other sections in this chapter, these costs have been updated to July 1989 dollars.

3.3.2 Development of Capital Costs

3.3.2.1 Installing a Floating Roof. Before an internal floating roof can be installed, the tank must be emptied, cleaned, and degassed. During degassing, the tank is emptied of all VOC vapors by replacing the VOC-laden air in the tank with fresh air. Fans or blowers are used to draw in the fresh air and exhaust the VOC vapors. The cost for this activity in terms of tank size is as follows:

$$\begin{array}{l} \text{Cleaning/} \\ \text{Degassing (July 1989 \$)} = 7.61 * (\text{Tank Size [gal]})^{0.5132} \\ \text{of tank} \end{array}$$

The cost of installing a floating roof depends upon the type of deck and seal system selected. The following equation estimates the cost of installing a new bolted, floating deck having a liquid-mounted primary seal and controlled deck fittings:

$$\begin{array}{l} \text{New Floating} \\ \text{Roof} \\ \text{(July 1989 \$)} \end{array} = 509 * (\text{Tank Diameter [ft]}) + 1160$$

This equation includes the retrofit cost of cutting vents or openings necessary for modifying the tank.

In some cases, an existing refrigerated condenser system may be removed before an internal floating roof is installed. The cost of this activity is estimated to be \$284 per tank based on 8 man-hours of labor.²⁷

3.3.2.2 Upgrading an Existing Internal Floating Roof Tank. Typical existing internal floating roof tanks have a bolted deck, vapor-mounted primary seal and uncontrolled deck fittings. To reduce emissions from such tanks, the deck may be upgraded by adding a secondary seal and controlled deck fittings (see Table 2-2).

The cost of adding a secondary seal to an existing floating roof is given in terms of tank diameter:

Addition of
a Secondary Seal (July 1989 \$) = $95.1 * (\text{Tank Diameter [ft]})$

The cost of controlled fittings is about 4 percent of the cost of a new deck having a vapor-mounted primary seal.³⁸ In terms of tank diameter, this cost is calculated as follows:

Addition of
Controlled Deck Fittings (July 1989 \$) = $16 * (\text{Tank Diameter [ft]}) + 46$

For this analysis, it is assumed that existing internal floating roof tanks will not be upgraded until a scheduled cleaning and degassing. Thus, the cleaning/degassing cost will not be a part of the control costs required for upgrading existing internal floating roof tanks.

3.3.3 Development of Total Annual Costs

The annual cost without product recovery credit is the sum of annualized capital cost, operating costs, and costs for taxes, insurance, and administration. Assuming an equipment life of 10 years and an interest rate of 10 percent, the capital recovery factor equals 0.163. Operating costs include the yearly maintenance charge of 5 percent of the capital cost and an inspection charge of 1 percent of the capital cost. Taxes, insurance, and administration are assumed to equal 4 percent of the capital cost.³⁹

The total annual cost, TAC, also accounts for the value of any recovered product. Product recovery credit is calculated by multiplying the market value of the chemical by the emission reduction achieved by the tank improvements.

The equation below summarizes the calculation of total annual cost. The first term represents the capital recovery costs, operating costs, and costs for taxes, insurance, and administration and the second term represents product recovery credit:

$$\text{TAC (July 1989 \$)} = (\text{Capital Costs} * 0.263) - (\text{Emission Reduction} * \text{Market Price})$$

where:

TAC = total annual cost.

3.4 COST METHODOLOGY FOR EQUIPMENT LEAK CONTROL TECHNOLOGIES

Emissions from different equipment types discussed in Chapter 2 can be reduced by either control equipment or work practices. Costs for reducing emissions from compressors, open-ended lines, sampling connections, and pressure relief devices are developed assuming control equipment will be installed. Costs for reducing emissions from pumps, valves, and connectors are developed assuming emissions will be reduced through work practices. The general methodology is developed to determine base costs for the control of emissions for each type of equipment. These base costs per component are then used with plant equipment counts to develop capital and annual costs for the control of emissions from equipment leaks for each plant.

3.4.1 Control Equipment

This section presents the costs associated with the purchase and installation of specified equipment for the control of VOC emissions from compressor seals, open-ended lines, sampling connections, and pressure relief devices. The costs for controlling emissions from product accumulator vessels with a combustion device are the same as the costs for controlling emissions from process vents, which are described in Section 3.1.

The base costs for all of these devices were developed assuming the same equipment requirements as stated for these devices in the Additional Information Document for Fugitive Emission Sources of Organic Compounds (SOCMI Fugitives AID).⁴⁰

Base cost information presented in the Background Information Document for Benzene Fugitive Emissions (Benzene BID) was also reviewed.⁴¹ The base cost estimates included here were made using Richardson Process Plant Construction Estimation Standards⁴² where possible and by contacting vendors for current prices. It was not possible to use these estimation techniques for all control equipment, however. The prices for control equipment for which estimates or quotations were not available were updated from the SOCFI Fugitives AID. All costs were converted to July 1989 dollars using the ratio of the CE Plant Cost Indices. These base cost estimates are presented in Table 3-13.

3.4.1.1 Compressors. The equipment for which cost estimates were developed for compressor seals includes a heavy liquid or non-VOC barrier fluid with a degassing reservoir which is connected by a closed vent system to an additional control device (e.g., flare) or vapor recovery header.⁴³

The base cost is calculated for the purchase of 122 m of 2-in. schedule 40 steel pipe, three 2-in. steel plug valves, and one metal gauze flame arrestor for each compressor.⁴⁰ The costs for piping and valves are estimated to be \$1,090 and \$2,440, respectively.⁴⁴ The cost for a metal gauze flame arrestor is estimated to be \$862.⁴⁵ A total of 82 hr of maintenance labor would be required for installation. The labor cost is estimated to be \$22.50/hr.^{46,47} The total installed base cost for a closed vent system is \$6,242 per compressor.

These costs are based on connecting the closed vent system to an existing enclosed combustion device or vapor recovery header and do not reflect the cost of adding a control device specifically to control the degassing vents. The actual base cost will vary depending on the length of pipe and number of valves required for each compressor within each plant.

3.4.1.2 Open-Ended Lines. A cap, plug, blind flange, or second valve will be installed on all open-ended lines to prevent emissions through the open end. The base cost is

TABLE 3-13. BASE COST DATA FOR EQUIPMENT LEAK
CONTROL DEVICES^a

Equipment Type: Item	Cost (July 1989 \$)	Basis
Compressors: closed vent system for degassing reservoirs		
• 122 m of 5.1-cm schedule 40 steel pipe	1,090	\$261.03/100 ft pipe CE Index(356.0/342.5)
• Installation (66 hr)	1,490	16.6 hr/100 ft pipe Labor \$22.50/hr
• Three 5.1-cm cast steel plug valves	2,440	\$783.75/valve CE Index(356.0/342.5)
• Installation (6 hr)	135	2 hr/valve Labor \$22.50/hr
• One metal gauze flame arrestor	862	10/90 quote of \$869.00 CE Index(356.0/358.7)
• Installation (10 hr)	225	Labor \$22.50/hr
TOTAL:	6,242	
Open-ended lines:	70	\$67.50/valve
One 2.5-cm gate valve		CE Index(356.0/342.5)
• Installation (1.4 hr)	32	1.4 hr/valve Labor \$22.50/hr
TOTAL:	102	
Sample connections: closed purge system		
• 6 m of 2.5-cm schedule 40 steel pipe	25	\$120.64/100 ft pipe CE Index(356.0/342.5)
• Installation (1 hr)	23	12.6 hr/100 ft Labor \$22.50/hr
• Three 2.5-cm ball valves, carbon steel	267	\$85.60/valve CE Index(356.0/342.5)
• Installation (6 hr)	95	1.4 hr/valve Labor \$22.50/hr
TOTAL:	409	

TABLE 3-13. BASE COST DATA FOR EQUIPMENT
LEAK CONTROL DEVICES^a (CONCLUDED)

Equipment Type: Item	Cost (July 1989 \$)	Basis
Pressure relief seals (new unit): rupture disk assembly		
• One 7.6-cm stainless steel rupture disk	78	10/90 quote of \$235/3-pak CE Index (356.0/358.7)
• One 7.6-cm carbon steel rupture disk holder	579	10/90 quote of \$583/holder CE Index (356.0/358.7)
• One 0.6-cm pressure gauge, dial face	24	Update from SOCM I NSPS AID CE Index (356.0/224.7)
• One 0.6-cm bleed valve, carbon steel	40	Update from SOCM I NSPS AID CE Index (356.0/224.7)
• Installation (16 hr)	360	Labor \$22.50/hr
• One 7.6-cm gate valve	401	\$385.60/valve CE Index (356.0/342.5)
• Installation (1.5 hr)	34	1.5 hr/valve Labor \$22.50/hr
• Offset: one 10.2- cm tee and one 10.2-cm elbow	49	\$35.63/tee, \$11.73/elbow CE Index (356.0/342.5)
• Installation (8.5 hr)	191	5.7 hr/tee, 2.8 hr/elbow Labor \$22.50/hr
TOTAL:	1,755	
Pressure relief seals (retrofitted unit): new assembly plus a new relief valve.	1,755	
• One 7.6-cm stainless steel body and trim relief valve	1,950	Update from SOCM I NSPS AID CE Index (356.0/224.7)
• Installation (10 hr)	225	Labor \$22.50/hr
TOTAL:	3,930	

^aCosts and hours of installation from Reference 46.

estimated based on the purchase price of a 1-in. gate valve of \$70 and 1.4 hr of labor for installation for each open-ended line.⁴⁶ The installed base cost for each open-ended line is estimated to be \$102, including labor.

The actual base cost for each open-ended line will vary based on the type of closure chosen. Caps, plugs, and blind flanges are all expected to cost less than a second valve for the same size line but cannot be used in all cases. For example, blind flanges can only be installed on lines larger than 1 in. in diameter.

3.4.1.3 Sampling Connection Systems. Sampling connection systems should be equipped with closed-loop purge systems to prevent losses of process fluid to the atmosphere. The base cost for each closed-loop purge system is based on the installation of 6 m of 1-in. schedule 40 steel pipe and three 1-in. carbon steel ball valves.⁴⁰ Base cost estimates of the piping and valves are \$25 and \$267, respectively.⁴⁴ The estimated total base cost after installation is \$409 per sample connection.

3.4.1.4 Pressure Relief Devices. The emissions from pressure relief valves in gas/vapor service can be controlled by installing a rupture disk in the line upstream from the relief valve. This requires the purchase and installation of a rupture disk and a rupture disk holder for each pressure relief valve. The purchase cost of these components are estimated at \$78 and \$579, respectively.⁴⁸

Performing maintenance on the rupture disk assembly during plant operation requires a valve upstream from the rupture disk, a pressure gauge, and a bleed valve.⁴⁰ The valve is estimated to cost \$401.⁴⁴ The pressure gauge and bleed valve are estimated to cost \$24, and \$40, respectively.⁴⁰ To protect the pressure relief valve from damage caused by fragments of the rupturing disk at the time of pressure relief, a piping offset is typically installed.⁴⁰ This offset comprises one 3-in. tee costing approximately \$36 and one 3-in. elbow costing approximately \$12.⁴⁰ Installing these components is estimated to require 26 hr, for a total

installed cost estimate of \$1,755 per assembly. This estimated base cost per pressure relief device is for new plants.

Existing plants which are required to retrofit pressure relief valves with this new equipment also typically require a new pressure relief valve. A 3-in. stainless-steel body and trim relief valve costing \$1,950 and requiring 10 hours to install is assumed.⁴⁰ This increases the total estimated base cost for existing plants to \$3,930 per pressure relief device.

3.4.2 Leak Detection and Repair Techniques

This section presents the costs associated with the initiation and continuation of a leak detection and repair program for the control of VOC emissions from pumps, valves, and connectors. Because leak frequency data for agitators are insufficient, no estimates of repair costs were made for this analysis. The base costs for this program are discussed in two parts: the leak detection program and the repair program.

3.4.2.1 Leak Detection Program. The various methods used to identify equipment leaks are discussed in Section 2.4. This section discusses the method used to develop base costs for a leak detection program based on equipment counts and monitoring frequencies. In the development of these base costs, only the methods of an individual component survey by instrument inspection and by visual inspection are considered. The procedure used to estimate the number of leaks detected annually based on equipment counts, monitoring frequencies, and leak frequencies is also discussed in this section. "Leak frequency" is the fraction of all pieces of equipment of each type which have screening values above a given action level, measured at any given time. Leak frequency must be considered in the context of this action level. For example, leak frequencies provided in the SOCMF Fugitives AID were based on an action level of 10,000 ppmv.

For the purposes of cost estimation, vendors who perform leak detection monitoring by instrument inspection were contacted.^{49,50} Based on the information provided by these contractors, the average cost of the first individual

component survey was determined to be \$2.50 per component monitored. An average cost of \$2.00 per component monitored per survey was determined for all surveys after the initial survey. The cost of the initial survey is higher because of difficulties in initially locating all components, increased numbers of leaking components, and increased time needed to number and tag equipment.^{49,50} The cost of an initial individual component survey will be different for each process unit. The cost will depend on the number of equipment components to be monitored and the actual initial leak frequencies of the process unit.

The monitoring costs cover the equipment rental and labor charges for a contractor's survey team to "screen" each piece of equipment using a portable monitoring instrument to determine which pieces of equipment require maintenance. This monitoring team would also tag leaking components and make a first attempt to repair the leak by tightening the packing gland on valves or by some other technique which does not require removing the equipment from service. If a process unit chose to conduct monitoring internally, it is assumed the cost would not be greater than the contracted cost.

For the purpose of cost estimates, different monitoring frequencies for each type of equipment being surveyed are assumed. The monitoring frequency assumed is monthly for pumps, agitators, and valves, and annually for connectors. These frequencies may vary from plant to plant, based on the number of pieces of equipment that have been identified for maintenance at each plant. For example, if a plant has under 2 percent of its valves requiring maintenance, that plant may monitor valves quarterly.⁴³

It is assumed for this cost analysis that each process unit is required to purchase one portable monitoring instrument. Very large process units may require more than one instrument, but plants with multiple process units may not need to purchase one for each process unit. This instrument would serve as a backup to the contractor's instrument and would be used to verify that components identified as

requiring repair by the contractor have been repaired. The cost of each monitoring instrument adjusted to July 1989 dollars is estimated at \$6,500.⁵¹

3.4.2.2 Repair Program. This section discusses the method used to develop base costs for the repair of equipment determined to be leaking by a monitoring team but not successfully repaired during initial repair efforts. These leaking components are required to be repaired within 15 days of leak detection. The costs for leak repair are based on the leak repair costs estimated in the SOCMF Fugitives AID.⁴⁰ These base costs can be used with average leak frequencies and equipment counts to determine an annual operating cost for leak repairs.

The average number of leaks detected by the leak detection program can be estimated based on the number of components monitored, the monitoring frequency, and the leak frequency. An estimate of the average number of leaks detected is necessary to estimate the cost of a repair program. By multiplying the leak frequency for each type of equipment by the number of components, the estimated number of leaks can be determined. Leak frequencies are estimated for the initial leak detection survey and for all subsequent surveys.

Initial leak frequencies for valves and connectors, based on the number of pieces of equipment with screening values over 10,000 ppmv, were estimated from data in the EPA 24-Unit Study.⁵² Initial leak frequencies for gas valves, light liquid valves, and connectors are 11.4 percent, 6.5 percent and 2.1 percent, respectively. The initial leak frequency for pumps is assumed to be two times the base performance level allowed under Phase III of the negotiated regulation. The leak frequency for pumps from the 24-Unit Study was not used in this analysis because the value (based on 10,000 ppmv) was below the final performance level for pumps provided in the negotiated regulation. Using the leak frequency from the 24-Unit Study would underestimate the number of pump seals to be repaired, because the action level under the negotiated

regulation is reduced to 1,000 ppmv. The frequencies for all surveys subsequent to the initial survey are estimated to be the base performance levels allowed under the final phase of the negotiated regulation, which are 2 percent for all valves, 0.5 percent for connectors, and 10 percent for pumps.⁴³

All of the costs and assumptions discussed above for the leak detection program are summarized in Table 3-14. These values can be substituted into the equations presented in Table 3-15 to determine the cost of the leak detection program. Repair of each equipment type is described below.

3.4.2.2.1 Pumps. The SOCMF Fugitives AID presents costs for leak detection and repair of pump seals. The LDAR model was used to estimate the number of pump seals with screening values above 10,000 ppmv, which would therefore require repair. Because repair of pump seals demands more extensive maintenance efforts than possible by a monitoring team, follow-up repair was assumed for 100 percent of the pump seals predicted to have screening values over 10,000 ppmv. This assumption is not consistent with the analysis needed to estimate cost impacts of the negotiated regulation.

According to the negotiated regulation, the leak definition for pump seals will be 1,000 ppmv of VOC.⁴³ However, the negotiated regulation also states that only pump seals with leakage through the seal with a monitored concentration of greater than or equal to 2,000 ppmv of VOC will require repair. Therefore, this analysis of cost impacts considers the number of pump seals with leakage equivalent to a monitored concentration above 2,000 ppmv.

To estimate the actual percentage of leaking pump seals that will require repair, screening value data from 19 ethylene oxide and butadiene plants were evaluated.⁵³ By counting the number of leaking pumps that had screening values greater than 2,000 ppmv VOC, it was determined that approximately 75 percent of all leaking pumps would require subsequent repairs.

It is assumed that all pumps requiring repair will require a new seal.⁴⁰ The cost for each replacement single

TABLE 3-14. BASE COSTS AND ASSUMPTIONS FOR A
LEAK DETECTION PROGRAM

Subcontractor monitoring fee:	
Initial monitoring	\$2.50/component
All subsequent monitoring	\$2.00/component
Monitoring frequencies:	
Pumps	Monthly
Connectors	Annually
Valves	Monthly
Initial leak frequencies:	
Pumps	20 percent
Connectors	2.1 percent
Gas/vapor valves	11.4 percent
Light liquid valves	6.5 percent
Leak frequencies (subsequent to initial monitoring):	
Pumps	10 percent
Connectors	0.5 percent
Gas/vapor valves	2 percent
Light liquid valves	2 percent
Visual inspection of pumps:	
Time	30 sec/pump
Labor charge	\$22.50/hr
Organic vapor detection instrument:	\$6,500

TABLE 3-15. EQUATIONS FOR DETERMINING COSTS AND NUMBER OF
LEAKS FOR A LEAK DETECTION PROGRAM

Initial Monitoring Costs:

$$\text{Cost of Monitoring} = \left(\begin{array}{c} \text{Number of Pieces} \\ \text{of Equipment} \end{array} \right) * \left(\begin{array}{c} \text{Subcontractor} \\ \text{Monitoring Fee} \end{array} \right)$$

Subsequent Monitoring Costs:

$$\begin{array}{c} \text{Annual} \\ \text{Cost} \\ \text{of} \\ \text{Monitoring} \end{array} = \left(\begin{array}{c} \text{Number of} \\ \text{Pieces of} \\ \text{Equipment} \end{array} \right) * \left(\begin{array}{c} \text{Subcontractor} \\ \text{Monitoring} \\ \text{Fee} \end{array} \right) * \left[\begin{array}{c} \text{Monitoring} \\ \text{Frequency} \\ \text{(x/yr)} \end{array} \right]$$

Initial Monitoring Number of Leaks:

$$\text{Initial Leaks} = \left(\begin{array}{c} \text{Number of Pieces} \\ \text{of Equipment} \end{array} \right) * \left(\begin{array}{c} \text{Leak} \\ \text{Frequency} \end{array} \right) * \left[\begin{array}{c} \text{Monitoring} \\ \text{Frequency (x/yr)} \end{array} \right]$$

seal is estimated to be \$180, which includes a 50 percent credit for the old seal.⁴⁰ The cost of a single seal is used for estimating replacement seal costs for all pumps. Double mechanical seals are not used because they are more expensive and they could be exempted from routine leak detection and repair by using a barrier fluid system with degassing reservoir connected to a closed vent system. The time required for these repairs is estimated to be 16 hr for each pump.⁴⁰ The labor costs for this repair time is estimated at \$22.50/hr.^{46,47}

3.4.2.2.2 Valves. In the SOCMF Fugitives AID, it was assumed that 25 percent of all valves identified as leaking would require repairs beyond any initial efforts made by the monitoring team. It was assumed that the remaining 75 percent of the leaking valves had been repaired by the initial efforts of the monitoring team.⁴⁰ This assumption has been retained for this analysis.

The time of the additional repairs is estimated to be 4 hours for each valve.⁴⁰ The labor costs for this repair time is estimated at \$22.50/hr.^{46,47} Costs for replacement valve seals are not included here. Replacement seals are considered to be covered by routine plant maintenance because these costs are typically low and not all valves requiring repair will require replacement seals.

3.4.2.2.3 Connectors. In the SOCMF Fugitives AID, it was assumed that 25 percent of all connectors identified as leaking would require repairs beyond any initial efforts made by the monitoring team. It was assumed that the remaining 75 percent of the leaking connectors had been repaired by the initial efforts of the repair team. Although the leak definitions have changed, this assumption has been retained for this analysis.

The time required for additional repairs is estimated to be 2 hours for each connector.⁴⁰ The labor cost for this repair time is estimated at \$22.50/hr.^{46,47} Costs for replacement connector seals are not included here. Replacement seals are considered to be covered by routine

plant maintenance because these costs are typically low and not all connectors requiring repair will require replacement seals.

3.4.3 Capital Costs

The capital costs for the control of emissions from equipment leaks include capital expenditures for all control equipment and the purchase cost of a portable organic vapor detection instrument. The cost of the initial leak detection and repair program is also included as a capital expense because the higher costs of the initial monitoring are only incurred in the first year.

The capital costs for control equipment are estimated by multiplying the base cost for each type of equipment (Table 3-13) by the equipment count for that type of equipment in the process unit. The purchase cost of a portable organic vapor detection instrument is \$6,500.⁵¹ The initial LDAR program costs are developed as discussed in Section 3.4.2 of this volume. An additional 40 percent is also added to the initial leak detection and repair program costs to cover administration and support expenses.⁴⁰

3.4.4 Annual Costs

The annual costs for the control of VOC emissions from equipment leaks have been divided into five sections: annualized capital costs, annual maintenance charges, annual miscellaneous charges, costs for leak detection and repair, and recovery credits. Each of these sections is discussed in detail below.

3.4.4.1 Annualized Capital Costs. The annualized capital costs are calculated by taking the appropriate factor from Table 3-16 and applying it to the corresponding capital cost. The capital recovery factors are calculated using the equation:

$$CRF = \frac{i(1 + i)^n}{(1 + i)^n - 1}$$

where:

CRF = capital recovery factors;

TABLE 3-16. DERIVATION OF ANNUAL COSTS FOR CONTROL
OF EQUIPMENT LEAKS^a

Capital recovery factor for
capital charges

• Rupture disks	0.58	* Capital
• Monitoring instruments	0.23	* Capital
• Other control equipment	0.163	* Capital
• Initial labor	0.163	* Costs
• Initial replacement seals	0.58	* Costs

Annual maintenance charges

• Control equipment	0.05	* Capital
• Replacement pump seals	Estimated Number of Leaks per Process Unit * 0.75 * \$180/Seal	
• Monitoring instrument	\$4,280	

Annual miscellaneous charges
(taxes, insurance,
administration)

• Control equipment	0.04	* Capital
• Monitoring instruments	0.04	* Capital
• Replacement pump seals	0.80	* Annual Maintenance Charge

Annual operating costs

• Contractor monitoring fee	\$2.00/Component/Monitoring
• Subsequent repair labor charges	22.50/Hr
• Administrative and support	1.4 * (Contractor Fee + Subsequent Repair Charges)

Recovery credits

• VOC raw material/product cost	Market Value ^b * Mass of VOC Saved
• HAP raw material/product cost	Market Value ^b * Mass of HAP Saved

^aFrom Reference 46.

^bDepends on the specific chemicals used and the composition of
the process stream.

i = interest rate, expressed as a decimal; and
n = economic life of the component, in years.

The interest rate is 10 percent. The expected life of the monitoring instrument is 6 years, compared to 10 years for control equipment. Rupture disks and pump seals are assumed to have a 2-year life.⁴⁰

3.4.4.2 Annual Maintenance Charges. The annual maintenance charge for control equipment is calculated by multiplying the appropriate cost by 5 percent.⁴⁶ This cost includes the expenses for labor, materials, and supervision to keep the control equipment in efficient operating condition.⁴⁰ The annual maintenance charge for the portable monitoring instrument is \$4,280/yr.⁴⁶ This cost was updated from the SOCFI Fugitives AID using the CE Plant Cost Index ratio.⁴⁶

The cost of replacement pump seals for the LDAR program is considered to be a maintenance expense. It is calculated by multiplying the replacement seal cost of \$180 per seal by the number of pump leaks repaired per year.

3.4.4.3 Annual Miscellaneous Charges. The annual miscellaneous charges cover taxes, insurance, administration, and other fees associated with operations. The miscellaneous charges for control equipment and for the portable organic vapor detection instrument are calculated by applying the factor of 4 percent from Table 3-16 to the appropriate capital costs.⁴⁶ The miscellaneous charge for replacement pump seals is calculated as 80 percent of the annual maintenance charge for pump seals. The calculation is based on the miscellaneous charge for control equipment being 4 percent of the capital cost and the annual maintenance charge being 5 percent of the capital cost. The miscellaneous charge is 80 percent of the annual maintenance charge.

3.4.4.4 Leak Detection and Repair Operating Costs. The annual operating costs for the LDAR program are developed as discussed in Section 3.4.2. The cost of administration and support of the LDAR program is estimated as an additional 40 percent of the leak detection and repair costs. This

factor is included in Table 3-16, which summarizes the annual LDAR operating costs.

3.4.4.5 Recovery Credits. The recovery credit is the value of the VOC which would have been lost through emissions if equipment leaks were not controlled. This represents a savings to the process unit because recovery will reduce the amount of raw material to be purchased or will increase the amount of product to be sold, at no additional expense to the process unit.

The recovery credit is calculated by multiplying the raw material/product cost by the amount of the reduction in VOC emissions from the process unit. This raw material/product cost and amount of pollutant emissions decrease will be different for each process unit and must be determined specifically for each process unit.

3.5 REFERENCES

1. U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards. OAQPS Control Cost Manual. Fourth Edition. EPA-450/3-90-006. Research Triangle Park, NC. January 1990.
2. Code of Federal Regulations, Title 40 Part 60.18. General Control Device Requirements. Washington, DC. U.S. Government Printing Office. January 21, 1986.
3. Memorandum from Zukor, C., Radian Corporation, to HON project file. August 23, 1991. Documentation of cost factors used by the HON project.
4. U.S. Environmental Protection Agency, Air and Energy Engineering Research Laboratory. Handbook--Control Technologies for Hazardous Air Pollutants. EPA-625/6-86-014. Research Triangle Park, NC. September 1986.
5. Hall, R.S., M.W. Vataavuk, J. Matley. Estimating Process Equipment Costs. Chem. Eng. November 21, 1988. p. 66-75
6. Vataavuk, W. Pricing Equipment for Air Pollution Control. Chem. Eng. May 1990. p. 126-130.
7. Telecon. Stone, D.K., Radian Corporation, with Dowd, E. ARI Technology. January 18, 1990. Incinerator sizes and turndown.
8. Blackburn, J.W. (IT Envirosience). Control Device Evaluation: Thermal Oxidation. In: Organic Chemical Manufacturing, Volume 4: Combustion Control Devices. Report 1. U.S. Environmental Protection Agency, Research Triangle Park, NC. Publication No. EPA-450/3-80-026. December 1980. pp. III-1 to III-5.
9. Memorandum from Scott, K., Radian Corporation, to HON project file. February 5, 1992. Composition of design molecule.
10. Memorandum and attachments from Farmer, J.R., EPA/ESD, to Ajax, B. et al. August 22, 1980. Thermal incinerators and flares.
11. Danielson, John, A. Air Pollution Engineering Manual, Second Edition. Air Pollution Control District County of LA. U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards. Research Triangle Park, NC. May 1973.

12. Treybal, R.E. Mass Transfer Operations, Third Edition. New York, McGraw-Hill Book Company.
13. Memorandum from Barbour, W., Radian Corporation, to HON project file. April 20, 1990. Estimating liquid and vapor schmidt numbers for acid streams.
14. Memorandum from Ferrero, B., Radian Corporation to HON project file. February 5, 1992. Estimating liquid to vapor flow rate ratios in scrubber columns.
15. Memorandum from Ferrero, B., Radian Corporation, to HON project file. February 5, 1992. Development of the slope of the equilibrium line and the absorption factor for acid gas scrubber design.
16. Memorandum. Probert, J.A., Radian Corporation, to HON project file. September 4, 1991. Cost equations for fan and motor in vapor collection system for transfer loading racks.
17. U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards. Bulk Gasoline Terminals --Background Information for Promulgated Standards. EPA-450/3-80-038b. Research Triangle Park, NC. August 1983. p. B-5.
18. Telecon. Ocamb, D.D., Radian Corporation, with Barbe, B., Exxon Corporation, Baton Rouge, LA. April 13, 1989. Cost for converting top loading railcars to bottom loading.
19. Memorandum from Olsen, T.R., Radian Corporation, to HON project file. August 21, 1991. Loading techniques utilized in the SOCMF.
20. Memorandum from Probert, J.A., Radian Corporation, to HON project file. August 28, 1991. Annual electricity costs for motors that power fans in a vapor collection system for transfer loading racks.
21. Memorandum from Probert, J.A., Radian Corporation, to HON project file. August 28, 1991. Costs for installing a nitrogen blanketing system on a transfer rack.
22. Memorandum from Probert, J.A., Radian Corporation, to HON project file. September 3, 1991. Costs for installing a vapor collection system on a transfer rack.
23. Edwards Engineering Corporation. Equipment Specifications for Solvent Vapor Recovery Units, Low Temperature Packaged Liquid Chillers, and Vapor Condensers. New York. September 1990.

24. U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards. OAQPS Control Cost Manual. Fourth Edition. Supplement 1. Chapter 8. EPA-450/3-90-006a. Research Triangle Park, NC. January 1992.
25. Letter and attachments from Waldrop, R., Edwards Engineering Corp., to Barbour, W., Radian Corporation. September 28, 1990. Response to letter of inquiry regarding cost of refrigerator surface condenser systems.
26. Vataavuk, W.M., and R.B. Neveril. Estimating Costs for Air Pollution Control Systems, Part II: Factors for Estimating Capital and Operating Costs. Chem. Eng. November 3, 1980. pp. 157-162.
27. Memorandum from Probert, J.A., Radian Corporation, to Docket No. A-90-21. September 10, 1990. Costs for condenser removal.
28. U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards. EAB Control Cost Manual Chapter 2: Cost Estimating Methodology. 4th Edition. Draft. Research Triangle Park, NC. March 1989. p. 2-5 to 2-8.
29. Corripio, A.B., K.S. Chrien, and L.B. Evans. Estimate Costs of Heat Exchangers and Storage Tanks via Correlations. Chem. Eng. January 25, 1982. p. 145.
30. Peters, M.S., and K.D. Timmerhaus. Plant Design and Economics for Chemical Engineers. Third Ed. New York, McGraw-Hill Book Company. 1980. pp. 768-773.
31. Mulet, A., A.B. Corripio, and L.B. Evans. Estimate Costs of Distillation and Absorption Towers via Correlations. Chem. Eng. December 28, 1981. p. 180.
32. Ref. 30, p. 572, Figure 13-58.
33. Telecon. Gitelman, A., Research Triangle Institute, with Oakes, D., Hoyt Corporation. September 9, 1986. Cost of flame arrestors.
34. Richardson Engineering Services, Inc. Process Plant Construction Estimating Standards, Volume 3. Section 15-40. Mesa, AZ. 1988.
35. Memorandum from Peterson, P., Research Triangle Institute, to S. Thorneloe, EPA/CPB. January 18, 1988. Basis for steam stripping organic removal efficiency and cost estimates used for source assessment model (SAM) analysis.
36. Ref. 28, p. 4-27.

37. U.S. Environmental Protection Agency. Control of Volatile Organic Compound Emissions from Volatile Organic Liquid Storage in Floating and Fixed Roof Tanks. Draft Report. Office of Air Quality Planning and Standards. Research Triangle Park, NC. June 1984.
38. U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards. VOC Emissions from Volatile Organic Liquid Storage Tanks--Background Information for Promulgated Standards. EPA-450/3-81-003b. Research Triangle Park, NC. January 1987.
39. Ref. 37, p. 5-6.
40. U. S. Environmental Protection Agency. Fugitive Emissions of Organic Compounds--Additional Information on Emissions, Emission Reductions, and Costs. EPA-450/3-82-010. Research Triangle Park, NC. April, 1982. Section 5.
41. U. S. Environmental Protection Agency, Office of Air Quality Planning and Standards. Benzene Fugitive Emissions--Background Information for Promulgated Standards. EPA-450/3-80-032b. Research Triangle Park, NC. June 1982.
42. Ref. 34, Sections 15-42, 15-43, 15-55.
43. National Emission Standards for Hazardous Air Pollutants; Announcement of Negotiated Regulation for Equipment Leaks. Federal Register, Vol. 56, No. 44, pp. 9315-9339. Washington, DC. Office of the Federal Register. March 6, 1991.
44. Ref. 34, Section 15-42.
45. Letter from Caracciolo, D., NAO Inc., to Whitt, D., Radian Corporation. October 31, 1990. Response to request for cost of NAO flame arrestor.
46. Memorandum from Whitt, D., and K. Hausle, Radian Corporation, to Markwordt, D., EPA/CPB. February 21, 1992. Final cost impacts analysis for HON equipment leaks.
47. Ref. 34, Section 15-0.
48. Personal Communication. Nagy, D., BS&B Safety Systems, Inc., Charlotte, NC, with D. J. Whitt, Radian Corporation. October 30, 1990.
49. Personal communication. Moretti, E., Radian Corporation, with Ponder, T., PEI Associates, Dallas, TX. February 2, 1990 and November 22, 1990.

50. Personal communication. Moretti, E., Radian Corporation with Darbonne Services, Inc., Lake Charles, LA. November 27, 1989.
51. Telecon. Whitt, D.J., Radian Corporation, with Johnson, J., Foxboro Equipment Company. October 26, 1990. Pricing of Foxboro equipment.
52. Langley, G.J., S.M. Dennis, L.P. Provost, J.F. Ward. (Radian Corporation). Analysis of SOCMI VOC Fugitive Emissions Data. Prepared for U.S. Environmental Protection Agency. Research Triangle Park, North Carolina. Publication No. EPA-600/2-81-111. June 1981.
53. Memorandum from Moretti, E., Radian Corporation, to Markwordt, D., EPA/CPB. May 22, 1989. Summary of Work Assignment 42 - Equipment Leak Emission Estimates, VOC/HAP.

APPENDIX A

EXAMPLE COSTS FOR INSTALLATION OF A FLARE TO A PROCESS VENT

This is an example calculation to determine the total annual costs associated with controlling an example process vent stream with a flare having a removal efficiency of 98 percent.

The purpose of this appendix is to demonstrate the approach used in the HON analysis. In the calculations below, all significant figures have been retained until the final calculation to make it easier for the reader to follow the calculation and to avoid potential error due to round off of intermediate calculations. It should not be inferred that the intermediate results represent the actual number of significant figures.

The example stream has a volumetric flow rate (Q_{tot}) of 7.39 scf/min and contains ethylbenzene. Other design and costing assumptions are presented in Tables A-1 and A-2. The design and costing procedure follows the one set forth in the Office of Air Quality Planning and Standards Control Cost Manual (OCCM)¹ and described in Section 3.1.1 of the text of this volume.

1. Determine the auxiliary fuel requirement necessary to meet the Federal requirement that a stream have a minimum net heating value (B_v) of 300 Btu/scf, as stated in 40 CFR Section 60.18 of the Federal Register. The example stream has a heating value of 321 Btu/scf, so no auxiliary fuel is required.

2. Calculate the minimum flare tip diameter (D_{min}) based on the maximum stream exit velocity. For a stream with a heat

TABLE A-1. DESIGN ASSUMPTIONS

Parameter Description	Value
Flare type	stream assisted, elevated
Minimum heat content of combusted stream (vent + fuel)	300 Btu/scf
Minimum flare tip diameter	1 in.
Minimum flare height	30 ft.
Maximum flare radiation level	500 Btu/hr-ft ²
Fraction of combustion heat radiated	0.2
Fraction of radiated heat transmitted	1.0
Minimum purge gas velocity	0.04 ft/sec
Steam to waste gas ratio	0.4 lb steam/lb waste gas
Minimum knockout drum diameter	6 in.

TABLE A-2. COST ASSUMPTIONS

Cost Item	Value
<u>Direct Annual Costs</u>	
Dollar basis	July, 1989 (CE Index = 355.9)
Operating labor	
Operator	630 hr/yr
Supervisor	15% of operator
Labor cost	\$13.20/hr
Maintenance	
Labor	1/2 hr/shift
Materials	100% of maintenance labor
Labor cost	\$14.50/hr
Utilities	
Electricity	All utilities equal to
Purge gas	(consumption rate) x (hrs/yr) x
Pilot gas	(unit cost)
Auxiliary fuel	
Steam	
Electricity cost	\$0.0509/kW•hr
Fuel cost	\$3.03/1000 scf
Steam cost	\$7.77/1000 scf
<u>Indirect Annual Cost</u>	
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

^aThe capital recovery factor is based on the assumption of an equipment life of 15 years and an interest rate of 10% with no inflation over the equipment life.

value greater than 300 Btu/scf and less than 1000 Btu/scf, the maximum gas velocity, V_{\max} (ft/sec), is given by:

$$\begin{aligned}\log_{10} (V_{\max}) &= (B_V + 1214)/852 \\ V_{\max} &= 10[(B_V + 1214)/852] \\ &= 10[(321 + 1214)/852] \\ &= 10(1.8) = 63.3 \text{ ft/sec}\end{aligned}$$

The minimum flare tip diameter (D_{\min}) is then given by:

$$D_{\min}(\text{in}) = 1.95 \sqrt{\frac{Q_{\text{tot}} (\text{ft}^3/\text{min})}{V_{\max} (\text{ft/sec})}}$$

Where:

Q_{tot} = vent flow rate + auxiliary fuel flow rate.

$$D_{\min}(\text{in}) = 1.95 \sqrt{\frac{7.39 \text{ ft}^3/\text{min}}{63.33 \text{ ft/sec}}} = 0.67 \text{ in}$$

The minimum flare tip diameter is 1 in, so the rest of the design and costing will be based on a 1-in diameter flare tip.

3. Next, the flare height L , is calculated using:

$$L^2 (\text{ft}^2) = \frac{\eta f R}{4\pi K}$$

where:

η = Fraction of heat intensity transmitted;
 f = Fraction of heat radiated;
 R = Net heat release (Btu/hr);
 K = Allowable radiation (500 Btu/hr·ft²); and
 $R(\text{Btu/hr}) = B_V \times Q_{\text{tot}} \times 60 \text{ min/hr}.$

In order to produce a conservative design, assume that all the heat produced is transmitted ($\eta = 1$) and the fraction of heat radiated, f , is 0.2.

$$R = (7.39 \text{ scfm}) \left(60 \frac{\text{min}}{\text{hr}} \right) \left(321 \frac{\text{Btu}}{\text{scf}} \right) = 142331 \frac{\text{Btu}}{\text{hr}}$$

and

$$L = \left[\frac{(1.0) (0.2) \left(142331 \frac{\text{Btu}}{\text{hr}} \right)}{4 \pi \left(500 \frac{\text{Btu}}{\text{hr} \cdot \text{ft}^2} \right)} \right]^{0.5} = 2.13 \text{ ft.}$$

The minimum design flare height is 30 ft. so costing will be based on a 30 ft. flare.

4. Purge gas is used to keep a minimum required positive flow throughout the system. The purge gas requirement (F_{pu}) is estimated according to flare tip diameter by:

$$F_{pu} \text{ (scf/yr)} = 6.88 \times 10^3 D^2 \text{ (in)}$$

$$F_{pu} \left(\frac{\text{scf}}{\text{yr}} \right) = 6.88 \times 10^3 (1)^2 = 6.88 \times 10^3 \frac{\text{scf purge gas}}{\text{yr}}$$

5. Pilot gas is needed to keep the burners lit. The pilot gas requirement (F_{pi}) is calculated by:

$$F_{pi} \text{ (scf/yr)} = (6.13 \times 10^5) N$$

where:

N = Number of pilot burners (see table below).

<u>Flare Tip Diameter (in)</u>	<u>Number of Pilot Burners (N)</u>
1-10	1
12-24	2
30-60	3
>60	4

The flare tip diameter is 1 in., so the number of pilot burners, N , is 1.

$$F_{pi} \left(\frac{\text{scf}}{\text{yr}} \right) = 6.13 * 10^5 (1) = 6.13 * 10^5 \frac{\text{scf pilot gas}}{\text{yr}}$$

6. A steam-assisted flare uses steam to create turbulence for even mixing of the vent stream in the burner flame. The amount of steam required is based on the mass flow of the vent stream (W). The ideal gas law is used to convert the volumetric vent stream flow to mass flow.

$$W \text{ (lb/hr)} = Q_{\text{vent}} \text{ (scfm)} * 60 \text{ (min/hr)} * P/RT * MW_{\text{avg}}$$

where:

P = Stream pressure (atm);

T = Steam temperature ($^{\circ}\text{R}$);

R = gas constant = $0.7302 \text{ ft}^3 * \text{atm}/\text{lb-mol} * ^{\circ}\text{R}$;

MW_{avg} = average molecular weight of stream (assume $29.0 \text{ lb}/\text{lb-mol}$).

$$W \left(\frac{\text{lb}}{\text{hr}} \right) = \frac{(7.39 \text{ scfm}) \left(60 \frac{\text{min}}{\text{hr}} \right) (1 \text{ atm}) \left(29 \frac{\text{lb}}{\text{lb-mole}} \right)}{\left(0.7302 \frac{\text{atm} \cdot \text{ft}^3}{\text{lb-mol} \cdot ^{\circ}\text{R}} \right) (110.7 + 460)^{\circ}\text{R}}$$

$$W = 30.8 \text{ lb/hr}$$

The steam requirement formula in the OCCM is

$$S \text{ (lb/yr)} = 3500 * W$$

For the example stream,

$$S \text{ (lb/yr)} = 3500 (30.8) = 1.08 * 10^5 \text{ lb/yr}$$

7. A knock-out drum is needed to eliminate mist entrainment in the vent stream. For the HON analysis, the densities of the pollutant vapor and liquid phases are estimated using propane (C_3H_8). Propane was selected because it has a similar molecular structure to that of the design molecule ($\text{C}_{2.85}\text{H}_{5.70}\text{O}_{0.63}$) described in Section 3.1.2.1.1 of the text.^{1,2} The vapor stream density, ρ_v , is $0.111 \text{ lb}/\text{ft}^3$,

and the condensed liquid density, ρ_l , is 37 lb/ft³. The maximum linear velocity of the stream, U , is calculated by:

$$U \text{ (ft/sec)} = G [(\rho_l - \rho_v) / \rho_v]^{0.5}$$

The factor G is assumed to be 0.2 as suggested on page 7-26 of the OCCM. For the example stream,

$$U \text{ (ft/sec)} = 0.2 [(37 - 0.111) / 0.111]^{0.5} = 3.65 \text{ ft/sec}$$

The cross sectional area of the drum, A , is calculated as the volumetric flow rate of the vent stream divided by the maximum linear velocity of the stream.

$$A \text{ (ft}^2\text{)} = Q_{\text{vent}} \text{ (scfm)} / [60 \text{ (sec/min)} * U \text{ (ft/sec)}]$$

$$A = 7.39 \text{ ft}^3/\text{min} / (60 \text{ sec/min} * 3.65 \text{ ft/sec})$$

$$A = 0.034 \text{ ft}^2$$

The minimum drum diameter, t_{\min} , is:

$$t_{\min} \text{ (in)} = 13.5 * A \text{ (ft}^2\text{)}^{0.5}$$

$$t_{\min} = 13.5 (0.034)^{0.5} = 2.5 \text{ in}$$

The minimum available drum size is 6 inches according to the OCCM. For drums of diameter less than 36 in., the drum wall thickness is 1/4 in. The drum height is typically three times the diameter. In this case, the drum height would be 3(6 in) = 18 inches.

8. The three main components of capital cost for flares are the flare itself, the knockout drum, and the piping for the vent stream. The flare cost, C_F , equation varies with flare height and diameter. For flares of height less than 100 feet the equation is:

$$C_F \text{ (March, 1990 \$)} = (78.0 + 9.14D + 0.749L)^2$$

For this example the flare cost is:

$$\begin{aligned} C_F &= [78.0 + 9.14(1.0) + 0.749(30)]^2 (355.9/354.6) \\ &= \$12,058 \end{aligned}$$

when adjusted to July, 1989 dollars.

9. The knockout drum cost, C_K , is dependent upon the height (h), diameter (d), and wall thickness (t) of the drum:

$$C_K \text{ (March, 1990 \$)} = 14.2 [d * t * (h + 0.812d)]^{0.737}$$

The cost of the example drum is:

$$\begin{aligned} C_K &= 14.2 [(6)(0.25)(18 + 0.812(6))]^{0.737} (355.9/354.6) \\ &= \$193 \text{ (July, 1989 dollars)} \end{aligned}$$

10. The piping cost, C_p , depends on the diameter and length of pipe required. The pipe diameter is assumed to equal to the flare tip diameter, and the piping length required is assumed to be 300 feet. The piping cost equation is:

$$C_p \text{ (March, 1990 \$ / 100 ft)} = 1.27D^{1.21}$$

For the example stream,

$$\begin{aligned} C_p &= (300 \text{ ft.}) (1.27) (1.0)^{1.21} (355.9/354.6) \\ &= \$381 \text{ (July, 1989 dollars)} \end{aligned}$$

11. The total equipment cost (EC) is the sum of the flare cost, the knockout drum cost, and the piping cost.

$$\begin{aligned} EC &= C_f + C_k + C_p = \$12,058 + \$193 + \$381 \\ &= \$12,632 \text{ (July, 1989 dollars)} \end{aligned}$$

12. Factors found in the OCCM allow the calculation of purchased equipment cost (PEC) then total capital investment (TCI).

The equations are:

$$PEC = 1.18 EC = 1.18 (\$12,632) = \$14,906$$

and

$$TCI = 1.92 PEC = 1.92 (\$14,906) = \$28,619$$

13. Direct annual costs include labor, materials, and utilities. Labor and materials costs are calculated as a function of hours of labor required and are shown in Table A-3. Utilities costs are calculated from the product of consumption rate and unit cost. The annual consumption values appearing in Table A-3 for the utilities are calculated as follows:

$$\begin{aligned} \text{Annual Gas Consumption} &= \text{Auxiliary Gas} + \text{Purge Gas} + \text{Pilot Gas} \\ &= 0 + (6.88 * 10^3 \text{ scf/yr}) + (6.13 * 10^5 \text{ scf/yr}) \\ &= 619880 \text{ scf/yr} \end{aligned}$$

$$\text{Annual Steam Consumption} = 1.08 * 10^5 \text{ lb/yr} = 108,000 \text{ lb/yr}$$

TABLE A-3. ESTIMATION OF TOTAL ANNUAL COST FOR A FLARE SYSTEM

Cost Component	Cost Factor	Annual Consumption	Annual Cost ^a
<u>Direct Annual Costs</u>			
Utilities			
Gas	\$3.03/1000 scf	619880 scf	\$1,878
Steam	\$7.77/2000 lb	108,000 lb	420
Electricity ^b	\$0.0509/kW-hr	1 kW-hr	0
Labor			
Operating Labor ^c	\$13.20/hr	630 hrs	\$8,316
Supervision & Administration	15% of Operating Labor		\$1,247
Maintenance			
Labor ^d	\$14.50/hr	547.5 hrs	\$7,939
Materials	100% of Maintenance Labor		\$7,939
TOTAL DIRECT ANNUAL COST (TDAC)			\$27,739
<u>Indirect Annual Costs</u>			
Overhead	60% of all labor and materials		\$15,265
General & Administrative			
Property Taxes	1% of TCI		\$286
Insurance	1% of TCI		\$286
Administrative Charges	2% of TCI		\$572
Capital Recovery ^e	13.15% of TCI		\$3,763
TOTAL INDIRECT ANNUAL COST (TIAC)			\$20,172
TOTAL ANNUAL COST (TAC)			TDAC + TIAC
			\$47,911

^aJuly 1989 dollars^bAssumed default value of 1 kW-hr (see Step 13).^cAssumed flare system operated continuously and required 630 hours of labor per year.^dAssumed flare system operated continuously and required 0.5 hour of labor per 8-hour shift.^eSee Table A-2.

$$\text{Annual Electricity Consumption} = (1.17 * 10^{-4}) \\ * (Q_{\text{vent}} \text{ scfm}) (\Delta p) 10.6$$

$$\Delta p \text{ (in H}_2\text{O)} = [(1.238 * 10^{-6}) (Q_{\text{vent}}) - 1.15 * 10^{-4}] \\ * (\text{pipe length})$$

$$\Delta p = [(1.238 * 10^{-6}) (7.39) - 1.15 * 10^{-4}] (300 \text{ ft}) \\ = -0.032 \text{ inches H}_2\text{O}$$

The above pressure drop correlation was developed for use in the HON costing procedure; however, the lower limit of its applicability is a vent stream flow of 93 scfm since a negative pressure drop is not technically realistic. For vent stream flow rates below 93 scfm, the subsequent calculation of power consumption is set to a default value of 1 kW•hr/yr. This default value was used for the example process vent because its vent stream flow rate is only 7.39 scfm.

14. Indirect annual costs include overhead, taxes, insurance, and administration costs. Factors and costs for each of these appear in Table A-3.

15. Total annual cost is the sum of the direct annual cost and the indirect annual cost. The total annual cost for the application of a flare to this example stream is approximately \$47,900.

REFERENCES

1. Blackburn, J.W. Control Device Evaluation Thermal Oxidation. In: Organic Chemical Manufacturing, Volume 4: Combustion Control Devices. Prepared for U.S. Environmental Protection Agency, Research Triangle Park, NC. December 1980. Report 1. pages III-2,5.
2. Memorandum from Scott, K., Radian Corporation, to HON project file. February 5, 1992. Composition of design molecule.

APPENDIX B

EXAMPLE COSTS FOR INSTALLATION OF AN INCINERATOR/SCRUBBER SYSTEM ON A TRANSFER LOADING RACK

The following example illustrates the procedure used to calculate the total annual costs and the total capital costs associated with the addition of an incinerator plus scrubber system to a tank truck or tank car transfer rack. The purpose of this appendix is to demonstrate the approach used in the HON analysis. In the calculations below, all significant figures have been retained until the final calculation to make it easier for the reader to follow the calculation and to avoid potential error due to round off of intermediate calculations. It should not be inferred that the intermediate results represent the actual number of significant figures.

The example tank truck transfer rack is located at an example facility which loads ethylene dichloride, formaldehyde, methanol, and vinyl chloride. The costing procedure for an incinerator and scrubber assigned to a tank car transfer rack is identical to the procedure presented below. Calculation data are provided in Table B-1 for the facility.

B.1 TOTAL CAPITAL COSTS

The total capital cost of the transfer rack control device includes the capital cost of the incinerator and the scrubber. For the purpose of the HON analysis, it was assumed that a nitrogen blanketing system, a vapor collection system, and a submerged loading system were already in place at each facility for each vehicle. Therefore, there were no capital costs associated with these three systems.

TABLE B-1. EXAMPLE FACILITY CALCULATION DATA

	Ethylene Dichloride	Formaldehyde	Methanol	Vinyl Chloride	TOTAL
Maximum Tank Car Throughput (MM gal/yr)	3.01	5.70	19.52	5.06	33.29
Maximum Tank Truck Throughput (MM gal/yr)	1.99	2.50	8.56	2.22	15.26
Actual Tank Car Throughput (MM gal/yr)	3.01	3.88	15.62	4.55	27.06
Actual Tank Truck Throughput (MM gal/yr)	1.99	1.70	6.85	2.00	12.53
Vapor Pressure (psi)	1.51	52.0	1.93	50.29	NA
Molecular Weight (lb/lb-mol)	98.96	30.03	32.04	62.50	NA
Heat of Combustion (Btu/lb)	3400.00	8190.00	8419.00	8136.00	NA
Emission Factor (lb/10 ³ gal)	2.08	0 ^a	0.86	0 ^a	NA
Uncontrolled Emissions from Tank Car Rack (Mg/yr)	2.84	0	6.11	0	8.95
Uncontrolled Emissions from Tank Truck Rack (Mg/yr)	1.874	0	2.680	0	4.55
State Control for Tank Car Rack	None	None	90%	None ^b	NA
State Control for Tank Truck Rack	None	None	None	None ^b	NA
Baseline Emissions from Tank Car Rack (Mg/yr)	2.84	0	0.611	0	3.46
Baseline Emissions from Tank Truck Rack (Mg/yr)	1.87	0	2.68	0	4.55

^aThe emission factor for this chemical is equal to zero, because the chemical has vapor pressure greater than 14.7 psi. It is assumed that the chemical is transferred under pressure, and there are not emissions to the atmosphere.

^bAlthough there is a NESHAP for transfer of vinyl chloride, it did not affect the calculation of baseline emissions for the HON analysis because it was assumed that vinyl chloride would be transferred under pressure without emissions to the atmosphere.

B.1.1 Incinerator Capital Investment

The incinerator capital cost includes the cost of the incinerator equipment and any associated ducts and fans. The cost of the incinerator is directly dependent on the total volumetric flow rate through the control device.

1. Calculation of the amount of VOC entering the incinerator. The amount of VOC entering the incinerator is calculated as follows:

$$\begin{aligned} \text{Mols VOC In} & \quad \left(\text{Baseline Emissions } \frac{\text{Mg}}{\text{yr}} \right) * \left(\frac{10^6 \text{ g}}{\text{Mg}} \right) * \\ (\text{lb-mol/hr}) & \quad \left(\frac{1 \text{ lb}}{453.593 \text{ g}} \right) * \left(\frac{1 \text{ yr}}{\text{Operating Hours}} \right) * \\ & \quad \left(\frac{1}{\text{Molecular Weight of VOC } \frac{\text{lb}}{\text{lb-mol}}} \right) \end{aligned}$$

The baseline emissions of VOC are the emissions generated during the vehicle loading operation which are routed to the incinerator. The operating hours represent the number of hours per year that the control device will operate, and the molecular weight is representative of the VOC molecules entering the incinerator.

- 1a. Calculation of baseline emissions.

To calculate the baseline emissions associated with tank car and tank truck transfer racks, the uncontrolled emissions corresponding to each material must first be calculated. This is done for each vehicle by multiplying the actual throughput of the material by the emission factor of the material. If the loading of a material is regulated by State or Federal control, the uncontrolled emissions for each material are reduced by the quantity (1 - the fractional control level efficiency) to obtain baseline emissions. If the loading of a material is not controlled at the State or Federal level, the baseline emissions are equal to the uncontrolled emissions.

The following equations illustrate the calculation of the tank truck baseline emissions for the methanol produced at the example facility.

The tank truck emission factor for methanol is calculated from the following equation:¹

$$\text{Emission Factor (lb/1000 gal)} = 12.46 * S * VP * MW / (T + 460)$$

where:

S = 0.6 = A saturation factor for submerged loading;

VP = 1.93 = Vapor pressure of methanol (psi);

MW = 32.04 = Molecular weight of methanol (lb/lb-mol);

T = 77 = Loading temperature (°F)

$$\text{Emission Factor} = 12.46 * 0.6 * 1.93 \text{ psi} * 32.04 \frac{\text{lb}}{\text{lb-mol}} \frac{1}{537} = 0.861 \frac{\text{lb}}{1000 \text{ gal}}$$

Therefore:

$$\begin{aligned} \text{Uncontrolled Emissions (Mg/yr)} &= \left(\text{Actual Throughput} \frac{\text{gal}}{\text{yr}} \right) * \left(\text{Emission Factor} \frac{\text{lb}}{1000 \text{ gal}} \right) \\ &\quad * \left(453.593 \frac{\text{g}}{\text{lb}} \right) * \left(\frac{\text{Mg}}{10^6 \text{ g}} \right) \end{aligned}$$

$$\text{Tank Truck Uncontrolled Emissions} = \left(6.85 * 10^6 \frac{\text{gal}}{\text{yr}} \right) * \left[\frac{0.861 \text{ lb}}{10^3 \text{ gal}} \right] * \left(453.593 \frac{\text{g}}{\text{lb}} \right) * \left[\frac{\text{Mg}}{10^6 \text{ g}} \right] = 2.68 \frac{\text{Mg}}{\text{yr}}$$

Assume the example facility is located in a State which requires 90 percent control of emissions from materials having a vapor pressure greater than or equal to 1.5 psi and an actual throughput to the tank truck rack greater than 40,000 gal/day. The throughput of a chemical in gallons per day is calculated by dividing the annual throughput by the number of days per year that a chemical is transferred. It is assumed that facilities transfer a chemical for the same number of days per year that they produce the chemical. Multiplying the number of days in a year by the capacity

utilization factor results in an estimate of the number of days in a year that the facility transfers product. The capacity utilization factor represents the difference between the amount of product actually produced and the maximum capacity. Of the four chemicals produced and transferred at the facility, ethylene dichloride and methanol meet the vapor pressure criteria for State control. However, none of the chemicals meet the tank truck throughput requirement for State control. Therefore, the baseline emissions for methanol and ethylene dichloride transfer are equal to the respective uncontrolled emissions because they do not meet the requirements for State control. The emission factors for vinyl chloride and formaldehyde are zero because these materials have a vapor pressure greater than 14.7 psi. Although there is a NESHAP for vinyl chloride, it is assumed that vinyl chloride and formaldehyde are transferred under pressure and their uncontrolled and baseline emissions are equal to zero. Therefore, there are no control costs associated with a chemical having a vapor pressure greater than atmospheric pressure. The data and results are shown in Table B-1.

To obtain the total baseline emissions for the tank truck transfer rack, the baseline emissions from tank truck loading operations for each material are summed. For the example facility, the tank truck rack total baseline emissions are 4.55 Mg/yr.

1b. Calculation of operating hours.

Operating hours for the control device are calculated by doubling the loading time. This allows for warm-up of the incinerator prior to loading and cool-down after loading is completed. Loading time is calculated by dividing the total annual rack actual throughput by the total filling rate. The total rack actual throughput is calculated by summing the actual rack throughput of every transferred chemical. The total filling rate is equal to the vehicle filling rate multiplied by the number of arms per rack. In the HON analysis, it is assumed that the vehicle filling rates of tank

trucks and tank cars are 170 gpm and 190 gpm, respectively. The number of arms per transfer rack are determined by the maximum total annual rack throughput and the total number of materials transferred as shown in Tables B-2 and B-3.

For the example facility, the total annual maximum throughput to the tank truck transfer rack is 15.26 MMgal/yr. Since this facility transfers four materials, it is assigned one 4-arm transfer rack. Therefore, the operating hours for the example facility are calculated from the total annual actual throughput as follows.

$$\text{Operating Hours (hr/yr)} = \left(\frac{12.53 \times 10^6 \frac{\text{gal}}{\text{yr}}}{170 \frac{\text{gal}}{\text{min} \cdot \text{arm}} * \frac{60 \text{ min}}{\text{hr}} * 4 \text{ arms}} \right) * 2 = 614.2 \frac{\text{hr}}{\text{yr}}$$

1c. Calculation of molecular weight.

The molecular weight of the stream sent to the incinerator is a weighted average of the molecular weights of the individual chemicals in the stream. To obtain a rack molecular weight, the molecular weight of each transferred chemical is weighted based on the quantity of the chemical sent to the incinerator. The quantity of the chemical routed to the incinerator is equal to the baseline emissions of the chemical. The following equations are used to calculate the number of lb-moles of ethylene dichloride and methanol entering the incinerator. Formaldehyde and vinyl chloride are not included in the molecular weight calculation, because it is assumed that the two chemicals have no baseline emissions (see Step 1a).

$$\begin{array}{l} \text{Ethylene} \\ \text{Dichloride} \end{array} \left(\frac{\text{lb-mol}}{\text{yr}} \right) = \frac{(1.874 \times 10^6 \text{ g/yr})}{\left(454 \frac{\text{g}}{\text{lb}} \right) * \left(98.96 \frac{\text{lb}}{\text{lb-mol}} \right)} = 41.71 \frac{\text{lb-mol}}{\text{yr}}$$

TABLE B-2. MODEL TANK CAR TRANSFER RACKS

Number of Materials	Throughput (TP) Range (MMgal/yr)	Number of Arms
1 - 3	$0 < TP \leq 10$	3
	$10 < TP \leq 40$	8
	$40 < TP \leq 80^a$	16
4 - 9	$0 < TP \leq 10$	3
	$10 < TP \leq 20$	6
	$20 < TP \leq 30$	10
	$30 < TP \leq 60^b$	16
10 - 22	$0 < TP \leq 3$	3
	$3 < TP \leq 80^c$	10
≥ 23	$0 < TP \leq 10$	4
	$10 < TP \leq 20^d$	9

^aFor throughputs above the maximum value, add an additional 3-arm rack per $10 * 10^6$ gal.

^bFor throughputs above the maximum value, add an additional 3-arm rack per $10 * 10^6$ gal.

^cFor throughputs above the maximum value, add an additional 3-arm rack per $3 * 10^6$ gal.

^dFor throughputs above the maximum value, add an additional 4-arm rack per $10 * 10^6$ gal.

TABLE B-3. MODEL TANK TRUCK TRANSFER RACKS

Number of Materials	Throughput (TP) Range (10^6 gal/yr)	Number of Arms
1 - 4	$0 < TP \leq 3$	1
	$3 < TP \leq 12$	2
	$12 < TP \leq 70^a$	4
5 - 12	$0 < TP \leq 3.5$	1
	$3.5 < TP \leq 7.5$	2
	$7.5 < TP \leq 21$	4
	$21 < TP \leq 54^b$	6
13 - 20	$0 < TP \leq 30^c$	1
≥ 21	$0 < TP \leq 12$	3
	$12 < TP \leq 24^d$	4

^aFor throughputs above the maximum value, add an additional 1-arm rack per $3 * 10^6$ gal.

^bFor throughputs above the maximum value, add an additional 1-arm rack per $3.5 * 10^6$ gal.

^cFor throughputs above the maximum value, add an additional 1-arm rack per $15 * 10^6$ gal.

^dFor throughputs above the maximum value, add an additional 4-arm rack per $12 * 10^6$ gal.

$$\text{Methanol} \left(\frac{\text{lb-mol}}{\text{yr}} \right) = \frac{(2.680 \times 10^6 \text{ g/yr})}{\left(454 \frac{\text{g}}{\text{lb}} \right) * \left(32.04 \frac{\text{lb}}{\text{lb-mol}} \right)} = 184.24 \frac{\text{lb-mol}}{\text{yr}}$$

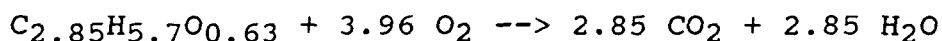
$$\begin{aligned} \text{Tank Truck Rack Molecular Weight} \left(\frac{\text{lb}}{\text{lb-mol}} \right) &= \left(\frac{41.71 \frac{\text{lb-mol}}{\text{yr}}}{41.71 + 184.24 \frac{\text{lb-mol}}{\text{yr}}} \right) * 98.96 \frac{\text{lb}}{\text{lb-mol}} \\ &+ \left(\frac{184.24 \frac{\text{lb-mol}}{\text{yr}}}{41.71 + 184.24 \frac{\text{lb-mol}}{\text{yr}}} \right) * 32.04 \frac{\text{lb}}{\text{lb-mol}} \\ &= 44.39 \frac{\text{lb}}{\text{lb-mol}} \end{aligned}$$

Therefore, to calculate the amount of VOC entering the incinerator for the tank truck transfer rack at the example facility:

$$\begin{aligned} \text{Mols VOC In} &= \left(4.55 \frac{\text{Mg}}{\text{yr}} \right) * \left(\frac{10^6 \text{ g}}{\text{Mg}} \right) * \left(\frac{1 \text{ lb}}{453.593 \text{ g}} \right) * \\ &\left(\frac{1 \text{ yr}}{614.2 \text{ hr}} \right) * \left(\frac{1}{44.39 \frac{\text{lb}}{\text{lb-mol}}} \right) \\ &= 0.368 \frac{\text{lb-mol}}{\text{hr}} \end{aligned}$$

2. Calculate the air required to combust the VOC molecules.

The combustion air requirements are based on a model VOC molecule, $C_{2.85}H_{5.7}O_{0.63}$ as explained in Section 3.1.2.1.1 of the text. Therefore, the combustion mole balance reads as follows:



There are 3.96 moles of oxygen required to combust 1 mole of VOC. Assuming air consists of 20.9 mole percent oxygen, the combustion air requirements for the example facility are:

$$\begin{aligned} \text{Combustion Air} &= \left(0.368 \frac{\text{lb-mol}}{\text{hr}} \text{ VOC} \right) * \left(3.96 \frac{\text{lb-mol } O_2}{\text{lb-mol VOC}} \right) * \left(\frac{1 \text{ lb-mol Air}}{0.209 \text{ lb-mol } O_2} \right) \\ &= 6.97 \frac{\text{lb-mol Air}}{\text{hr}} \end{aligned}$$

3. Calculation of dilution air.

The incinerator exhaust is required to contain at least 3 mole percent oxygen. Since all of the oxygen in the combustion air is reacted in the incinerator, a certain amount of dilution air must be fed to the incinerator. Using a total balance around the incinerator, it is found that the volumetric flow of exhaust is equal to the sum of the volumetric flow of dilution air, combustion air, and VOC entering the incinerator. If it is assumed that all of the combustion oxygen is reacted in the incinerator, and that volume percent is approximately equal to mole percent at atmospheric pressure, then the volumetric flow rate of the dilution air is calculated as follows:

$$0.209 * \left(\text{Flow Rate of Dilution Air} \right) = 0.03 \left(\text{Flow Rate of Exhaust} \right)$$

$$\text{Exhaust (scfm)} = \text{VOC (scfm)} + \text{Combustion Air (scfm)} + \text{Dilution Air (scfm)}$$

For the tank truck rack at the example facility,

$$\begin{aligned} \text{Combustion Air (scfm)} &= 6.97 \frac{\text{lb-mol}}{\text{hr}} * \frac{1 \text{ hr}}{60 \text{ min}} * \frac{392 \text{ scf}}{\text{lb-mol}} \\ &= 45.5 \text{ scfm} \end{aligned}$$

$$\begin{aligned} \text{VOC Stream (scfm)} &= \text{Total filling rate} \\ &= \left(170 \frac{\text{gal}}{\text{min.arm}} \right) * (4 \text{ arms}) * \left(\frac{1 \text{ ft}^3}{7.48 \text{ gal}} \right) \\ &= 90.91 \text{ scfm} \end{aligned}$$

$$0.209 * \left(\frac{\text{Dilution}}{\text{Air}} \right) = 0.03 * \left(\frac{90.91 \text{ scfm} + 45.5 \text{ scfm}}{\text{+ Dilution Air}} \right)$$

$$\text{Dilution Air} = 22.87 \text{ scfm}$$

4. Calculate the auxiliary fuel requirement.

Auxiliary fuel is required for startup of the incinerator unit and is also required to maintain the reactor temperature and to stabilize the flame. Auxiliary fuel is also needed to sustain combustion if the heat content of the VOC stream is low. For the HON analysis, the energy provided by the auxiliary fuel must be at least five percent of the total energy input to the incinerator.

4a. Calculate the volumetric flow rate of the auxiliary fuel.

The volumetric flow rate of auxiliary fuel is calculated using the following equation which appears in the OAQPS Control Cost Manual.²

$$\text{Qaf (scfm)} = \frac{\rho_{fi} Q_{fi} [C_{p\text{air}} (1.1 T_{fi} - T_{fo} - .1 T_{ref}) - (-\Delta h_{cfo})]}{\rho_{af} [(-\Delta h_{caf}) - 1.1 C_{p\text{air}} (T_{fi} - T_{ref})]}$$

where:

ρ_{fi}	=	density of the inlet gas (lb/ft ³)
Q_{fi}	=	flow rate of the inlet gas (ft ³ /min)
C_{pair}	=	specific heat of air (Btu/lb °F)
T_{fi}	=	operating temperature of the incinerator (°F)
T_{fo}	=	temperature of the outlet gas (°F)
T_{ref}	=	reference temperature (°F)
Δh_{cfo}	=	heat of combustion of outlet gas (Btu/lb)
Δh_{caf}	=	heat of combustion of auxiliary fuel (Btu/lb)
ρ_{af}	=	density of the auxiliary fuel (lb/ft ³)

Because the inlet and outlet gas streams consist mainly of air, it is assumed that the density of these streams is equal to the density of air. The operating temperature of the incinerator is 1600 °F if the inlet stream is nonhalogenated and 2000 °F if the inlet stream is halogenated. The reference temperature is 77 °F, and it is assumed that the temperature of the outlet gas stream is also 77 °F. It is also assumed that the heat content of the outlet stream is approximately equal to the heat content of the inlet stream, because these streams consist mainly of air. The auxiliary fuel is methane.

The heat content of VOC stream is calculated and then used to find the total energy input. The heat content of the VOC generated by the loading operation is the sum of the respective heat contents of each transferred chemical divided by the operating hours and the VOC stream flow rate.

$$\text{Heat Content (Btu/scf)} = \frac{\sum \left[\left(\begin{array}{c} \text{Chemical} \\ \text{Baseline} \\ \text{Emissions g/yr} \end{array} \right) * \left(\begin{array}{c} \text{Chemical} \\ \text{Heat} \\ \text{of Combustion} \\ \text{Btu/lb} \end{array} \right) \right]}{\left(\begin{array}{c} \text{VOC Stream} \\ \text{Flow Rate} \\ \text{SCFM} \end{array} \right) * \left(\begin{array}{c} \text{Operating} \\ \text{Hours} \\ \text{hr/yr} \end{array} \right) * \left(60 \frac{\text{min}}{\text{hr}} \right) * \left(453.593 \frac{\text{g}}{\text{lb}} \right)}$$

For the example facility, the VOC stream contains ethylene dichloride and methanol.

$$\begin{aligned} \text{Heat Content (Btu/scf)} &= \frac{[(1.874 * 10^6) * (3400.0) + (2.679 * 10^6) * (8419)]}{(614.2) * (90.91) * (60) * (453.593)} \\ &= 19.03 \frac{\text{Btu}}{\text{scf}} \end{aligned}$$

The heat content of the VOC stream is then adjusted to incorporate the lower heat content of the combustion and dilution air fed to the incinerator with the VOC. The total inlet gas flow rate is equal to the sum of the flow rates of the combustion air, the dilution air, and the VOC stream.

$$\begin{aligned} \text{New Heat Content (Btu/scf)} &= \frac{\text{Heat Content of VOC Stream}}{\frac{\text{Btu}}{\text{scf}}} * \left(\frac{\text{VOC Stream Flow Rate (scfm)}}{\text{Total Inlet Gas Flow Rate (scfm)}} \right) \\ &= 19.03 \frac{\text{Btu}}{\text{scf}} * \left(\frac{90.91 \text{ scfm}}{159.28 \text{ scfm}} \right) = 10.86 \frac{\text{Btu}}{\text{scf}} \end{aligned}$$

The new heat content represents the heat content of the incinerator inlet stream and is therefore approximately equal to the heat content of the outlet stream. If the new heat content is still larger than 98 Btu/scf for nonhalogenated streams or larger than 95 Btu/scf for halogenated streams, more dilution air is required to prevent exceeding design specifications.³ The heat content is then converted to units of Btu/lb.

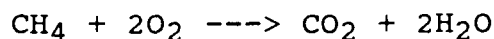
$$\Delta h_{cfo} = \left(10.86 \frac{\text{Btu}}{\text{scf}} \right) * \left(\frac{1 \text{ scf}}{0.0739 \text{ lb}} \right) = 146.96 \frac{\text{Btu}}{\text{lb}}$$

For the example facility, the following input data were used to calculate the flow rate of the auxiliary fuel. Since the example facility transfers a halogenated chemical, ethylene dichloride, the incinerator operates at 2000 °F.

$$\begin{aligned} p_{fi} &= p_{air} = 0.0739 \text{ lb/ft}^3 \\ Q_{fi} &= \text{VOC Stream} + \text{Combustion Air} + \text{Dilution Air} \\ &= 90.91 \text{ scfm} + 45.5 \text{ scfm} + 22.87 \text{ scfm} \\ &= 159.28 \text{ scfm} \\ C_{pair} &= 0.255 \text{ Btu/lb } ^\circ\text{F} \\ T_{fi} &= 2000 ^\circ\text{F} \\ T_{fo} &= 77 ^\circ\text{F} \\ T_{ref} &= 77 ^\circ\text{F} \\ \Delta h_{cfo} &= -146.96 \text{ Btu/lb} \\ \Delta h_{caf} &= -21,502 \text{ Btu/lb} \\ p_{af} &= 0.0408 \text{ lb/ft}^3 \end{aligned}$$

$$\begin{aligned} Q_{af} &= \frac{(0.0739) * (159.28) * [0.255 * (1.1 * (2000) - 77 - (0.1 * 77)) - 146.96]}{(0.0408) * [21,502 - 1.1 * (0.255 * (2000 - 77))]} \\ &= 5.40 \text{ ft}^3/\text{min} \end{aligned}$$

4b. Calculate the air required to combust the auxiliary fuel.



To combust one mole of methane, two moles of oxygen are required. It is assumed that air consists of 20.9 mole percent oxygen.

$$\begin{aligned}
&\text{Air Required} \\
&\text{to Combust} \\
&\text{Methane} \\
&\text{(scfm)} = \left(\frac{2 \text{ mol O}_2}{\text{mol CH}_4} \right) * \left(\frac{1 \text{ mol Air}}{0.209 \text{ mol O}_2} \right) * (5.40 \text{ scfm CH}_4) * \\
&\quad \left(0.0408 \frac{\text{lb}}{\text{scf}} \text{ CH}_4 \right) * \left(\frac{1 \text{ lb-mol}}{16 \text{ lb CH}_4} \right) * \left(\frac{392 \text{ scf}}{1 \text{ lb-mol Air}} \right) \\
&= 51.65 \text{ scfm}
\end{aligned}$$

4c. Calculate the total energy input.

The total energy input is the energy contained in the VOC stream entering the incinerator.

$$\text{TEI} = \rho_{fi} * Q_{fi} * (T_{fi} - T_{ref}) * C_{pfi}$$

The specific heat of the outlet stream, C_{pfi} , is approximately equal to the specific heat of air. The inlet stream flow rate, Q_{fi} , will now include the flow rate of the auxiliary fuel and the flow rate of the air required to combust the auxiliary fuel. The total energy input for the example facility is shown below.

$$\begin{aligned}
\text{TEI} &= \left(0.0739 \frac{\text{lb}}{\text{scf}} \right) * (159.28 \text{ scfm} + 5.40 \text{ scfm} + 51.65 \text{ scfm}) \\
&* (2000^\circ\text{F} - 77^\circ\text{F}) * \left(0.255 \frac{\text{Btu}}{\text{lb}^\circ\text{F}} \right) \\
&= 7839.4 \frac{\text{Btu}}{\text{min}}
\end{aligned}$$

4d. Calculate the auxiliary fuel energy input.

The auxiliary fuel energy input equals the energy contained in the methane entering the incinerator.

$$\text{AFEI} = \rho_{af} * Q_{af} * (-\Delta h_{caf})$$

where:

$$\begin{aligned}
\rho_{af} &= \text{density of methane (lb/ft}^3\text{)} \\
Q_{af} &= \text{volumetric flow rate of methane (scfm)} \\
\Delta h_{caf} &= \text{heat of combustion of methane (Btu/lb)}
\end{aligned}$$

For the example facility,

$$\begin{aligned} \text{AFEI} &= \left(0.0408 \frac{\text{lb}}{\text{ft}^3} \right) * (5.40 \text{ scfm}) * \left(21,502 \frac{\text{Btu}}{\text{lb}} \right) \\ &= 4737.3 \frac{\text{Btu}}{\text{min}} \end{aligned}$$

The energy input of the auxiliary fuel must be greater than 5 percent of the total energy input.

$$\text{TEI} * 0.05 = 7839.4 \frac{\text{Btu}}{\text{min}} * 0.05 = 392 \frac{\text{Btu}}{\text{min}}$$

The auxiliary fuel energy input is much greater than 392 Btu/min.

5. Calculate the capital cost of the incinerator.

The cost of the incinerator is based on the total volumetric flow to the control device. The total inlet volumetric flow is equal to the sum of the flow rates of the VOC stream, the combustion and dilution air, and the auxiliary fuel and associated combustion air. For the example facility, the total volumetric flow to the incinerator is equal to 216.3 scfm. However, in the HON analysis, it is assumed that the smallest incinerator is designed for a flow of 500 scfm. Therefore, the equipment cost of any incinerator with a flow less than 500 scfm will be based on a flow of 500 scfm.

The equipment cost equation is based on zero percent heat recovery, because the stream is halogenated. Cost equations for the incinerator are given in Table 3.1-5 of the text.

$$\text{EC} = 10294 * (\text{Qtot})^{0.2355} * \left(\frac{\text{Price Index}}{\$340.1} \right)$$

The price index will adjust the cost of the incinerator to July, 1989 dollars.

$$EC = 10294 * (500)^{0.2355} * \left(\frac{\$355.9}{\$340.1} \right)$$

$$= 46,549.30$$

It is assumed that the ducting associated with the incinerator is 300 feet long with two elbows per 100 feet. The ducting also has a diameter of 24 inches and is made of 1/8 inch thick carbon steel. The ducting cost is also adjusted to July, 1989 dollars.

$$EC_{(duct)} = [210 * (24)^{0.839} + 2 * 4.52 * (24)^{1.43}] * 3$$

$$* \frac{\$355.9}{\$352.4} = \$11,732.41$$

The cost of the fan is proportional to the volumetric flow rate of the VOC which the fan must send to the incinerator. The cost is adjusted to July, 1989 dollars.

$$EC(fan) = (96.96418 * (VOC \text{ Flowrate SCFM})^{0.547169})$$

$$* \frac{\text{Cost Index}}{\$342.5}$$

$$Ec(fan) = (96.96418 * (90.91 \text{ scfm})^{0.547169})$$

$$* \frac{\$355.9}{\$342.5}$$

$$= \$1188.42$$

The basic equipment cost (BEC) is equal to the cost of the incinerator and the associated fans and ducting.

$$BEC = \$46,549.30 + \$11,732.41 + \$1188.42 = \$59,470.13$$

The purchased equipment cost (PEC) includes the BEC and the cost of instrumentation, sales tax, and freight.

Instrumentation = 0.1 * BEC
 Sales Tax = 0.03 * BEC
 Freight = 0.05 * BEC

PEC = 1.18 * BEC

PEC = 1.18 * 59,470.13 = \$70,174.75

The total capital investment is calculated by multiplying the purchased equipment cost by an installation factor of 1.25.

TCI = 1.25 * PEC

TCI = 1.25 * \$70,174.75 = \$87,718.44

B.2.1 Scrubber Capital Investment

For the HON analysis, the scrubber design and capital investment is based on a packed bed, countercurrent scrubber tower. The scrubber capital investment includes the cost of the scrubber, packing, platform, stack, and any associated ducts and fans. The capital investment is directly related to the size of the scrubber and the pressure drop across the scrubber, which is dependent on the liquid and vapor flow rates through the column. The absorbing liquid is water, and the vapor consists mainly of air. The design and costing procedures closely follow the procedure presented in Chapter 9 of the OCCM (Gas Absorbers)², and the Handbook on Control Technologies for Hazardous Air Pollutants (HAP Manual)⁴.

1. Calculate the flow rate of acid gas to the scrubber.

The amount of vapor flowing to the scrubber is equal to the amount of vapor leaving the incinerator. The vapor stream is predominantly air.

$$\begin{aligned} \text{Mols Vapor} \\ \left(\frac{\text{lb-mol}}{\text{hr}} \right) &= (216.3 \text{ scfm}) * \left(60 \frac{\text{min}}{\text{hr}} \right) * \left(\frac{1 \text{ lb-mol}}{392 \text{ scf (at } T = 77^{\circ}\text{F)}} \right) \\ &= 33.11 \frac{\text{lb-mol}}{\text{hr}} \end{aligned}$$

2. Calculate the liquid flow rate through the scrubber.

It is assumed that the liquid and vapor flow rates through the column are essentially constant. Therefore, the

liquid to vapor flow rate ratio throughout the column is also constant and, for the purpose of the HON analysis, equal to 17 gpm/1000 scfm.⁵ By converting both quantities to lb-moles per hour and multiplying the ratio by the vapor flow rate, the liquid flow rate is found.

$$\frac{17 \text{ gpm}}{1000 \text{ scfm}} = \frac{\left(17 \frac{\text{gal}}{\text{min}}\right) * \left(60 \frac{\text{min}}{\text{hr}}\right) * \left(8.34 \frac{\text{lb}}{\text{gal}}\right) * \left(\frac{1 \text{ lb-mol}}{18 \text{ lb}}\right)}{\left(1000 \frac{\text{ft}^3}{\text{min}}\right) * \left(\frac{1 \text{ lb-mol}}{392 \text{ scf}}\right) * \left(60 \frac{\text{min}}{\text{hr}}\right)}$$

$$= \frac{472.6 \frac{\text{lb-mol}}{\text{hr}} \text{ water}}{153.06 \frac{\text{lb-mol}}{\text{hr}} \text{ vapor}}$$

$$\begin{array}{l} \text{Liquid} \\ \text{Flow Rate} \\ \text{Through} \\ \text{Scrubber} \\ \text{(lb-mol/hr)} \end{array} = \left(\frac{472.6 \frac{\text{lb-mol}}{\text{hr}} \text{ water}}{153.06 \frac{\text{lb-mol}}{\text{hr}} \text{ vapor}} \right) * 33.11 \frac{\text{lb-mol Vapor}}{\text{hr}} = 102.23 \frac{\text{lb-mol}}{\text{hr}}$$

3. Calculate the diameter of the tower.

As explained in Section 3.1.2.1.5 of the text, a correlation for randomly packed towers based on flooding considerations is used to determine the tower diameter. The vapor flow rate per cross-sectional area of the column is first determined by the correlation. By knowing the vapor flow rate, the cross-sectional area of the column can then be found.

$$\begin{aligned}
 \text{ABSCISSA} &= \frac{\text{Liquid Flowrate} \left(\frac{\text{lb}}{\text{hr}} \right)}{\text{Vapor Flowrate} \left(\frac{\text{lb}}{\text{hr}} \right)} * \left(\frac{\text{Density of Vapor } (\rho_V)}{\text{Density of Liquid } (\rho_L)} \right)^{.5} \\
 &= \frac{\left(102.23 \frac{\text{lb-mol}}{\text{hr}} \right) * \left(\frac{18 \text{ lb}}{\text{lb-mol}} \right)}{\left(33.11 \frac{\text{lb-mol}}{\text{hr}} \right) * \left(\frac{1 \text{ lb}}{\text{lb-mol}} \right)} * \left(\frac{0.0739 \frac{\text{lb}}{\text{ft}^3}}{62.2 \frac{\text{lb}}{\text{ft}^3}} \right)^{.5} \\
 &= 0.0661
 \end{aligned}$$

$$\begin{aligned}
 \text{ORDINATE} &= -0.9809237 * (\text{ABSCISSA}) (-0.0065226 * \ln (\text{ABSCISSA})) \\
 &\quad + (\text{ABSCISSA}) (-0.021897) \\
 &= 0.126
 \end{aligned}$$

The vapor flow rate per cross-sectional area is calculated by the following equation. For the purpose of the HON analysis, the column is assumed to operate at 60 percent of flooding.

$$\begin{aligned}
 &\text{Vapor Flowrate} \\
 &\text{Per Cross-Sectional} \\
 &\text{Area} \left(\frac{\text{lb}}{\text{sec} \cdot \text{ft}^2} \right) = \left[\frac{\text{ORDINATE} * \rho_V * \rho_L * g_c}{\left(\frac{a}{\epsilon^3} \right) * \left(\frac{\mu_L}{2.42} \right)^{.2}} \right]^{1/2} * f
 \end{aligned}$$

where:

$$\begin{aligned}
 \rho_V &= \text{density of the vapor} = 0.0739 \text{ lb/ft}^3 \\
 \rho_L &= \text{density of water} = 62.2 \text{ lb/ft}^3 \\
 g_c &= \text{gravitational constant} = 32.2 \text{ ft} \cdot \text{lbm/lbf} \cdot \text{sec}^2 \\
 \mu_L/2.42 &= \text{viscosity of solvent} = 0.85 \text{ cp}
 \end{aligned}$$

a/ϵ^3 = the void space of packing per surface area to volume ratio of packing = 69.1 ft

f = flooding factor = 0.6

$$\begin{array}{l} \text{Vapor} \\ \text{Flowrate} \\ \text{per Cross-} \\ \text{Sectional} \\ \text{Area} \end{array} \left(\frac{\text{lb}}{\text{sec} \cdot \text{ft}^2} \right) = \left(0.126 * 0.0739 * 62.2 * \frac{32.2}{69.1} * \frac{1}{(0.85)^{0.2}} \right)^{1/2} * 0.6$$

$$\begin{array}{l} \text{Vapor Flowrate} \\ \text{per Cross-Sectional} \\ \text{Area} \end{array} = 0.3168 \frac{\text{lb}}{\text{sec} \cdot \text{ft}^2}$$

$$\begin{array}{l} \text{Area of Tower} \\ (\text{ft}^2) \end{array} = \frac{\left(\text{Vapor Flowrate} \frac{\text{lb}}{\text{sec}} \right)}{\text{Vapor Flowrate per Cross-Sectional Area}}$$

$$\begin{aligned} \text{Area of Tower} (\text{ft}^2) &= \frac{\left(33.11 \frac{\text{lb mol}}{\text{hr}} \right) * \left(29 \frac{\text{lb}}{\text{lbmol}} \right) * \left(\frac{1 \text{ hr}}{3600 \text{ sec}} \right)}{0.3168 \frac{\text{lb}}{\text{sec} \cdot \text{ft}^2}} \\ &= 0.842 \text{ ft}^2 \end{aligned}$$

$$A = \pi * (\text{radius})^2 = \pi * \left(\frac{\text{Diameter}}{2} \right)^2$$

$$\text{Diameter} (\text{ft}) = \left(\frac{4}{\pi} * A \right)^{\frac{1}{2}}$$

$$\text{Diameter} = \left(\frac{4}{\pi} * 0.842 \right)^{\frac{1}{2}} = 1.035 \text{ ft}$$

4. Determine the number of transfer units.

The number of transfer units represent the number of theoretical equilibrium stages required to absorb the vapor pollutant. This number is determined from a quantity called the absorption factor which is calculated by dividing the

liquid to vapor ratio by the slope of the equilibrium line. The slope of the equilibrium line is represented by the difference in the vapor mole fractions divided by the difference in the liquid mole fractions from the top to the bottom of the column. For the purpose of the HON analysis, the slope is equal to 0.1.⁷

$$\begin{aligned} \text{Absorption Factor} &= \left(\frac{472.6 \text{ lb-mol/hr water} * \frac{18 \text{ lb}}{\text{lb-mol}}}{153.06 \text{ lb-mol/hr vapor} * \frac{29 \text{ lb}}{\text{lb-mol}}} \right) * \left(\frac{1}{0.1} \right) \\ &= 19.16 \end{aligned}$$

$$\begin{aligned} \text{Number of Transfer Units (NOG)} &= \ln \left[\left(\frac{\text{Hal-Conc}}{0.02 * \text{Hal-Conc}} \right) * (1 - 1/AF) + (1/AF) \right] \\ &\quad * \left(\frac{1}{1 - 1/AF} \right) \\ \text{NOG} &= 4.072 \end{aligned}$$

5. Determine the height of a transfer unit.

The height of a transfer unit is the height of one theoretical equilibrium stage. This number is determined from the vapor and liquid flow rates per cross-sectional area of the tower and from the Schmidt numbers^{8,9} of the vapor and the liquid.

$$\begin{aligned} \frac{\text{Vapor Flowrate}}{\text{per Cross-Sectional Area} \left(\frac{\text{lb}}{\text{ft}^2 \cdot \text{hr}} \right)} &= \left(0.3168 \frac{\text{lb}}{\text{ft}^2 \cdot \text{sec}} \right) * \left(\frac{3600 \text{ sec}}{\text{hr}} \right) \\ &= 1140.48 \frac{\text{lb}}{\text{ft}^2 \cdot \text{hr}} \end{aligned}$$

$$\begin{aligned} \frac{\text{Liquid Flowrate per Cross-Sectional Area} \left(\frac{\text{lb}}{\text{ft}^2 \cdot \text{hr}} \right)}{\left(\frac{\text{lb}}{\text{ft}^2 \cdot \text{hr}} \right)} &= \frac{\left(102.23 \frac{\text{lb-mol}}{\text{hr}} \right) * \left(18 \frac{\text{lb}}{\text{lbmol}} \right)}{0.842 \text{ ft}^2} \\ &= 2185.4 \frac{\text{lb}}{\text{ft}^2 \cdot \text{hr}} \end{aligned}$$

$$\text{Height of a Transfer Unit} = \text{HG} + \left(\frac{1}{\text{AF}} \right) * \text{HL}$$

where:

$$\text{HG} = b * \frac{(\text{Vapor Flowrate per Area})^c}{(\text{Liquid Flowrate per Area})^d} * \left(\frac{\text{Vapor Schmidt Number}}{\text{Number}} \right)^{0.5}$$

$$\text{HL} = Y * \left(\frac{\text{Liquid Flowrate per Area}}{\text{Viscosity of the Liquid}} \right)^s * \left(\frac{\text{Liquid Schmidt Number}}{\text{Number}} \right)^{0.5}$$

The values b, c, d, Y, and s⁹ and the Schmidt numbers are all constants which are dependent on the packing and the liquid and vapor flow rates through the tower. The height (HOG) of a transfer unit is calculated below for the example facility.

$$\text{HG} = \frac{3.82 * \left(1140.48 \frac{\text{lb}}{\text{ft}^2 \cdot \text{hr}} \right)^{0.41} * (0.809)^{0.5}}{\left(2185.4 \frac{\text{lb}}{\text{ft}^2 \cdot \text{hr}} \right)^{0.45}} = 1.93$$

$$\begin{aligned} \text{HL} &= 0.0125 * \left(\frac{2185.4 \frac{\text{lb}}{\text{ft}^2 \cdot \text{hr}}}{0.85 \text{ cp} * 2.42 \frac{\text{lb}}{\text{hr} \cdot \text{ft} \cdot \text{cp}}} \right)^{0.22} * (381)^{0.5} \\ &= 1.13 \end{aligned}$$

$$\text{HOG} = 1.93 + \left[\left(\frac{1}{19.16} \right) * 1.13 \right] = 1.989 \text{ ft}$$

6. Determine the height of the column.

The height of the column is determined from the number of transfer units and the height of a transfer unit.

$$\text{Height (ft)} = (\text{HOG} * \text{NOG}) + 2 + (0.25 * \text{Diameter})$$

$$\begin{aligned}\text{Height (ft)} &= (1.989 * 4.072) + 2 + (0.25 * 1.035) \\ &= 10.36 \text{ ft}\end{aligned}$$

7. Determine the pressure drop across the tower.

The pressure drop is related to the height of the tower, the liquid and vapor flow rates per cross-sectional area, and the liquid and vapor densities.

$$\begin{aligned}\frac{\text{Pressure}}{\text{Drop}} &= \frac{\text{NOG} * \text{HOG} * g * (10^{-8})}{5.2} \\ &\quad * (10) (r * L'' / \rho_L) * \left(\frac{(G'')^2}{\rho_V} \right)\end{aligned}$$

Where:

NOG = Number of Transfer Units

HOG = Height of a Transfer Unit (ft)

ρ_L = Density of the liquid (lb/ft³)

ρ_V = Density of the vapor (lb/ft³)

L'' = Liquid Flow rate per Area (lb/ft²•hr)

G'' = Vapor Flow rate per Area (lb/ft²•hr)

5.2 = Conversion between lb/ft² and inches of H₂O

The values of g (11.13) and r (0.00295) are constants dependent on the tower design parameters.⁹

$$\begin{aligned} \text{Pressure Drop} &= \left(\frac{1.989 * 4.072}{5.2} \right) * 11.13 * (10^{-8}) * \\ &\quad (10) \left(\frac{0.00295 * 2185.4}{62.2} \right) * \\ &\quad \left(\frac{(1140.48)^2}{0.0739} \right) = 3.87 \text{ inches H}_2\text{O} \end{aligned}$$

8. Calculate the weight of the tower.

As explained in Section 3.1.2.2 of the text, the weight of the tower is dependent on the height and diameter of the tower.

$$\begin{aligned} \text{Weight} &= (48 * \text{Diameter (ft)} * \text{Height (ft)}) + 39 * (\text{Diameter})^2 \\ \text{Weight} &= (48 * 1.035 * 10.36) + 39 * (1.035)^2 = 556.5 \text{ lb} \end{aligned}$$

9. Calculate the capital cost of the scrubber.

The capital cost of the tower and associated equipment is calculated from the design parameters. The procedure follows the procedure presented in Section 3.1.2.2 of the text. All costs are reported in July, 1989 dollars.

$$\begin{aligned} \text{Tower Cost} &= \left(1.900604 * \left(\frac{\text{wt (lb)}}{1000 \text{ lb}} \right)^{0.93839} \right) * \$1000 * \frac{\text{Cost Index}}{\$298.2} \\ \text{Tower Cost} &= \left(1.900604 * \left(\frac{556.5 \text{ lb}}{1000 \text{ lb}} \right)^{0.93839} \right) * \$1000 * \frac{\$355.9}{\$298.2} \\ &= \$1308.76 \end{aligned}$$

The volume of packing is needed to determine packing costs.

The cost of the ducting and fans are based on 100 feet of 24-inch diameter ducting that is 1/8 inch thick carbon steel and has two elbows per 100 feet.

$$\begin{aligned}
\text{Volume (ft}^3\text{)} &= \pi * (\text{Radius (ft)})^2 * \text{HOG (ft)} * \text{NOG} \\
&= \pi * \left(\frac{\text{Diameter (ft)}}{2}\right)^2 * \text{HOG (ft)} * \text{NOG} \\
&= \pi * \left(\frac{1.035 \text{ (ft)}}{2}\right)^2 * (1.989) \text{ ft} * (4.072) \\
&= 6.814 \text{ ft}^3
\end{aligned}$$

$$\text{Packing Cost} = \frac{\$9.7}{\text{ft}^3} * \frac{\$392.8}{\$324.4} * \text{Volume} = \$11.75 * 6.814 = \$80.03$$

$$\begin{aligned}
\text{Duct Cost} &= [(210 * (24)^{0.839}) + (2 * 4.52 * (24)^{1.43})] * \\
&\quad \left(\frac{\$355.9}{\$352.4}\right) \\
\text{Duct Cost} &= \$3910.80
\end{aligned}$$

$$\text{Fan Cost} = 79.1239 * (24)^{0.5612} * \left(\frac{\$355.9}{\$342.5}\right) = \$489.27$$

The platform cost is dependent on the diameter of the column.

$$\begin{aligned}
\text{Platform Cost} &= [(10) (0.7884 * \ln (\text{diameter}) + 3.325)] * \\
&\quad \left(\frac{\$355.9}{\$298.2}\right) \\
&= [(10) (0.7884 * \ln (1.035) + 3.325)] * \left(\frac{\$355.9}{\$298.2}\right) \\
&= \$2684.99
\end{aligned}$$

The stack cost is always equal to \$5000.00.¹⁰ The basic equipment cost (BEC) is the sum of the tower, packing, duct, fan, platform, and stack cost.

$$\begin{aligned} \text{BEC} &= \$1308.76 + \$80.03 + \$3910.80 + \$489.27 + \\ &\quad \$2684.99 + \$5000.00 = \$13,473.85 \end{aligned}$$

The purchased equipment cost includes the BEC and instrumentation, freight, and sales tax.

$$\text{Instrumentation} = 0.1 * \text{BEC}$$

$$\text{Freight} = 0.05 * \text{BEC}$$

$$\text{Sales Tax} = 0.03 * \text{BEC}$$

$$\text{PEC} = 1.18 * \text{BEC} = 1.18 * \$13,473.85 = \$15,899.14$$

The total capital cost of the scrubber system incorporates an installation factor of 2.2.

$$\text{TCI} = 2.2 * \text{PEC} = 2.2 * \$15,899.14 = \$34,978.11$$

The total capital investment of the incinerator and scrubber system is the sum of the two capital costs.

$$\begin{aligned} \text{TCI} &= \text{Incinerator TCI} + \text{Scrubber TCI} \\ &= \$87,718.44 + \$34,978.11 = \$122,696.55 \end{aligned}$$

B.2 TOTAL ANNUAL COST

The total annual cost includes the direct and indirect costs associated with the incinerator and scrubber. Direct costs include costs such as labor, utilities, and maintenance, and the annual cost associated with the nitrogen blanketing system. Indirect costs include overhead, taxes, insurance, administrative costs, capital recovery, and the annual costs associated with maintaining the transfer rack and vehicles.

B.2.1 Incinerator Annual Cost

1. Calculate the annual incinerator direct cost.

The direct annual cost associated with the incinerator is the sum of operating and supervising labor costs, maintenance labor and material costs, and utility costs. Labor wages are reflective of July, 1989.

1a. Calculate operating and supervisory labor costs.

It is assumed that the supervisory labor cost is 15% of the operating labor cost.

$$\begin{aligned}\text{Operating} \\ + \text{Supv Labor} \\ \text{Cost} &= (1.15) * \left(\frac{0.5 \text{ hrs}}{8 \text{ hr shift}} \right) * \left(\text{Operating Hrs } \frac{\text{hr}}{\text{yr}} \right) \\ &\quad * \left(\text{Labor Wage } \frac{\$}{\text{hr}} \right) \\ &= (1.15) * \left(\frac{0.5}{8} \right) * \left(614.2 \frac{\text{hr}}{\text{yr}} \right) * \left(\frac{\$13.20}{\text{hr}} \right) \\ &= \$582.72/\text{yr}\end{aligned}$$

1b. Calculate maintenance material and maintenance labor costs.

It is assumed that the cost of maintenance material is equal to the cost of maintenance labor.

$$\begin{aligned}\text{Maintenance} &= \left(\frac{0.5 \text{ hrs}}{8 \text{ hr shift}} \right) * \left(\frac{\text{Operating Hours}}{\text{hr/yr}} \right) * \left(\frac{\text{Labor Wage}}{\$/\text{hr}} \right) \\ &= \left(\frac{0.5}{8} \right) * \left(614.2 \frac{\text{hr}}{\text{yr}} \right) * \left(\frac{\$14.50}{\text{hr}} \right)\end{aligned}$$

$$\text{Maintenance Labor} = \$556.62/\text{yr}$$

$$\text{Maintenance Material} = \$556.62/\text{yr}$$

1c. Calculate utility costs.

The utility costs associated with the incinerator include the costs of natural gas and electricity.

$$\begin{aligned}\text{Natural Gas} \\ \text{Cost} &= \left(\frac{\$3.03}{1000 \text{ SCF}} \right) * \left(\frac{\text{Natural Gas}}{\text{Flowrate SCFM}} \right) * \\ &\quad \left(60 \frac{\text{min}}{\text{hr}} \right) * \left(\text{Operating Hours } \frac{\text{hr}}{\text{yr}} \right) \\ &= \left(\frac{\$3.03}{1000 \text{ SCF}} \right) * (5.35 \text{ SCFM}) * \left(60 \frac{\text{min}}{\text{hr}} \right) * \left(614.2 \frac{\text{hr}}{\text{yr}} \right) \\ &= \$597.39/\text{yr}\end{aligned}$$

The electricity costs are the costs associated with running the incinerator fan.

The pressure drop across an incinerator with zero heat recovery is 4 inches water.² The efficiency of the fan and motor is assumed to be 60 percent.

$$\begin{aligned}\text{Power} \\ (\text{KW}) &= (1.17 * 10^{-4} * \text{Volumetric Flow (SCFM)} * \\ &\quad \text{Pressure Drop/Efficiency}) \\ &= \left(1.17 * 10^{-4} * 216.3 \text{ SCFM} * \frac{4}{0.6}\right) = 0.1687 \text{ KW}\end{aligned}$$

$$\begin{aligned}\text{Electricity} \\ \text{Cost} &= \frac{\$0.0509}{\text{KW} \cdot \text{hr}} * \text{Power (KW)} * \frac{\text{Operating (Hr)}}{\text{Hours}} \left(\frac{\text{Hr}}{\text{yr}}\right) \\ &= \frac{\$0.0509}{\text{KW} \cdot \text{hr}} * 0.1687 \text{ KW} * 614.2 \frac{\text{hr}}{\text{yr}} = \$5.27/\text{yr}\end{aligned}$$

The direct annual cost associated with the incinerator is the sum of the costs presented above.

$$\begin{aligned}\text{Incinerator} \\ \text{Direct Annual} &= \text{Operating/Supv. Labor} + \text{Maintenance and} \\ \text{Cost} &\quad \text{Labor} \\ &\quad \text{Materials + Utilities} \\ &= \$582.72 + \$556.62 + \$556.62 + \$597.39 + \$5.27 \\ &= \$2298.62/\text{yr}\end{aligned}$$

2. Calculate the indirect annual cost for the incinerator.

The indirect annual cost associated with the incinerator includes the overhead costs; taxes, insurance, and administrative costs; and the capital recovery factor.

2a. Calculate the overhead cost.

The overhead cost is a function of the operating and supervisory labor costs and the maintenance labor and material costs.

$$\begin{aligned}\text{Overhead} &= (\text{Operating/Supv. Labor} + \text{Maintenance} \\ &\quad \text{Labor and Materials}) * 0.6 \\ \text{Overhead} &= (\$582.72 + \$556.62 + \$556.62) * 0.6 = \$1017.58/\text{yr}\end{aligned}$$

2b. Calculate the taxes, insurance, and administrative costs and the capital recovery. Capital recovery is based on equipment life of 10 years and an interest rate of 10 percent.

The taxes, insurance, and administrative costs, and the capital recovery costs are all dependent on the total capital investment of the incinerator.

$$\text{Taxes} = 0.01 * \text{TCI}$$

$$\text{Insurance} = 0.01 * \text{TCI}$$

$$\text{Administration} = 0.02 * \text{TCI}$$

$$\text{Capital Recovery Cost} = 0.16275 * \text{TCI}$$

$$\begin{aligned} \text{Total Cost} &= 0.20275 * \text{TCI} = 0.20275 * \$87,718.44 \\ &= \$17,784.91/\text{yr} \end{aligned}$$

The indirect annual cost for the incinerator is the sum of the above costs.

$$\begin{aligned} \text{Incinerator Indirect Annual Cost} &= \text{Overhead} + \text{Taxes} + \text{Insurance} + \\ &\quad \text{Administration} + \text{Capital Recovery} \\ &= \$1017.58 + \$17,784.91 \\ &= \$18,802.49/\text{yr} \end{aligned}$$

The total direct annual cost of the incinerator is the sum of the direct annual cost and the indirect annual cost.

$$\begin{aligned} \text{Incinerator Total Annual Cost} &= \text{Direct Annual Cost} + \text{Indirect Annual Cost} \\ &= \$2298.62 + \$18,802.49 = \$21,101.11/\text{yr} \end{aligned}$$

B.2.2 Scrubber Annual Cost

1. Calculate the direct annual cost.

The direct annual cost associated with the scrubber includes the operating and supervisory labor costs, the maintenance material and labor costs, and the utility costs. It is assumed that the scrubber operates for the same number

of hours per year as the associated incinerator. Labor rates are reported in July, 1989 dollars.

1a. Calculate the operating and supervisory labor.

It is assumed that the supervisory labor cost is 15 percent of the operating labor cost.

$$\frac{\text{Operating/Supervising Labor}}{\text{Labor}} = (1.15) * \left(\frac{0.5 \text{ hrs}}{8 \text{ hr shift}} \right) *$$

$$\frac{\text{Operating Hours (hr/yr)}}{\text{Hours (hr/yr)}} * \frac{\text{Labor Wage (\$/hr)}}{\text{Wage (\$/hr)}}$$

$$= 1.15 * \frac{0.5 \text{ hrs}}{8 \text{ hr}} * 614.2 \frac{\text{hr}}{\text{yr}} * \frac{\$13.20}{\text{hr}}$$

$$= \$582.72/\text{yr}$$

1b. Calculate the maintenance material and maintenance labor cost.

It is assumed that maintenance material is equal to maintenance labor.

$$\text{Maintenance} = \left(\frac{0.5 \text{ hr}}{8 \text{ hr shift}} \right) * \left(\frac{\text{Operating Hours}}{\text{Hours}} \frac{\text{hr}}{\text{yr}} \right) * \left(\frac{\text{Labor Wage}}{\text{Wage}} \frac{\$}{\text{yr}} \right)$$

$$= \left(\frac{0.5 \text{ hr}}{8 \text{ hr}} \right) * \left(614.2 \frac{\text{hr}}{\text{yr}} \right) * \left(\frac{\$14.50}{\text{hr}} \right)$$

$$\text{Maintenance Labor} = \$556.62/\text{yr}$$

$$\text{Maintenance Material} = \$556.62/\text{yr}$$

1c. Calculate utility costs.

The scrubber utility costs include water and electricity costs. The electricity costs depend on the vapor flow and pressure drop through the column.

$$\begin{aligned}
 \text{Water Cost} &= \frac{\text{water (lb/hr)}}{8.34 \text{ (lb/gal)}} * \frac{\$0.22}{1000 \text{ gal}} * 614.2 \frac{\text{hr}}{\text{yr}} \\
 &= \frac{\left(102.23 \frac{\text{lbmol}}{\text{hr}} * 18 \frac{\text{lb}}{\text{lbmol}}\right)}{8.34 \frac{\text{lb}}{\text{gal}}} * \frac{\$0.22}{1000 \text{ gal}} * 614.2 \frac{\text{hr}}{\text{yr}} \\
 &= \$29.81/\text{yr}
 \end{aligned}$$

$$\begin{aligned}
 \text{Electricity Cost} &= 0.0002 * \frac{\text{Vapor Flow Through Scrubber (SCFM)}}{\text{Operating Hours}} * \frac{\text{Pressure Drop (inches H}_2\text{O)}}{\left(\frac{\text{hr}}{\text{yr}}\right)} * \frac{\$0.0509}{\text{KW} \cdot \text{hr}} \\
 &= 0.0002 * 216.3 \text{ SCFM} * 3.87 \text{ inches} * \\
 &\quad 614.2 \frac{\text{hr}}{\text{yr}} * \frac{\$0.0509}{\text{KW} \cdot \text{hr}} \\
 &= \$5.23/\text{yr}
 \end{aligned}$$

The direct annual cost of the scrubber is the sum of the costs presented above.

$$\begin{aligned}
 \text{Scrubber Direct Annual Cost} &= \text{Operating Supv. Labor} + \text{Maintenance Labor and Cost} \\
 &\quad \text{Materials + Utilities} \\
 &= \$582.72 + \$556.62 + \$556.62 + \$29.81 + \$5.23 \\
 &= \$1731.00/\text{yr}
 \end{aligned}$$

2. Calculate the scrubber indirect annual cost.

The scrubber indirect annual costs include the overhead, taxes, insurance, administrative costs, and capital recovery.

2a. Calculate the overhead cost.

The overhead cost is a function of the operating and supervisory labor costs, and the maintenance labor and materials costs.

$$\text{Overhead} = (\text{Operating/Supv. Labor} + \text{Maintenance Labor and Materials}) * 0.6$$

$$= *\$582.72 + \$556.62 + \$556.62) * 0.6$$

$$= \$1017.58/\text{yr}$$

2b. Calculate the taxes, insurance, and administrative costs, and the capital recovery costs. These costs are dependent on the total capital cost of the scrubber. Capital recovery is based on equipment life of 10 years and an interest rate of 10 percent.

$$\text{Taxes} = 0.01 * \text{TCI}$$

$$\text{Insurance} = 0.01 * \text{TCI}$$

$$\text{Administration} = 0.02 * \text{TCI}$$

$$\text{Capital Recovery Cost} = 0.16275 * \text{TCI}$$

$$\begin{aligned} \text{Total Cost} &= 0.20275 * \text{TCI} = 0.20275 * \$34,978.11 \\ &= \$7091.81/\text{yr} \end{aligned}$$

The indirect annual cost is the sum of the above costs.

$$\begin{aligned} \text{Scrubber Annual Indirect Cost} &= \text{Overhead} + \text{Taxes} + \text{Insurance} + \\ &\quad \text{Administration} + \text{Capital Recovery} \\ &= \$1017.58 + \$7121.42 = \$8139.00/\text{yr} \end{aligned}$$

The total annual cost of the scrubber is equal to the sum of the direct annual cost and the indirect annual cost.

$$\begin{aligned} \text{Scrubber Total Annual Cost} &= \$1731.00 + \$8139.00 = \$9870.00/\text{yr} \end{aligned}$$

B.2.3 Total Annual Cost

The total annual cost includes the total annual cost of the incinerator and scrubber, the annual costs associated with the nitrogen blanketing system, and the annual costs associated with maintaining the transfer rack and vehicles.

1. Calculate the annual cost associated with the nitrogen blanketing system.

This cost is dependent on the total material throughput to the transfer vehicle.

$$\begin{aligned} \text{Tank Truck} \\ \text{Nitrogen} &= \$274 * \text{Vehicle Annual Throughput (MMgal/yr)} \\ \text{Annual Cost} \\ &= \$274 * 12.53 \frac{\text{MMgal}}{\text{yr}} = \$3433.22/\text{yr} \end{aligned}$$

2. Calculate the annual cost of maintaining the vapor collection system (rack annual cost). The cost (\$200) is adjusted to July, 1989 dollars using CE Construction Labor indices.

$$\begin{aligned} \text{Rack} \\ \text{Annual} &= \frac{\$200 * 268.6}{220} = \$244.18/\text{yr} \\ \text{Cost} \end{aligned}$$

3. Calculate the vehicle annual cost for vapor tightness testing.

The vehicle annual cost is dependent on the vehicle fleet size, which is dependent on the number of loads.¹¹ The number of loads is equal to the total maximum production capacity of all transferred chemicals divided by the vehicle capacity. The tank truck and tank car vehicle capacities are equal to 10,000 gallons and 20,000 gallons, respectively.

$$\begin{aligned} \text{Number} &= \frac{\text{Maximum Capacity (gal/yr)}}{\text{of Loads} \quad \text{Vehicle Capacity (gal)}} \\ \text{Number of} &= \frac{15.26 * 10^6 \frac{\text{gal}}{\text{yr}}}{\text{Tank Truck} \quad 10,000 \text{ gal}} = 1526 \text{ loads} \\ \text{Loads} \end{aligned}$$

If the number of loads is greater than 1463 and less than or equal to 1646, then the number of vehicles is equal to nine.

The cost (\$300) is also adjusted to July, 1989 dollars using CE Construction Labor indices.

$$\begin{aligned}\text{Vehicle Annual Cost} &= \frac{300 * 268.6}{220} * \text{Vehicle Fleet Size} \\ \text{Tank Truck Vehicle Annual Cost} &= \frac{300 * 268.6}{220} * 9 = \$3296.45/\text{yr}\end{aligned}$$

The total annual cost is the sum of the incinerator annual cost, the scrubber annual cost, the nitrogen annual cost, and the annual cost of the transfer rack and vehicles.

$$\begin{aligned}\text{Total Annual Cost} &= \$21,101.11 + \$9870.00 + \$3433.22 +\end{aligned}$$

$$\begin{aligned}&\$244.18 + \$3296.45 \\ &= \$37,944.96/\text{yr}\end{aligned}$$

Thus, the total annual cost of controlling the tank truck rack at the example facility is approximately \$37,900/yr.

REFERENCES

1. Compilation of Air Pollutant Emission Factors, Volume 1: Stationary Point and Areas Sources. U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards. Research Triangle Park, NC. Publication No. EPA/AP-42. September 1985. pp. 4.4-1 through 4.4-17.
2. U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards. OAQPS Control Cost Manual. Fourth Edition. Section 3.9. EPA-450/3-90-006. Research Triangle Park, NC. January 1990.
3. Memorandum from Pandullo, R.F., Radian Corporation, to Barbour, W., and D. Stone, Radian Corporation; L. Evans, and B. Rosensteel, EPA/CPB; and B. Vataavuk, EPA/SDB. April 27, 1990. Summary of April 11 meeting to discuss thermal incinerator cost issues.
4. U.S. Environmental Protection Agency, Air and Energy Engineering Research Laboratory. Handbook--Control Technologies for Hazardous Air Pollutants. EPA-625/6-86-014. Research Triangle Park, North Carolina. September 1986.
5. Memorandum from Ferrero, B., Radian Corporation, to HON project file. February 5, 1992. Estimating liquid to vapor flow rate ratios in scrubber columns.
6. Treybal, R.E. Mass-Transfer Operations, Third Edition. New York, McGraw-Hill Classic Book Company. 1980. p. 196.
7. Memorandum from Ferrero, B., Radian Corporation, to HON project file. February 5, 1992. Development of the slope of the equilibrium line and the absorption factor for acid gas scrubber design.
8. Memorandum from Barbour, W., Radian Corporation, to HON project file. April 20, 1990. Estimating liquid and vapor Schmidt numbers for acid streams.
9. Danielson, John, A. Air Pollution Engineering Manual, Second Edition. Air Pollution Control District County of LA. U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards. Research Triangle Park, NC. May 1973. p. 214-217.
10. Memorandum from Self, P., Radian Corporation, to HON project file. February 5, 1992. Estimating scrubber stack cost.
11. Memorandum from Scott, K., Radian Corporation, to HON Project file. February 3, 1992. Estimating vehicle fleet size for HON transfer.

APPENDIX C

EXAMPLE COSTS FOR INSTALLATION OF A REFRIGERATED CONDENSER TO A FIXED ROOF TANK

This is an example calculation to determine the total annual costs associated with controlling an example storage tank with a refrigerated condenser having a removal efficiency of 95 percent. The purpose of this appendix is to demonstrate the approach used in the HON analysis. In the calculations below, all significant figures have been retained until the final calculation to make it easier for the reader to follow the calculation and to avoid potential error due to round off of intermediate calculations. It should not be inferred that the intermediate results represent the actual number of significant figures.

The example tank has a capacity of 20,000 gallons and stores vinylidene chloride ($\text{CH}_2=\text{CCl}_2$). Design and cost assumptions are presented in Tables C-1 and C-2.

1. Determine the filling rate for the example tank which is dependent upon the tank capacity. Table C-3 presents tank filling rates which were developed from engineering judgement for five tank capacity ranges. The example model tank has a storage capacity of 20,000 gallons; therefore, its filling rate is 500 gallons/minute.

2. Calculate the annual number of hours a tank experiences working losses. Working losses occur during tank filling operations when the incoming fluid displaces vapors in the tank head space. Therefore, working hours are the number of hours spent each year filling a tank.

TABLE C-1. DESIGN AND COST^a ASSUMPTIONS

Parameter Description (Framework Variable Names)	Value
Annual tank breathing hours (b__hours)	8,760 hr/yr
Storage tank operates at standard conditions	Total pressure of 760 mmHg and a temperature of 25 °C
Volume of ideal gas at standard conditions	392 scf/lb-mol of gas
Molecular weight of air	29 lb/lb mol
Condenser HAP removal efficiency (rem__eff)	0.95
Condenser mechanical work efficiency	0.85
Electricity costs (eleccost)	\$0.0509/kW-hr
Water costs (watercst)	\$0.00022/gal
Labor costs (laborcst)	\$13.20/hr-labor
Maintenance labor costs (mtlbrcst)	\$14.50/hr-labor

^aCosts are given in July 1989 dollars.

TABLE C-2. MODEL TANK DESIGN PARAMETERS

Parameter Description	Units	Value
Number of tanks	--	1
Tank volume (tank_size)	gal	20,000
Annual tank throughput (vol_per_tk)	gal/yr	2,472,727
Tank orientation and type	--	vertical, fixed roof
Tank diameter (tank_dia)	feet	15
Tank height	feet	15
Average tank vapor space height	feet	7.5
Adjustment factor for small diameter tanks (C _d)	dimensionless	0.7306
Tank paint factor - white roof and aluminum color shell (F _p)	dimensionless	1.3
Average ambient diurnal temperature change	°F	20
Product factor - organic liquid other than crude oil (K _C)	dimensionless	1.0
Stored product - HAP	--	vinylidene chloride
Product molecular weight (mol-weight)	lb/lb mol	96.94
Product specific gravity (actual)	--	1.21
Product vapor pressure at 25 °C	mm Hg psia	601.17 11.6279
Product Antoine coefficients		
A	dimensionless	6.972
B	dimensionless	1099.4
C	dimensionless	237.2
Product heat capacity (default)	Btu/lb mol - °F	23
Product heat of vaporization (actual)	Btu/lb	129.660
Product average market price	\$/kg	1.57

TABLE C-3. TANK FILLING RATES

Tank Capacity (gallons)	Filling Rate (gallons/minute)
Capacity \leq 10,000	250
10,000 < Capacity \leq 20,000	500
20,000 < Capacity \leq 40,000	750
40,000 < Capacity \leq 200,000	1,000
Capacity > 200,000	2,000

$$\text{Working Hours} = \frac{\text{annual volume throughput of a single tank}}{\text{filling rate}}$$

$$w\text{-hours} = \frac{\text{vol-per-tk (gal/yr)}}{\text{fill rate (gal/min)}} \left(\frac{1 \text{ hr}}{60 \text{ min}} \right)$$

$$w\text{-hours} = \frac{2,472,727 \text{ (gal/yr)}}{500 \text{ (gal/min)}} * 0.0167 \frac{\text{hr}}{\text{min}}$$

$$w_hours = 82.589 \text{ hr/yr}$$

3. Calculate the number of tank turnovers per year.

$$\text{Turnovers} = \frac{\text{annual throughput of a single tank}}{\text{tank capacity}}$$

$$\text{turn} = \frac{\text{vol-per-tk (gal/yr)}}{\text{tank-size (gal)}}$$

$$\text{turn} = \frac{2,472,727 \text{ (gal/yr)}}{20,000 \text{ (gal)}}$$

$$\text{turn} = 123.64 \text{ turnovers/yr}$$

4. Calculate uncontrolled working and breathing losses from the example tank using AP-42 emission equations adjusted to yield results in Mg/yr.

4a. Fixed roof tank working losses:

$$\begin{array}{l} \text{Fixed Roof} \\ \text{Tank} \\ \text{Working} \\ \text{Losses} \end{array} = 1.089 \times 10^{-8} * \begin{array}{l} \text{HAP} \\ \text{Molecular} \\ \text{Weight} \\ \text{(lb/lb-mole)} \end{array} * \begin{array}{l} \text{HAP} \\ \text{Vapor Pressure} \\ \text{at } 25^{\circ}\text{C} \\ \text{(psia)} \end{array} \\ \quad \quad \quad \begin{array}{l} \text{Annual} \\ \text{Volume} \\ * \text{Throughput} \\ \text{for a tank} \\ \text{(gal/yr)} \end{array} * \begin{array}{l} \text{Tank} \\ \text{Turnover} \\ \text{Factor} \end{array} * \begin{array}{l} \text{Product} \\ \text{Factor} \end{array}$$

$$\text{work-u} = 1.089 \times 10^{-8} * \text{mol-weight} * \text{vp-psia}$$

$$* \text{vol-per-tank} * \left(\frac{180 + \text{turn}}{6 * \text{turn}} \right) * 1.0$$

$$\text{work-u} = 1.089 \times 10^{-8} * 96.94 * 11.6279$$

$$* 2,472,727 * \left(\frac{180 + 123.64}{6 * 123.64} \right) * 1.0$$

$$\text{work_u} = 12.424 \text{ Mg/yr}$$

4b. Fixed roof tank breathing losses:

Fixed Roof Tank Breathing Losses	=	1.025 x 10 ⁻⁵	*	HAP Molecular Weight (lb/lb-mole)	*	Partial Pressure Ratio of HAP to Air
				Tank * Diameter (ft)		Average Tank Vapor Space Height (ft)
						Average Ambient Diurnal Temperature Change (°F)
				Tank * Paint Factor		Tank Adjustment Factor for Small Diameter Tanks
						* Product Factor

$$\text{breath-u} = 1.025 \times 10^{-5} * \text{mol-weight}$$

$$* \left(\frac{\text{vp-psia}}{\text{total pressure} - \text{vp-psia}} \right)^{0.68}$$

$$* (\text{tank-dia})^{1.73} * (\text{dh})^{0.51} * (\text{dt})^{0.50}$$

$$* (F_p) * (C_d) * (K_c)$$

$$\text{breath_u} = 1.025 \times 10^{-5} * 96.94$$

$$* \left(\frac{11.6279}{14.7 - 11.6279} \right)^{0.68}$$

$$* (15)^{1.73} * (7.5)^{0.51} * (20)^{0.50}$$

$$* (1.3) * (0.7306) * (1.0)$$

$$\text{breath_u} = 3.158 \text{ Mg/yr}$$

5. Determine the mass flow rate of HAP emissions into the condenser based on working and breathing losses from the tank.

$$\frac{\text{Uncontrolled Working Losses Entering Condenser}}{\text{Working Hours}} = \frac{\text{Uncontrolled Working Loss Emissions}}{\text{Working Hours}}$$

$$\text{work_in} = \frac{\text{work_u (Mg/yr)}}{\text{w_hours (hr/yr)}} \left(\frac{10^6 \text{ g}}{\text{Mg}} \right) \left(\frac{1 \text{ lb}}{453.6 \text{ g}} \right)$$

$$\text{work_in} = \frac{\text{work_u (Mg/yr)} * 2204.6 \text{ (lb/Mg)}}{\text{w_hours (hr/yr)}}$$

$$\text{work_in} = \frac{12.424 \text{ (Mg/yr)} * 2204.6 \text{ (lb/Mg)}}{82.589 \text{ (hr/yr)}}$$

$$\text{work_in} = 331.642 \text{ lb/hr}$$

$$\frac{\text{Uncontrolled breathing emissions entering condenser}}{\text{breathing hours}} = \frac{\text{uncontrolled breathing loss emissions}}{\text{breathing hours}}$$

$$\text{bth_in} = \frac{\text{breath_u (Mg/yr)} * 2204.6 \text{ (lb/Mg)}}{\text{b_hours (hr/yr)}}$$

$$\text{bth_in} = \frac{3.158 \text{ (Mg/yr)} * 2204.6 \text{ (lb/Mg)}}{8760 \text{ (hr/yr)}}$$

$$\text{bth_in} = 0.7948 \text{ lb/hr}$$

6. Calculate the HAP concentration in working losses at the inlet to the refrigerated condenser.

$$\begin{array}{l} \text{HAP concentration} \\ \text{in working} \\ \text{losses at the} \\ \text{condenser inlet} \\ \text{(concen-w)} \end{array} = \frac{\text{number of lb-moles of} \\ \text{HAP in working losses}}{\text{total number of} \\ \text{lb-moles in working} \\ \text{loss gas stream}} \left(\frac{10^6 \text{ ppm}}{1} \right)$$

6a. First, calculate the number of lb-moles of HAP in the working losses:

$$\begin{array}{l} \text{lb-moles of} \\ \text{HAP in working} \\ \text{losses} \end{array} = \frac{\text{work-u (Mg/yr)} * 2204.6 \text{ (lb/Mg)}}{\text{mol-weight (lb/lb-mol)}}$$

6b. Then, calculate the total number of lb-moles in the working losses:

$$\begin{array}{l} \text{Total lb-moles} \\ \text{in working losses} \end{array} = \frac{\text{vol-per-tk (gal/yr)}}{392 \text{ (scf/lb-mol)}} \left(\frac{0.1337 \text{ ft}^3}{\text{gal}} \right)$$

$$\begin{array}{l} \text{Total lb-moles} \\ \text{in working losses} \end{array} = \text{vol-per-tk} * 3.411 \times 10^{-4} \left(\frac{\text{lb mol}}{\text{yr}} \right)$$

6c. Divide equation 6a by equation 6b:

$$\text{concen-w} = \frac{\left(\frac{\text{work-u} * 2204.6}{\text{mol-weight}} \right) * \left(\frac{10^6 \text{ ppm}}{1} \right)}{\text{vol-per-tk} * 3.411 \times 10^{-4}}$$

$$\text{concen-w} = \frac{6.46 \times 10^{12} * (\text{work-u})}{(\text{mol-weight}) (\text{vol-per-tk})} \text{ ppm}$$

$$\text{concen_w} = \frac{6.46 \times 10^{12} * (12.424)}{(96.94) * (2,472,727)} \text{ ppm}$$

$$\text{concen_w} = 334,823 \text{ ppmv}$$

7. Calculate the HAP concentration in the breathing losses at the inlet to the refrigerated condenser.

$$\begin{array}{l} \text{HAP concentration} \\ \text{in breathing} \\ \text{losses at the} \\ \text{condenser inlet} \\ (\text{concen_b}) \end{array} = \frac{\begin{array}{l} \text{number of lb-moles of} \\ \text{HAP in breathing losses} \end{array}}{\begin{array}{l} \text{total number of} \\ \text{lb-moles in breathing} \\ \text{loss gas stream} \end{array}} \left(\frac{10^6 \text{ ppm}}{1} \right)$$

7a. First, calculate the number of lb-moles of HAP in the breathing losses:

$$\begin{array}{l} \text{lb-moles of HAP} \\ \text{in breathing losses} \end{array} = \frac{\text{breath_u (Mg/yr)} * 2204.6 \text{ (lb/Mg)}}{\text{mol_weight (lb/lb-mol)}}$$

7b. Then, calculate the total number of lb-moles in the breathing losses. It was assumed that the HAP liquid phase was in equilibrium with the HAP gaseous phase and Raoult's Law applies. Thus, the ratio of the HAP partial pressure divided by the total pressure is equal to the HAP mole fraction in the gaseous phase. Therefore, total pressure (760 mm Hg) and vap_pr_25C (HAP partial pressure in mmHg) are shown below in units of lb-moles.

$$\begin{array}{l} \text{Total} \\ \text{lb-moles} \\ \text{in} \\ \text{breathing} \\ \text{losses} \end{array} = \frac{\text{breath_u (Mg/yr)} * \text{Total Pressure (lb-mol)}}{\text{mol_weight (lb/lb-mol)} * \text{vap_pr_25C (lb-mol)}} * 2204.6 \left(\frac{\text{lb}}{\text{Mg}} \right)$$

7c. Divide equation 7a by equation 7b:

$$\text{concen_b} = \frac{\left[\frac{\text{breath_u} * 2204.6 \text{ (lb/Mg)}}{\text{mol_weight}} \right]}{\left[\frac{\text{breath_u} * \text{total pressure} * 2204.6 \text{ (lb/Mg)}}{\text{mol_weight} * \text{vap_pr_25C}} \right]} * \left(\frac{10^6 \text{ ppm}}{1} \right)$$

$$\text{concen_b} = 1316 * \text{vap_pr_25C}$$

$$\text{concen_b} = 1316 * (601.17)$$

$$\text{concen_b} = 791,140 \text{ ppm}$$

8. Calculate total volumetric flow rates at standard conditions (760 mmHg and 25°C) for working losses and breathing losses at the refrigerated condenser inlet.

$$\begin{array}{l} \text{Volumetric flow} \\ \text{rate of working} \\ \text{losses} \end{array} = \begin{array}{l} \text{Filling rate of} \\ \text{storage tank} \end{array}$$

$$\text{w_flow} = \text{fillrate (gal/min)} * 0.1337 \text{ (ft}^3\text{/gal)}$$

$$\text{w_flow} = 500 \text{ (gal/min)} * 0.1337 \text{ (ft}^3\text{/gal)}$$

$$\text{w_flow} = 66.85 \text{ ft}^3\text{/min @ 25°C}$$

$$\begin{array}{l} \text{Volumetric flow} \\ \text{rate of breathing} \\ \text{losses at the} \\ \text{condenser inlet} \end{array} = \frac{\begin{array}{l} \text{Uncontrolled HAP emissions} \\ \text{from breathing losses at} \\ \text{the condenser inlet} \end{array}}{\begin{array}{l} \text{HAP} \\ \text{molecular} \\ \text{weight} \end{array} * \begin{array}{l} \text{HAP concentration} \\ \text{in breathing losses} \\ \text{at the condenser inlet} \end{array}}$$

$$\text{b_flow} = \left(\frac{\text{bth_in (lb/hr)}}{\text{mol_weight (lb/lb-mol)}} \right) * \left(\frac{10^6 \text{ (lb-mol total)}}{\text{concen_b (lb-mol HAP)}} \right) * \left(\frac{392 \text{ scf}}{\text{lb-mol}} \right) * \left(\frac{1 \text{ hr}}{60 \text{ min}} \right)$$

$$b_flow = \frac{(6,533,333 \text{ scfm}) (bth-in)}{(mol-weight) (concen-b)}$$

$$b_flow = \frac{(6,533,333) (0.7948)}{(96.94) (791,140)}$$

$$b_flow = 0.06771 \text{ ft}^3/\text{min} @ 25^\circ\text{C}$$

9. Calculate the mass flow rate of carrier gas (e.g., air) which will enter the refrigerated condenser along with the working and breathing losses.

Carrier gas		
mass flow rate	Volumetric flow	Concentration
for working	= rate of working	* of air in
losses at the	losses	working losses
condenser inlet		

$$w_air = w_flow \text{ (scfm)} * \left[\frac{10^6 - concen-w}{10^6} \left(\frac{\text{lb-mol air}}{\text{lb-mol total}} \right) \right]$$

$$* \left(\frac{1 \text{ lb-mol total}}{392 \text{ scf}} \right) * \left(\frac{29 \text{ lb air}}{\text{lb-mol air}} \right) * \left(\frac{60 \text{ min}}{1 \text{ hr}} \right)$$

$$w_air = (4.439 * 10^{-6} \text{ lb/hr}) (w_flow) (10^6 - concen-w)$$

$$w_air = (4.439 * 10^{-6} \text{ lb/hr}) (66.85) (10^6 - 334,823)$$

$$w_air = 197.39 \text{ lb/hr}$$

Carrier gas		
mass flow rate	Volumetric flow	Concentration
for breathing	= rate of	* of air in
losses at the	breathing losses	breathing losses
condenser inlet		

$$b_{\text{air}} = b_{\text{flow}} \text{ (scfm)} * \left[\frac{10^6 - \text{concen}_{\text{b}}}{10^6} \left(\frac{\text{lb-mol air}}{\text{lb-mol total}} \right) \right] \\ * \left(\frac{1 \text{ lb-mol total}}{392 \text{ scf}} \right) * \left(\frac{29 \text{ lb air}}{\text{lb-mol air}} \right) * \left(\frac{60 \text{ min}}{1 \text{ hr}} \right)$$

$$b_{\text{air}} = (4.439 \times 10^{-6} \text{ (lb/hr)}) (b_{\text{flow}}) (10^6 - \text{concen}_{\text{b}})$$

$$b_{\text{air}} = (4.439 \times 10^{-6} \text{ (lb/hr)}) (0.06771) (10^6 - 791,140)$$

$$b_{\text{air}} = 0.06278 \text{ lb/hr}$$

10. Calculate the mass flow rate of HAP working and breathing losses removed by a refrigerated condenser with a removal efficiency of 95%.

$$\begin{array}{l} \text{Working loss} \\ \text{HAP emissions} \\ \text{removed by} \\ \text{the condenser} \end{array} = \begin{array}{l} \text{Working loss} \\ \text{HAP emissions} \\ \text{entering the} \\ \text{condenser} \end{array} * \begin{array}{l} \text{condenser} \\ \text{efficiency} \end{array}$$

$$\text{work}_{\text{rem}} = \text{work}_{\text{in}} \text{ (lb/hr)} * \text{rem}_{\text{eff}}$$

$$\text{work}_{\text{rem}} = 331.642 \text{ (lb/hr)} * 0.95$$

$$\text{work}_{\text{rem}} = 315.060 \text{ (lb/hr)}$$

$$\begin{array}{l} \text{Breathing loss} \\ \text{HAP emissions} \\ \text{removed by} \\ \text{the condenser} \end{array} = \begin{array}{l} \text{Breathing loss} \\ \text{HAP emissions} \\ \text{entering the} \\ \text{condenser} \end{array} * \begin{array}{l} \text{condenser} \\ \text{efficiency} \end{array}$$

$$\text{bth}_{\text{rem}} = \text{bth}_{\text{in}} \text{ (lb/hr)} * \text{rem}_{\text{eff}}$$

$$\text{bth}_{\text{rem}} = 0.7948 \text{ (lb/hr)} * 0.95$$

$$\text{bth}_{\text{rem}} = 0.7551 \text{ lb/hr}$$

11. Calculate the mass flow rate of HAP working and breathing losses at the condenser outlet.

$$\begin{array}{rcl} \text{Working loss} & & \text{Working loss} \\ \text{HAP emissions} & = & \text{HAP emissions} \\ \text{exiting the} & & \text{entering the} \\ \text{condenser} & & \text{condenser} \end{array} - \begin{array}{r} \text{Working loss} \\ \text{HAP emissions} \\ \text{removed by the} \\ \text{condenser} \end{array}$$

$$\text{work_out} = \text{work_in (lb/hr)} - \text{work_rem (lb/hr)}$$

$$\text{work_out} = 331.642(\text{lb/hr}) - 315.060 (\text{lb/hr})$$

$$\text{work_out} = 16.582 \text{ lb/hr}$$

$$\begin{array}{rcl} \text{Breathing loss} & & \text{Breathing loss} \\ \text{HAP emissions} & = & \text{HAP emissions} \\ \text{exiting the} & & \text{entering the} \\ \text{condenser} & & \text{condenser} \end{array} - \begin{array}{r} \text{Breathing loss} \\ \text{HAP emissions} \\ \text{removed by the} \\ \text{condenser} \end{array}$$

$$\text{bth_out} = \text{bth_in (lb/hr)} - \text{bth_rem (lb/hr)}$$

$$\text{bth_out} = 0.7948 (\text{lb/hr}) - 0.7551 (\text{lb/hr})$$

$$\text{bth_out} = 0.0397 \text{ lb/hr}$$

12. Calculate the vapor pressure of the non-condensed working losses exiting the condenser. The HAP vapor pressure will be the partial pressure as determined from Raoult's Law.

[Note: The cost of a refrigerated condenser system is a function of the temperature to which the HAP vapor must be cooled in order to condense. Systems designed to achieve lower temperatures are more expensive because they can require different coolants and materials of construction. Also, different operating temperatures can increase the amount of energy that must be put into the system.

For a given HAP removal efficiency, the condenser outlet temperature is a function of the level of saturation of the stream entering the condenser and, therefore, the vapor pressure of the stream exiting the condenser. Saturated streams such as tank breathing losses condense more easily; lower temperatures are required to condense unsaturated

streams such as tank working losses. As a conservative estimate of condenser outlet temperature (and therefore, system cost), the condenser outlet temperature will be calculated based on the working losses only.]

$$\begin{array}{l} \text{HAP vapor} \\ \text{pressure from} \\ \text{working losses} \\ \text{at the} \\ \text{condenser outlet} \end{array} = \begin{array}{l} \text{total} \\ \text{pressure} \end{array} * \begin{array}{l} \text{HAP mol fraction} \\ \text{in working losses} \\ \text{at the condenser} \\ \text{outlet} \end{array}$$

$$vpout = 760 \text{ mm Hg} * \left(\frac{\text{lb-mol HAP}}{\text{lb-mol air} + \text{lb-mol HAP}} \right)$$

$$vpout = 760 \text{ mmHg} * \frac{\left(\frac{\text{work-out (lb/hr)}}{\text{mol-weight (lb/lb-mol)}} \right)}{\left(\frac{\text{w-air (lb/hr)}}{29 \text{ lb/lb-mol}} \right) + \left(\frac{\text{work-out} \left(\frac{\text{lb}}{\text{hr}} \right)}{\text{mol-weight} \left(\frac{\text{lb}}{\text{lb-mol}} \right)} \right)}$$

$$vpout = 760 \text{ mm Hg} * \frac{\left(\frac{16.582 \text{ lb/hr}}{96.94 \text{ lb/lb-mol}} \right)}{\left(\frac{197.29 \text{ lb/hr}}{29 \text{ lb/lb-mol}} \right) + \left(\frac{16.582 \text{ lb/hr}}{96.94 \text{ lb/lb-mol}} \right)}$$

$$vpout = 18.640 \text{ mm Hg}$$

13. Calculate the temperature at the outlet of the refrigerated condenser (tout_c) using Antoine's equation:

$$\log_{10} (vpout, \text{ mm Hg}) = A - \frac{B}{(tout_c, ^\circ\text{C}) + C}$$

Where: A, B, and C are Antoine coefficients.

It is assumed that the gaseous (noncondensed) HAP emissions exiting the condenser were in equilibrium with the condensed HAP emissions. Thus, the temperature calculated

from Antoine's equation will be the temperature at the condenser outlet and the condensation temperature necessary to yield a 95% removal efficiency.

Condenser outlet temperature = Antoine's equation solved for temperature

$$\log_{10} (vpout) = A - \frac{B}{(tout-c) + C}$$

$$\log_{10} (vpout) - A = - \frac{B}{(tout-c) + C}$$

$$tout-c + C = - \frac{B}{\log_{10} (vpout) - A}$$

$$tout-c = -C - \frac{B}{\log_{10} (vpout) - A}$$

$$tout-c = - (antcof-c) - \left(\frac{antcof-b}{\log_{10} (vpout) - antcof-a} \right)$$

$$tout-c = - (237.20) - \left(\frac{1099.40}{\log_{10} (18.640) - 6.972} \right)$$

$$tout_C = -44.38^{\circ}\text{C}$$

14. Convert the condenser outlet temperature from degrees Celsius to degrees Kelvin. The condenser outlet temperature is assumed to be the same for both working and breathing losses.

Condenser outlet temperature
in degrees Kelvin = Condenser outlet
for working or temperature in + 273.15
breathing losses degrees celcius

$$\begin{array}{l} \text{temp_w_k} \\ \text{or} \\ \text{temp_b_k} \end{array} = \text{tout_c} + 273.15 \text{ (}^\circ\text{K)}$$

$$\begin{array}{l} \text{temp_w_k} \\ \text{or} \\ \text{temp_b_k} \end{array} = (-44.38 + 273.15) \text{ (}^\circ\text{K)}$$

$$\begin{array}{l} \text{temp_w_k} \\ \text{or} \\ \text{temp_b_k} \end{array} = 228.77^\circ\text{K}$$

15. Convert the condenser outlet temperature from
degrees Celsius to degrees Fahrenheit.

Condenser outlet Condenser outlet
temperature in = temperature in
degrees Fahrenheit degrees Celcius

$$\text{tout_f} = (\text{tout_c} * 9/5) + 32$$

$$\text{tout_f} = (-44.38 * 9/5) + 32$$

$$\text{tout_f} = -47.88 \text{ }^\circ\text{F}$$

16. Calculate the contribution to the condenser heat
load made by the HAP sensible heat loss from working and
breathing losses:

$$\begin{array}{ccccccc} \text{HAP sensible} & & \text{Molar flow} & & & & \text{Temperature} \\ \text{heat loss} & & \text{rate of} & & \text{HAP} & & \text{difference} \\ \text{from} & = & \text{HAP working} & * & \text{molar} & * & \text{between the} \\ \text{the condenser} & & \text{losses at} & & \text{heat} & & \text{condenser inlet} \\ \text{inlet} & & & & \text{capacity} & & \text{and outlet} \end{array}$$

$$w_{\text{ht-gas}} = \left(\frac{\text{work-in lb/hr}}{\text{mol-weight lb/lb-mol}} \right) \left(\text{ht-cap} \frac{\text{Btu}}{\text{lb-mol} \cdot ^\circ\text{F}} \right) \\ * (25^\circ\text{C} - t_{\text{out-c}}) * \left(1.8 \frac{^\circ\text{F}}{^\circ\text{C}} \right)$$

$$w_{\text{ht-gas}} = \left(\frac{331.642 \text{ lb/hr}}{96.94 \text{ lb/lb-mol}} \right) \left(23 \frac{\text{Btu}}{\text{lb-mol} \cdot ^\circ\text{F}} \right) \\ * (25^\circ\text{C} - [-44.38^\circ\text{C}]) * \left(1.8 \frac{^\circ\text{F}}{^\circ\text{C}} \right)$$

$$w_{\text{ht-gas}} = 9827 \text{ Btu/hr}$$

$$\begin{array}{ccccccc} \text{HAP sensible} & & \text{Molar flow rate} & & & & \text{Temperature} \\ \text{heat loss from} & = & \text{of HAP breathing} & * & \text{HAP molar} & * & \text{difference} \\ \text{breathing} & & \text{losses at the} & & \text{heat} & & \text{between the} \\ \text{losses} & & \text{condenser inlet} & & \text{capacity} & & \text{condenser inlet} \\ & & & & & & \text{and outlet} \end{array}$$

$$b_{\text{ht-gas}} = \left(\frac{\text{bth-in lb/hr}}{\text{mol-weight lb/lb-mol}} \right) \left(\text{ht-cap} \frac{\text{Btu}}{\text{lb-mol} \cdot ^\circ\text{F}} \right) \\ * (25^\circ\text{C} - t_{\text{out-c}}) \left(1.8 \frac{^\circ\text{F}}{^\circ\text{C}} \right)$$

$$b_{ht-gas} = \left(\frac{0.7948 \text{ lb/hr}}{96.94 \text{ lb/lb-mol}} \right) \left(23 \frac{\text{Btu}}{\text{lb-mol} \cdot ^\circ\text{F}} \right) \\ * (25^\circ\text{C} - [44.38^\circ\text{C}]) * \left(1.8 \frac{^\circ\text{F}}{^\circ\text{C}} \right)$$

$$b_{ht_gas} = 23.55 \text{ Btu/hr}$$

17. Calculate the contribution to the condenser heat load made by the air sensible heat loss from working and breathing losses.

$$\begin{array}{ccccccc} \text{Air sensible} & & \text{Molar flow} & & & & \text{Temperature} \\ \text{heat loss} & & \text{rate of} & & & & \text{difference} \\ \text{from working} & = & \text{air in working} & * & \text{Air molar} & * & \text{between the} \\ \text{losses} & & \text{losses at the} & & \text{heat} & & \text{condenser} \\ & & \text{condenser} & & \text{capacity} & & \text{inlet and} \\ & & \text{inlet} & & & & \text{outlet} \end{array}$$

$$w_{ht-air} = \left(\frac{w_{air} \text{ lb/hr}}{29 \text{ lb/lb-mol}} \right) [6.713 (298.15 - \text{temp-w-k}) \\ + \frac{4.699 * 10^{-4}}{2} (298.15^2 - \text{temp-w-k}^2) \\ + \frac{1.147 * 10^{-6}}{3} (298.15^3 - \text{temp-w-k}^3) \\ - \frac{4.994 * 10^{-10}}{4} (298.15^4 - \text{temp-w-k}^4)] \left(\frac{\text{cal}}{\text{gmol}} \right) \\ * \left(3.9685 * 10^{-3} \frac{\text{Btu}}{\text{cal}} \right) * \left(453.6 \frac{\text{gmol}}{\text{lb-mol}} \right)$$

$$w_{ht-air} = 0.062 * (197.39) *$$

$$[6.713 (298.15 - 228.77)$$

$$+ \frac{4.699 * 10^{-4}}{2} (298.15^2 - 228.77^2)$$

$$+ \frac{1.147 * 10^{-6}}{3} (298.15^3 - 228.77^3)$$

$$- \frac{4.994 * 10^{-10}}{4} (298.15^4 - 228.77^4)]$$

$$w_{ht-air} = 5865 \text{ Btu/hr}$$

Air sensible heat loss from breathing losses	= Molar flow rate of air in breathing losses at the condenser inlet	* Air molar heat capacity	* Temperature difference between the condenser inlet and outlet
--	---	---------------------------------	--

$$b_{ht-air} = \left(\frac{b_{air} \text{ lb/hr}}{29 \text{ lb/lb-mol}} \right) [6.713 (298.15 - \text{temp-b-k})$$

$$+ \frac{4.699 * 10^{-4}}{2} (298.15^2 - \text{temp-b-k}^2)$$

$$+ \frac{1.147 * 10^{-6}}{3} (298.15^3 - \text{temp-w-k}^3)$$

$$- \frac{4.994 * 10^{-10}}{4} (298.15^4 - \text{temp-b-k}^4)] \left(\frac{\text{cal}}{\text{gmol}} \right)$$

$$* \left(3.9685 * 10^{-3} \frac{\text{Btu}}{\text{cal}} \right) * \left(453.6 \frac{\text{gmol}}{\text{lb-mol}} \right)$$

$$\begin{aligned}
b_{ht-air} = & 0.062 * (0.06278) * \\
& [6.713 (298.15 - 228.77) \\
& + \frac{4.699 * 10^{-4}}{2} (298.15^2 - 228.77^2) \\
& + \frac{1.147 * 10^{-6}}{3} (298.15^3 - 228.77^3) \\
& - \frac{4.994 * 10^{-10}}{4} (298.15^4 - 228.77^4)]
\end{aligned}$$

$$b_{ht-air} = 1.865 \text{ Btu/hr}$$

18. Calculate the contribution to the condenser heat load made by the HAP heat of vaporization.

$$\begin{array}{lcl}
\text{HAP heat of} & \text{Mass flow rate} & \\
\text{vaporization} & \text{of condensed} & \\
\text{from working} & \text{HAP working} & * \text{HAP heat of} \\
\text{losses} & \text{losses from} & \text{vaporization} \\
& \text{the condenser} &
\end{array}$$

$$w_{ht-lq} = \left(\text{work-rem} \frac{\text{lb}}{\text{hr}} \right) \left(\text{ht-vap} \frac{\text{Btu}}{\text{lb}} \right)$$

$$w_{ht-lq} = \left(315.060 \frac{\text{lb}}{\text{hr}} \right) \left(129.660 \frac{\text{Btu}}{\text{lb}} \right)$$

$$w_{ht-lq} = 40,851 \frac{\text{Btu}}{\text{hr}}$$

$$\begin{array}{lcl}
\text{HAP heat of} & \text{Mass flow rate} & \\
\text{vaporization} & \text{of condensed HAP} & \\
\text{from breathing} & \text{breathing losses} & * \text{HAP heat of} \\
\text{losses} & \text{from the condenser} & \text{vaporization}
\end{array}$$

$$b_{ht-lq} = \left(b_{th-rem} \frac{lb}{hr} \right) \left(h_{t-vap} \frac{Btu}{lb} \right)$$

$$b_{ht-lq} = \left(0.7551 \frac{lb}{hr} \right) \left(129.660 \frac{Btu}{lb} \right)$$

$$b_{ht-lq} = 97.91 \frac{Btu}{hr}$$

19. Calculate the total heat load that the condenser must remove from the entering working and breathing losses at standard conditions (25°C and 760 mm Hg) to achieve a condenser exit temperature of -44.38°C.

	HAP sensible	Air sensible	
	heat loss from	heat loss from	
Total	the temperature	the temperature	Heat of
heat =	difference	difference	+ vaporization
load	between the	between the	of the HAP
	condenser inlet	condenser inlet	
	and outlet	and outlet	

Total condenser heat load from working losses:

$$w_{ht_tot} = w_{ht_gas} + w_{ht_air} + w_{ht_lq}$$

$$w_{ht_tot} = 9827 \text{ Btu/hr} + 5865 \text{ Btu/hr} + 40,851 \text{ Btu/hr}$$

$$w_{ht_tot} = 56,543 \text{ Btu/hr}$$

Total condenser heat load from breathing losses:

$$b_{ht_tot} = b_{ht_gas} + b_{ht_air} + b_{ht_lq}$$

$$b_{ht_tot} = 23.55 \text{ Btu/hr} + 1.865 \text{ Btu/hr} + 97.91 \text{ Btu/hr}$$

$$b_{ht_tot} = 123.33 \text{ Btu/hr}$$

20. Calculate the refrigeration capacity required from the condenser total heat load. Refrigeration is defined as the quantity of energy to melt one ton of ice at 32°F in 24 hours, or 12,000 Btu/hr-ton.

$$\frac{\text{Refrigeration capacity required for the condenser heat load due to working losses}}{\text{Total condenser heat load due to working losses}} = \frac{12,000 \frac{\text{Btu}}{\text{hr-ton}}}{12,000 \frac{\text{Btu}}{\text{hr-ton}}}$$

$$w\text{-refrig} = \frac{\left(w\text{-ht-tot} \frac{\text{Btu}}{\text{hr}} \right)}{12,000 \frac{\text{Btu}}{\text{hr-ton}}} = \frac{56,543 \frac{\text{Btu}}{\text{hr}}}{12,000 \frac{\text{Btu}}{\text{hr-ton}}}$$

$$w\text{-refrig} = 4.71 \text{ tons}$$

$$b\text{-refrig} = \frac{\left(b\text{-ht-tot} \frac{\text{Btu}}{\text{hr}} \right)}{12,000 \frac{\text{Btu}}{\text{hr-ton}}} = \frac{\left(123.33 \frac{\text{Btu}}{\text{hr}} \right)}{12,000 \frac{\text{Btu}}{\text{hr-ton}}}$$

$$b\text{-refrig} = 0.0103 \text{ tons}$$

21. Calculate the number of hours the condenser processes working losses assuming that the condenser operates 1/2 hour prior and 1/2 hour after each tank filling as well as during tank filling.

$$\text{Condenser run time for working losses} = \frac{\text{Hours per year spent filling a tank}}{\text{One additional hour of condenser operation per tank filling}}$$

$$w_run = \left(w_hours \frac{hr}{yr} \right) + \left(1 \frac{hr}{turnover} \right) \left(turn \frac{turnovers}{yr} \right)$$

$$w_run = \left(82.589 \frac{hr}{yr} \right) + \left(1 \frac{hr}{turnover} \right) \left(123.64 \frac{turnovers}{yr} \right)$$

$$w_run = 206.23 \frac{hr}{yr}$$

NOTE: If $w_run \geq 8760$ hr/yr then assume $w_run = 8760$ hr/yr.

22. Calculate the number of hours the condenser operates to process breathing losses. It is assumed that when the condenser is not processing working losses, it operates continuously to process breathing losses.

$$\begin{array}{l} \text{Condenser} \\ \text{run time} \\ \text{for breathing} \\ \text{losses} \end{array} = \begin{array}{l} \text{Number of hours} \\ \text{tank experiences} \\ \text{breathing} \\ \text{losses} \end{array} - \begin{array}{l} \text{Condenser run} \\ \text{time for} \\ \text{working losses} \end{array}$$

$$b_run = \left(b_hours \frac{hr}{yr} \right) - \left(w_run \frac{hr}{yr} \right)$$

$$b_run = \left(8760 \frac{hr}{yr} \right) - \left(206.23 \frac{hr}{yr} \right)$$

$$b_run = 8553.77 \frac{hr}{yr}$$

23. Calculate the basic equipment costs (BEC) for a refrigerated condenser system based on the maximum operating limit of the system, which occurs during the control of working losses. Table C-5 contains equations for calculating

BEC as a function of the condenser outlet temperature and refrigeration capacity requirement.

For the example storage tank, the designed condenser has an outlet temperature of -47.85°F (-44.36°C); therefore, the basic equipment costs are calculated from cost algorithm #12.

$$w\text{--cap} = \left(8,380 \frac{\$}{\text{ton}}\right) * w\text{--refrig (tons)} + 49,991(\$)$$

$$w\text{--cap} = \left(8,380 \frac{\$}{\text{tons}}\right) * 4.71 \text{ (tons)} + 49,991(\$)$$

$$w_\text{cap} = \$89,461$$

(Note: If w_refrig had been less than the minimum value (1.58 tons) listed in the table, then setting w_refrig equal to that listed minimum value would yield a conservative estimate of w_cap .)

24. Calculate the total capital investment (TCI) from the basic equipment costs (BEC) using the approach summarized in Table C-6.

25. Calculate the annual condenser electricity costs for controlling working and breathing losses. Electricity cost algorithms were developed from vendor data as a function of condenser outlet temperature, refrigeration requirement, condenser hours of operation, and unit electrical costs. These electricity cost algorithms are presented in Table C-7. For the example storage tank, the designed condenser has an outlet temperature of -47.88°F (-44.38°C); therefore, the annual electricity costs are calculated from algorithm #4.

TABLE C-5. REFRIGERATED CONDENSER SYSTEM -
BASIC EQUIPMENT COST ALGORITHMS*

Equation Number	Temperature at Condenser Outlet [°F]	Basic Equipment Cost (BEC) Algorithms [July 1989 \$]	Minimum (maximum) Value for w_refrig in BEC Algorithm [tons]
1	$T \geq 40$	$(1813 * w_refrig + 13521)$	0.85
2	$40 > T \geq 30$	$(2274 * w_refrig + 13830)$	0.63
3	$30 > T \geq 20$	$(2926 * w_refrig + 13776)$	0.71
4	$20 > T \geq 10$	$(3997 * w_refrig + 17465)$	0.44
5	$10 > T \geq 0$	$(5016 * w_refrig + 18034)$	0.32
6	$0 > T \geq -10$	$(6977 * w_refrig + 16789)$	0.21 (<10)
7		$(3600 * w_refrig + 53834)$	10
8	$-10 > T \geq -20$	$(9450 * w_refrig + 16813)$	0.13 (<5)
9		$(4475 * w_refrig + 51083)$	5
10	$-20 > T \geq -30$	$(5417 * w_refrig + 50111)$	2.42
11	$-30 > T \geq -40$	$(6824 * w_refrig + 50103)$	1.92
12	$-40 > T \geq -50$	$(8380 * w_refrig + 49991)$	1.58
13	$-50 > T \geq -60$	$(8940 * w_refrig + 54549)$	1.25
14	$-60 > T \geq -70$	$(11172 * w_refrig + 52141)$	1.33
15	$T < -70$	$(22248 * w_refrig + 58633)$	0.67

*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. All 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.

TABLE C-6. ESTIMATION OF TOTAL CAPITAL INVESTMENT FOR
A REFRIGERATED CONDENSER SYSTEM^{a, b}

Cost Component	Cost Factor	Component Cost ^a
<u>Direct Equipment Costs^b</u>		
Base Equipment Cost (BEC) TABLE 5		\$89,461
Instrumentation	Included in BEC	
Sales Tax and Freight	8% of BEC	\$7,157
PURCHASED EQUIPMENT COST (PEC)		\$96,618
<u>Installation Cost</u>		
Packaged System	15% of PEC	\$14,493
TOTAL INSTALLATION COST		\$14,493
TOTAL CAPITAL INVESTMENT (TCI)		\$111,111

^aJuly 1989 dollars

^bBased on a HAP condensation temperature of -44°F and a refrigeration requirement of 4.71 tons.

TABLE C-7. REFRIGERATED CONDENSER ELECTRICITY COST ALGORITHMS

Equation Number	Temperature at Condenser Outlet [°F]	Annual Electricity Cost [July 1989 \$/Yr]	Minimum Refrig Value for Electricity Cost Algorithm [tons]
1	$T \geq 40$	$1.3 * (\text{refrig}/0.85) * \text{run} * \text{eleccost}$	0.85
2	$40 > T \geq 20$	$2.2 * (\text{refrig}/0.85) * \text{run} * \text{eleccost}$	0.71
3	$20 > T \geq -20$	$4.7 * (\text{refrig}/0.85) * \text{run} * \text{eleccost}$	0.13
4	$-20 > T \geq -50$	$5.0 * (\text{refrig}/0.85) * \text{run} * \text{eleccost}$	1.58
5	$T < -50$	$11.7 * (\text{refrig}/0.85) * \text{run} * \text{eleccost}$	0.67

aTo calculate refrigerated condenser electricity costs for controlling tank working losses:

refrig = w_refrig

run = w_run

bTo calculate refrigerated condenser electricity costs for controlling tank breathing losses:

refrig = b_refrig

run = b_run

Celeccost = \$0.0509/kW-hr

Annual condenser electricity costs for controlling working losses = Refrigeration requirement for working losses * Condenser run time for working losses

Unit
* electricity + Mechanical efficiency
costs of the condenser

$$w_elec = \left(5.0 \frac{\text{kW}}{\text{ton}}\right) * \left(w_refrig \frac{\text{ton}}{\text{ton}}\right) * \left(w_run \frac{\text{hr}}{\text{yr}}\right) * \frac{\left(eleccost \frac{\$}{\text{kW-hr}}\right)}{0.85}$$

$$w_elec = \frac{(5.0 \text{ kW/ton}) (4.71 \text{ (ton)}) (206.22 \text{ hr/yr}) (0.0509 \text{ \$/kW-hr})}{0.85}$$

$$w_elec = \$290.82/\text{yr}$$

NOTE: If w_refrig had been less than the minimum value (1.58 tons) listed in Table C-7, then setting w_refrig equal to the listed minimum value would yield a conservative estimate of w_elec .

Annual condenser electricity costs for controlling breathing losses = Refrigeration requirement for breathing losses * Condenser run time for breathing losses

Unit
* electricity + Mechanical efficiency
costs of the condenser

$$b_elec = \frac{\left(5.0 \frac{\text{kW}}{\text{ton}}\right) * \left(b_refrig \frac{\text{ton}}{\text{ton}}\right) * \left(b_run \frac{\text{hr}}{\text{yr}}\right) * \left(eleccost \frac{\$}{\text{kW-hr}}\right)}{0.85}$$

As calculated in Step 20, $b_{\text{refrig}} = 0.0103$ tons. Because the lower bound of applicability for equation #4 in Table C-7 is 1.58 tons, b_{elec} must be calculated using this minimum value. Since the condenser will actually be operated at a much lower level (0.0103 tons), use of 1.58 tons gives a conservative estimate of the electricity cost for the system.

$$\frac{b_{\text{elec}} = \left(5.0 \frac{\text{kw}}{\text{ton}}\right) (1.58 \text{ (ton)}) \left(8553.78 \frac{\text{hr}}{\text{yr}}\right) \left(\frac{\$0.0509}{\text{kw-hr}}\right)}{0.85}$$

$$b_{\text{elec}} = \$4,047/\text{yr}$$

26. Calculate the condenser's annual consumption of electricity for controlling working and breathing losses.

$$\begin{array}{l} \text{Condenser's annual} \\ \text{consumption of} \\ \text{electricity for} \\ \text{controlling} \\ \text{working losses} \end{array} = \frac{\text{Annual condenser electricity} \\ \text{costs for controlling working losses}}{\text{Unit electricity costs}}$$

$$w_{\text{kw-hr}} = \frac{w_{\text{elec}} \left(\frac{\$}{\text{yr}}\right)}{\text{eleccost} \left(\frac{\$}{\text{kw-hr}}\right)}$$

$$w_{\text{kw-hr}} = \frac{290.82 \left(\frac{\$}{\text{yr}}\right)}{0.0509 \left(\frac{\$}{\text{kw-hr}}\right)}$$

$$w_{\text{kw-hr}} = 5,714 \frac{\text{kw-hr}}{\text{yr}}$$

$$\begin{array}{l} \text{Condenser's annual} \\ \text{consumption of} \\ \text{electricity for} \\ \text{controlling} \\ \text{breathing losses} \end{array} = \frac{\text{Annual condenser electricity} \\ \text{costs for controlling breathing losses}}{\text{Unit electricity costs}}$$

$$b_kW_hr = \frac{b_elec \left(\frac{\$}{yr} \right)}{eleccost \left(\frac{\$}{kW-hr} \right)}$$

$$b_kW_hr = \frac{4047 \left(\frac{\$}{hr} \right)}{0.0509 \left(\frac{\$}{kW-hr} \right)}$$

$$b_kW_hr = 79,509 \frac{kW-hr}{yr}$$

Condenser's total annual consumption of electricity = Sum of electricity consumption for controlling working and breathing losses

$$tot_kW_hr = w_kW_hr + b_kW_hr$$

$$tot_kW_hr = (5,714 + 79,509) \text{ } \$/yr$$

$$tot_kW_hr = 85,223 \text{ } \$/yr$$

27. Calculate emission reduction as the working and breathing losses removed by the refrigerated condenser.

$$emis_red = \left[\frac{\text{work-rem} \left(\frac{lb}{hr} \right) * \text{w-hours} \left(\frac{hrs}{yr} \right) + \text{bth-rem} \left(\frac{lb}{hr} \right) * \text{8760-w-hours} \left(\frac{hrs}{yr} \right)}{2204.6 \frac{lb}{Mg}} \right]$$

$$emis_red = \frac{(315.060) * (82.589) + (0.7551 * [8760 - 82.589])}{2204.6}$$

$$emis_red = 14.775 \text{ } Mg/yr$$

28. Calculate recovery credit as the product of emission reduction and average market price.

$$\text{rec_credit} = \text{emis_red (Mg/yr)} * (1.57 \text{ \$/kg}) * (1000 \text{ kg/Mg})$$
$$\text{rec_credit} = 14.775 * 1.57 * 1000$$
$$\text{rec_credit} = 23,197 \text{ \$/yr}$$

29. Calculate total annual cost (TAC) using the approach summarized in Table C-8. After rounding, the total annual cost for controlling the example tank is approximately \$38,900/yr.

TABLE C-8. ESTIMATION OF TOTAL ANNUAL COST FOR
A REFRIGERATED CONDENSER SYSTEM

Cost Component	Cost Factor	Annual Consumption	Annual Cost ^a
<u>Direct Annual Costs</u>			
Utilities			
Electricity	\$0.0509/kW-hr	85,223 kW-hr	\$4,338
Labor			
Operating Labor ^b	\$13.20/hr	547.5 hrs	\$7,227
Supervision & Administration	15% of Operating Labor		\$1,084
Maintenance			
Labor ^b	\$14.50/hr	547.5 hrs	\$7,939
Materials	100% of Maintenance Labor		\$7,939
TOTAL DIRECT ANNUAL COST (TDAC)			\$28,527
<u>Indirect Annual Costs</u>			
Overhead	60% of all labor and materials		\$14,513
General & Administrative			
Property Taxes	1% of TCI		\$1,112
Insurance	1% of TCI		\$1,112
Administrative Charges	2% of TCI		\$2,224
Capital Recovery	15 yrs @ 10%		\$14,621
TOTAL INDIRECT ANNUAL COST (TIAC)			\$33,582
RECOVERY CREDIT (RC)			\$23,197
TOTAL ANNUAL COST (TAC)	TDAC + TIAC - RC		\$38,912

^aJuly 1989 dollars

^bAssumes refrigerated condenser system operated continuously and required 0.5 hour of labor per 8-hour shift.

APPENDIX D

APPLICATION OF STEAM STRIPPER COSTING METHODOLOGY

D.1 INTRODUCTION

This appendix presents an application of the costing methodology given in Section 3.2.3 of this volume to an example steam stripper system. The example steam stripper system is based on the configuration shown in Figure 2-8 of this volume and is designed to treat an example wastewater stream with a flow of 500 lpm (see Table D-1). Because the cost of a steam stripper does not vary greatly with inlet organic concentration, the wastewater composition is not shown.

D.2 SYSTEM DESIGN

The first step in costing a steam stripper system is to design the system.

D.2.1 Data Collection for Steam Stripper Design

Information on the design and operation of steam stripper systems is available from studies conducted by EPA.¹ During the Industrial Wastewater Project, EPA obtained information on approximately 15 steam strippers from facility responses to CAA Section 114 information requests in 1987.² Information was also gathered during site visits to eight chemical manufacturing facilities operating steam strippers to remove organic compounds from wastewater.^{3,4,5,6,7,8,9,10} The EPA also gathered data on steam stripper operation as a part of the Hazardous Waste TSDF Project. During this project, data were gathered through field testing at three steam strippers.^{11,12,13}

TABLE D-1. DESIGN AND OPERATING BASIS FOR A 500- ℓ pm
STEAM STRIPPING SYSTEM

-
-
1. Wastewater Flow: 500 ℓ /min (132 gal/min)
 2. Stripper Operating Period: 24 hr/day x 300 day/yr =
7,200 hr/yr
 3. Wastewater Storage: Wastewater feed collection tank with
48-hr retention time
 4. Steam Stripping Column:

Configuration: countercurrent flow, 9.0-m sieve
tray column (10 trays)

Steam Flow Rate: 0.096 kg of steam/ ℓ of waste feed
(0.8 lb steam/gallon)

Wastewater Feed Temperature: 35 $^{\circ}$ C (95 $^{\circ}$ F)

Column Diameter: 0.98 m (3.2 ft)

Active Column Height: 6.5 m (21.3 ft)

Total Column Height: 9.0 m (29.5 ft)

Liquid Loading: 39,900 ℓ /hr/m² (979 gal/hr/ft²)
 5. Condenser:

Configuration: Water-cooled

Primary Condenser Outlet Vapor
Temperature: 50 $^{\circ}$ C (122 $^{\circ}$ F)
 6. Overhead Control: Vent to existing on-site combustion
or other control device
 7. Bottoms Control: Feed to existing on-site wastewater
treatment facility or POTW
-
-

The EPA also gathered data on steam strippers during the development of effluent guidelines for the OCPSF, Pesticide, and Pharmaceutical Manufacturing industries. In response to CWA Section 308 information requests, 63 OCPSF facilities reported using a total of 108 steam strippers as in-plant control for process wastewater.¹⁴ Field testing also contributed data on steam strippers operating at three OCPSF facilities.^{15,16,17} Data available for steam strippers in use at pesticide and pharmaceutical manufacturing industry facilities came from eight pesticide facilities and eight pharmaceutical facilities.^{18,19} These strippers are similar to strippers used in the SOCMI.

The steam stripper systems that are considered treat wastewater streams that vary in flow rate and composition, and some streams contain relatively high levels of suspended solids. However, although the wastewater characteristics vary, the basic steam stripper system shown in Figure 2-8 can be designed and operated to achieve high efficiency in removing organic compounds from most streams.

The organic removal performance of 13 steam stripper systems was measured during field tests sponsored by the EPA and provided through CAA Section 114 questionnaire responses and site visits to industrial facilities. Information gathered during these efforts is supplied in Appendix C of the Control Technology Center Document for Industrial Wastewater Volatile Organic Compound Emissions--Background Information for BACT/LAER Determinations. These data are summarized in Table D-2. All significant figures have been retained for intermediate values reported in this table to make it easier for the reader to follow the calculation and to avoid potential error due to round-off. It should not be inferred that the intermediate values represent the actual number of significant figures. The organic removals presented in Table D-2 range from 76 percent for Site 7 to greater than 99.96 percent for Site C.

TABLE D-2. STEAM STRIPPER ORGANIC COMPOUND REMOVAL PERFORMANCE

Site I.D.	Organic Compounds	Column Design		Column Operation			Column Performance		
		Height (m)	Diameter (m)	Feed Rate (kg/hr)	Steam:Feed Ratio (kg/kg)	Feed Conc. (mg/L)	Bottoms Conc. (mg/L)	Organic Compound Removal (%)	
A	CHC ^a , Benzene, Chlorobenzene	NA ^b	NA	48,960	0.03	5,900	9.8	>99.8	
B	CHC	3.0 ^c	0.2	1,260	0.1	3,900	5.2	99.8	
C	Benzene, Chlorobenzene, 1,2-, 1,3-, and 1,4-Dichlorobenzene, Ethylbenzene, Toluene, Xylene	49.0	2.5	6,624	0.14	7.98	0.305	99.96	
D	1,1-Dichloroethane, 1,1-Dichloroethene, trans-1,1-Dichloroethane, 1,1,1-Trichloroethane, Trichloroethane, Trichloroethene, 1,1,2-Trichloroethane, 1,1,2,2-Tetrachloroethane, Tetrachloroethene,	NA	NA	143,556	0.14	1,984	4.9	99.75	
E	Benzene, Toluene, Ethylbenzene, Isophorone, Naphthalene, 5-Ethyl-1,2-methylpyridine, 1,2,3,4-Tetrahydronaphthalene, Acetophenone, 2-Methyl-1,3-Cyclopentanediol	26	NA	1,831	0.70	22.9	0.079	92.25	
F	Chloromethane, Methylene Chloride, Chloroform, Carbon tetrachloride, Trichloroethylene, 1,2-Trichloroethane	NA	1.5	2,440	0.10	5,860	0.033	>99.9	
G	Nitrobenzene, Nitrotoluene	19.2	0.46	29,900	0.07	634	47.8	92	
7	Benzene, Nitrobenzene, Phenol, Nitrophenols	9.8 ^c	0.61	5,452	0.4	1,192	250	76	
10	CHC, Chlorinated ethers, Phenol, Chlorinated phenols	6.1 ^c	1.07	12,693	NA	453	3.4	99	
Z	Benzene, Toluene, Chlorobenzene, Ethylbenzene, Methylene chloride	12.2	1.22	68,100	0.20	2,073	0.04	>99.8	

TABLE D-2. STEAM STRIPPER ORGANIC COMPOUND REMOVAL PERFORMANCE
(CONTINUED)

Site I.D.	Organic Compounds	Column Design		Column Operation		Column Performance		
		Height (m)	Diameter (m)	Feed Rate (kg/hr)	Steam:Feed Ratio (kg/kg)	Feed Conc. (mg/l)	Bottoms Conc. (mg/l)	Organic Compound Removal (%)
151	Benzene, Nitrobenzene, 2,4-Dinitrobenzene, Aniline, Nitrophenol, Phenol	11.6	1.83	8,402. ^d	0.625	2,317. ^e	23. ^f	99
154	Aniline, Nitrobenzene, Benzene	11.9	1.07	3,885. ^g	NA	3,170	31.7 ^f	99
163	Acrolein, Vinyl acetate, Acetaldehyde	7.32	0.61	5,892. ^{h,i}	0.05	1,150. ^j	12	99
611	Methylene chloride, Methanol, Dimethyl ether	8.32	0.51	3,772. ^d	0.23	55,000. ^k	5,500. ^f	>90
921	Ethylene dichloride	13.9	0.91	16,707. ^d	0.113	324. ^j	0.649	99.8
312	Monoester, n-Butanol, Dibutyl phthalate	8.46	0.46	13,728. ^l	0.11	128,000	2,300	99.8
321	Benzene, Aniline, Caprolactam	14.7	0.91	9,945. ^d	0.12	34,300	31,205	99.9
131	n-Butanol, 4-Aminobiphenyl, Aniline	22.0	0.51	737. ^d	0.37	30,000	21	99.9
716	Benzene	12.2	0.41	487	0.155	280	2.8 ^f	99
4113	iso-Butanol	15.2	0.91	1,816. ^d	0.5	400,000. ^j	4,000	99
414	n-Butanol							
3310	Ethylmorpholine, Methylmorpholine, Morpholine	12.2	0.30	686. ^m	0.36 ⁿ	28,300	2,700	95 90 10

TABLE D-2. STEAM STRIPPER ORGANIC COMPOUND REMOVAL PERFORMANCE
(CONTINUED)

^aCHC = Chlorinated hydrocarbons.

^bNA = not available.

^cHeight of packed section only. Total height is not available.

^dThe feed flow rate was calculated by dividing the annual steam usage by the steam-to-feed ratio (SFR).

^eFeed stream volatile organic concentration is the weighted average of volatile organic concentrations in several combined process wastewater streams.

^fThe bottoms volatile organic concentration is calculated from the following relationship:

$$\text{Bottoms Concentration} = \text{Feed Concentration} (1 - \text{Stripper Efficiency Fraction})$$

^gThe feed flow rate was calculated from the sum of process streams into the stripper identified in the facility process flow diagram.

^hAssumed wastewater feed temperature of 35 °C.

ⁱCalculated the feed flow rate from known bottoms flow rate using the following relationship:

$$\text{Feed Flow Rate} = (\text{Bottoms Flow Rate}) / [0.99 (1 + \text{SFR})]$$

^jThe feed volatile organic concentration is calculated from the following relationship:

$$\text{Feed Concentration} = [\text{Bottoms Concentration}] / (1 - \text{Stripper Efficiency Fraction})$$

^kAssumed that the two reported wastewater feed streams were of equal flow and calculated an average volatile organic concentration.

^lBatch steam stripper--Assumed median flow rate of reported range.

^mThe flow rate was calculated from stripper bottom flow range code. The mid-point of the flow range was used for the calculation.

ⁿThe reported steam flow rate of 250 kg/hr was divided by the calculated feed flow rate of 686 kg/hr.

D.2.2 Development of Equipment Design and Operating Parameters

The equipment design and operating parameters for the example wastewater stream were developed through a design evaluation performed using Advanced System for Process Engineering (ASPEN)²⁰, a computer software program intended for the rigorous design of distillation columns. The major design parameters in the ASPEN steam stripper model are based on field experience and published information. The diameter was calculated assuming a velocity of 80-percent flooding conditions. In addition, the following engineering assumptions were made:

- Operating pressure of one atmosphere;
- Isothermal column operation;
- Constant molal overflow (i.e., one mole of aqueous phase vaporized for each mole of steam condensed); and
- Linear equilibrium and operating equations (i.e., Henry's Law is valid for each organic compound at the concentrations encountered in the stripping column).

The design stripper contains 10 trays. A tray efficiency of 80 percent was assumed to estimate the actual number of stages for the column. A tray spacing of 0.50 m (1.64 ft) was assumed to estimate the active column height. To approximate the total column height, a total of 2.5 m (8.20 ft) of nonactive entrance and exit column was assumed. The system components are discussed in the following paragraphs.

The wastewater storage tank is sized to provide a desired retention time of 48 hours for the stripper feed stream, assuming 1500 m (4921 ft) of piping will be needed to combine multiple batch and/or continuous streams for treatment by the same steam stripper.

The remaining equipment in the steam stripper system was designed using ASPEN. The steam stripper column is designed as a sieve tray unit with countercurrent flow. The column is operated at a typical steam-to-wastewater-feed ratio of

0.096 kg of steam per ℓ (0.80 lb/gal) of wastewater. The liquid loading of the column is $39,900 \ell/\text{hr}/\text{m}^2$ ($979 \text{ gal}/\text{hr}/\text{ft}^2$).

Emphasis was placed on generating a design that would be most cost effective, would be within practical design parameters, and would remove virtually 100 percent of the highly volatile compounds. (The controlling compound used for design purposes was benzene.) A column height of 9 m (30 ft) with a total of 10 sieve trays is used for the steam stripper unit.

The overheads from the steam stripper are recovered with a water-cooled condenser. The condenser is designed for an outlet vapor temperature of 50°C (122°F) with an overall heat-transfer coefficient of $1,000 \text{ j}/\text{m}^2/\text{s}/^\circ\text{K}$ ($0.05 \text{ Btu}/\text{ft}^2/\text{s}/^\circ\text{F}$). The organic phase of the overheads stream is recovered from the overheads decanter. The overheads vapor from the primary condenser is assumed to be vented to the feed storage tank which is routed to an existing on-site combustion device or other control device.

The bottoms from the steam stripper are fed to the existing wastewater treatment facility. Before discharge from the stripper system, the bottoms pass through a feed preheater to enhance the efficiency of the steam stripper. The overall heat-transfer coefficient used by ASPEN for the feed preheater is $1,000 \text{ j}/\text{m}^2/\text{s}/^\circ\text{K}$ ($0.05 \text{ Btu}/\text{ft}^2/\text{s}/^\circ\text{F}$).

Pumps are installed to transfer the wastewater from the feed tank to the stripper, from the stripper to the feed/bottoms heat exchanger, from the decanter to the collection pot, and from the collection pot to storage.

There are five vents included in the system design for the storage tank and decanter. The vapors from these vents are ducted to a flare or other control device.

D.3 SYSTEM COSTING

Total capital investment and total annual cost for the example steam stripper system are estimated using the methodology described in Section 3.2.3 of this volume.

D.3.1 Total Capital Investment

Steam stripper costs are estimated using the equipment size generated by ASPEN. The purchase cost of each piece of process equipment is determined from various costing algorithms. Purchased equipment costs are then used to estimate total installation costs, and finally total capital investment.²¹ The cost-estimating techniques presented are based upon the size or capacity of the equipment and are derived from actual construction projects. Table D-3 summarizes the costing algorithms and Table D-4 summarizes the estimated equipment costs calculated for each component, the Table D-4 estimated size or capacity, and the reference or information source used to obtain the cost estimate for treating 500 lpm (132 gpm). The initial estimates were based on the equipment costs for the year in which the textbook or journal article was published. These costs were then adjusted to July 1989 dollars using the *Chemical Engineering* fabricated equipment index for the appropriate month and year. The cost for each individual component was summed to yield the basic equipment cost for the example wastewater stream.

The basic equipment cost was then used to calculate the purchased equipment cost, direct installation costs, and indirect installation costs according to the equations given in Section 3.2.3 and summarized here in Table D-5. Table D-5 also shows the cost for each of these items.

D.3.2 Total Annual Cost

Total annual cost for the example steam stripper system is based on the operating parameters developed using ASPEN, and the methodology given in Section 3.2.3 of this volume. Table D-6 summarizes the cost equations and the costs calculated for utilities, labor, maintenance, and indirect annual costs.

A credit was also calculated for the organics that are recovered through steam stripping. Although there are several options available for disposal or use of the recovered organic stream, for this cost estimate it is assumed that the organics

TABLE D-3. COSTING EQUATIONS FOR A STEAM STRIPPING UNIT

Equipment Component	Costing Equation	Cost Index ^a	Qualifier ^b	Cost Reference
Feed Tanks	$\exp[2.331+1.367*\ln(v)-0.063088*\ln(v)^2]$ $\exp[11.362-0.6104*\ln(v)+0.45355*\ln(v)^2]$	230.9	1,300 gal < V ≤ 21,000 gal 21,000 gal < V < 11,000,000 gal	22 22
Feed Preheater	$\exp[8.551-0.30863*\ln(A)+0.06811*\ln(A)^2]$ $(3\exp[-1.1156+0.0906*\ln(A)])$	230.9	150 ft ² < A < 12,000 ft ²	22
Steam Stripper	(1A + 1B + 1C + 2A + 2B + 2C + 2D)/2			
1A. Column Shell, Skirts, Nozzles and Manholes	$\exp[6.823+0.14178*\ln(W_S)+0.02468*\ln(W_S)^2]$	230.9	9020 lb < W _S < 2,470,000 lb	24
1B. Platforms and Ladders	$151.81*D_i(0.633616)*L_t(0.80161)$	230.9	3 ft < D _i < 24 ft; 57.5 ft < L _t < 150 ft	24
1C. Sieve Stainless Steel Trays	$0.85*(1.189+0.0577*D)^*$ (No. of Trays)*278.38*exp[0.1739*D]	230.9	2 ft < D < 16 ft	24
2A. Column Shell	83.69*W _S (0.6124)	225.9	1,000 lb < W _S < 70,000 lb	23
2B. Manholes	(No. of Manholes)*18*(53.83+40.71*TS)	225.9	0.375 inch < T _S < 2 inch	23
2C. Nozzles	each nozzle length (No. of nozzels)*(length of i=1 nozzles, inch)*[24.57+35.94*T _S]	225.9	0.375 inch < T _S < 2 inch	23
2D. Sieve Stainless Steel Trays	(no. of Trays)*214.54*exp[0.2075*D _i]	225.9	2 ft < D _i < 12.5 ft	23
Primary Condenser	2228.8*exp[0.00411*A]	343	150 ft ² < A < 240 ft ²	25
	5328.8*exp[0.0008762*A]	343	240 ft ² < A < 1500 ft ²	25

TABLE D-3. COSTING EQUATIONS FOR A STEAM STRIPPING UNIT
(Continued)

Equipment Component	Costing Equation	Cost Index ^a	Qualifier ^b	Cost Reference
Overhead Collection Decanter	$74.55 \cdot (V/0.5)^{0.5662}$	225.9	100 gal < V < 100,000 gal	26
Flame Arrestor	\$100/arrestor	319	NA	27
Pumps	$8740.7 \cdot W^{0.4207}$; W in hp	347.8	10 gpm < Q < 150 gpm	25
	$13783.4 \cdot W^{0.2890}$; W in hp	347.8	30 gpm < Q < 900 gpm	25

^aThe July 1989 Chemical Engineering fabricated equipment cost index is 356.0.

^b

- A = Surface Area
- D = Steam stripper outside column diameter
- D_i = Steam stripper inside column diameter
- L_t = Steam stripper tangent-to-tangent length
- Q = Flow rate through component
- T_s = Steam stripper column wall thickness
- V = Volume of component
- W = Work
- W_s = Steam stripper column weight

TABLE D-4. ESTIMATION OF BASIC EQUIPMENT COST FOR A STEAM STRIPPING UNIT

Equipment Component	Equipment Size ^a	Construction Material	Equipment Costs ^b (\$)	Cost Reference
Feed Tanks	1600 m ³ (56,500 ft ³)	Carbon Steel	99,000	22
Feed Preheater (shell and tube)	400 m ² (4300 ft ²)	Carbon Steel	50,000	22
Steam Stripper	0.98 m diameter (3.2 ft)	Carbon Steel	59,000	23, 24
Trays	9 m height (30 ft)	Trays-Stainless		
Primary Condenser (water cooled, shell and tube)	36.1 m ² (388 ft ²)	Carbon Steel	8,000	25
Overhead Collection Decanter	10.8 m ³ (381 ft ³)	Carbon Steel	11,000	26
Flame Arrestor	One arrestor per ventline	NAC	1,000	27
Pumps (4)	10,800 total Watts (14.5 Hp)	Stainless Steel	56,000	25
TOTAL EQUIPMENT COST (BEC) ^d			\$284,000	

^aBased on 500 lpm wastewater flow.^bJuly 1989 dollars.^cNAC = Not applicable.^dBEC = Base equipment cost.

**TABLE D-5. ESTIMATION OF TOTAL CAPITAL INVESTMENT FOR A
STEAM STRIPPING UNIT^{a,b}**

Cost Component	Cost Factor	Component Cost ^a	Total Capital Investment	Cost Reference
<u>Direct Equipment Costs^b</u>				
Base Equipment Cost (BEC) ^c	TABLE D-4	\$284,000		
Piping ^d	\$36.83/m (\$11.23/ft)	\$58,000		28
Instrumentation	0.1 * [BEC + Pipe]	\$34,000		29
Sales Tax and Freight	0.08 * [BEC + Pipe]	\$30,000		29
Purchased Equipment Cost (PEC) ^e			\$406,000	
<u>Direct Installation Costs</u>				
Foundations and Supports	12% of PEC	\$49,000		29
Electrical	1% of PEC	\$4,000		29
Erection and Handling	40% of PEC	\$162,000		29
Painting	1% of PEC	\$4,000		29
Insulation	1% of PEC	\$4,000		29
TOTAL DIRECT INSTALLATION COST			\$223,000	
<u>Indirect Installation Costs</u>				
Engineering and Supervision	10% of PEC	\$41,000		29
Construction & Field Expense	10% of PEC	\$41,000		29
Construction Fee	10% of PEC	\$41,000		29
Startup and Testing	1% of PEC	\$4,000		29
Contingency	3% of PEC	\$12,000		29
TOTAL INDIRECT INSTALLATION COST			\$140,000	
TOTAL CAPITAL INVESTMENT			\$768,000	

^aJuly 1989 dollars.

^bBased on 500 lpm wastewater flow.

^cBEC = Base equipment cost.

^dAdditional piping for combination of five wastewater streams is assumed to total approximately 1,500 m (4,900 ft). Vapor vent lines required for storage tanks and decanters. Each vent line was assumed to be 11 m (36 ft) in length and constructed of 5.1-cm (2-in) diameter schedule 40 steel pipe.

^ePEC = Purchased equipment cost.

**TABLE D-6. ESTIMATION OF TOTAL ANNUAL COST
FOR A STEAM STRIPPING UNIT^{a,b}**

Cost Component	Cost Factor	Annual Consumption	Annual Cost	Cost Reference
<u>Direct Annual Costs</u>				
Utilities				
Electricity	\$0.0509/kWhr	78,000 kWhr ^c	\$4,000	30
Steam	\$7.68/Mg (\$7.77/ton)	29,000 Mg ^d (32,000 tons)	\$220,000	30
Water	\$0.058/1,000 ℓ (\$0.22/1,000 gal)	780,000,000 ℓ ^e (206,900,000 gal)	\$45,000	31
Labor				
Operating Labor	\$13.20/hr	450 hrs	\$6,000	29
Supervision & Administration	15% of Operating Labor		\$1,000	29
Maintenance				
Labor	\$14.50/hr	450 hrs	\$7,000	29
Materials	100% of Maintenance Labor		\$7,000	29
TOTAL DIRECT ANNUAL COST			\$290,000	
<u>Indirect Annual Costs</u>				
Overhead	60% of All Labor and Materials		\$13,000	29
Property Taxes	1% of TCI ^f		\$8,000	29
Insurance	1% of TCI		\$8,000	29
Administrative Charges	2% of TCI		\$15,000	29
Capital Recovery	10% @ 15 yrs		\$101,000	
TOTAL INDIRECT ANNUAL COST (TIAC)			\$145,000	
RECOVERY CREDIT			\$17,000 ^g	
TOTAL ANNUAL COST			\$418,000	
ANNUAL WASTE THROUGHPUT		220,000 Mg/yr		
COST PER UNIT WASTEWATER (\$/MG)	TAC/AWT ^h		\$1.90	
COST PER LITER WASTEWATER FEED (\$/ℓ)	TAC/FLOW ^j	260,000,000 ℓ/yr	\$0.0016	

^aJuly 1989 dollars.

^bBased on 500 ℓpm (132 gpm) wastewater flow.

^c260 kWhr/day, 300 days/yr.

^d97,000 kg/day (107 tons/day), 300 days/yr.

^e2,600,000 ℓ/day (690,000 gpd), 300 days/yr.

^fTCI = Total capital investment

^gRecovery credit based on approximately 28,000 KJ/Kg (12,000 Btu/lb) heating value.

^hTDAC + TIAC = RC = Total direct annual cost + Total indirect annual cost - Recovery credit.

ⁱTAC/AWT = Total annual cost per annual waste throughput.

^jTAC/FLOW = Total annual cost per flow.

will be used as fuel for an existing boiler. A heating value of approximately 28,000 KJ/Kg (12,000 Btu/lb) was developed to estimate the recovery credit in this example.³²

D.4 References

1. U. S. Environmental Protection Agency, Office of Air Quality Planning and Standards. Control Technology Center Document for Industrial Wastewater Volatile Organic Compound Emissions -- Background Information for BACT/LAER Determinations. EPA 450/3-90-004. Research Triangle Park, NC. January 1990. pp. 4-4 to 4-5.
2. Letter and attachments from Radian Corporation to P.E. Lassiter, P.E., EPA/OAQPS. July 17, 1987. The summary of data provided in confidential Section 114 questionnaire responses.
3. Trip Report. Howle, R.H. and Vancil, M.A., Radian Corporation, to file. May 12, 1987. Visit to Allied Fibers.
4. Trip Report. Herndon, D.J. and Buchanan, S.K., Radian Corporation, to file. May 22, 1987. Visit to Rhone-Poulenc/Agricultural (RP Ag) Company.
5. Trip Report. Herndon, D.J. and Buchanan, S.K., Radian Corporation, to file. September 2, 1987. Visit to Fritzsche Dode & Olcott.
6. Trip Report. Herndon, D.J., Radian Corporation, to file. May 6, 1987. Visit to PPG Industries.
7. Trip Report. Herndon, D.J. and Buchanan, S.K., Radian Corporation, to file. May 21, 1987. Visit to Mobay Chemical Company.
8. Trip Report. Herndon, D.J., Radian Corporation, to file. May 4, 1987. Visit to Borden Chemical Company.
9. Trip Report. Herndon, D.J., Radian Corporation, to file. May 5, 1987. Visit to Union Carbide Corporation.
10. Trip Report. Herndon, D.J., Radian Corporation, to file. May 8, 1987. Visit to Dow Chemical Company.
11. U. S. Environmental Protection Agency. Hazardous Waste Treatment, Storage, and Disposal Facilities (TSDF) - Background Information for Proposed RCRA Air Emission Standards, Volume 2: Appendices. Preliminary Draft. Research Triangle Park, NC. March 1988. pp. F-146 to F-149.
12. Ref. 11, pp. F-151 to F-147.
13. Ref. 11, pp. F-151 to F-155.

14. Memorandum from Herndon, D. J., Radian Corporation, to Industrial Wastewater file. May 20, 1988. Summary of facilities reporting use of steam stripping.
15. Environmental Science & Engineering, Inc., and SAIC. Plant No. 4 - Organic Chemicals Best Available Technology Long-Term Field Sampling. Prepared for U. S. Environmental Protection Agency, Office of Water Regulations and Standards. July 1985.
16. Ref. 15, Plant No. 7.
17. Ref. 15, Plant No. 15.
18. Code of Federal Regulations. Title 40, Part 455. Pharmaceutical Chemicals Category Effluent Limitations Guidelines, Pretreatment Standards, and New Source Performance Standards. Washington, DC. U.S. Government Printing Office. October 4, 1985.
19. Code of Federal Regulations. Title 40, Part 439. Pharmaceutical Manufacturing Point Source Category Effluent Limitations Guidelines, Pretreatment Standards, and New Source Performance Standards. Washington, DC. U.S. Government Printing Office. October 27, 1983.
20. Advanced System for Process Engineering (ASPEN). Massachusetts Institute of Technology for the Department of Energy. DOE/MC/16481-(3 vols) 1201, 1202, 1203. 1981.
21. U. S. Environmental Protection Agency. Office of Air Quality Planning and Standards. OAQPS Control Cost Manual. Chapter 2: Cost Estimating Methodology. 4th Edition. EPA-450/3-90-006. Research Triangle Park, N.C. January 1990. pp. 2-5 to 2-8.
22. Corripio, A.B., K.S. Chrien, and L.B. Evans. Estimate Costs of Heat Exchangers and Storage Tanks via Correlations. Chem.Eng. p. 145. January 25, 1982.
23. Peters, M.S., and K.D. Timmerhaus. Plant Design and Economics for Chemical Engineers. Third Ed. New York, McGraw-Hill Book Company. 1980. pp. 768-773.
24. Corripio, A.B., A. Mulet, and L.B. Evans. Estimate Costs of Distillation and Absorption Towers via Correlations. Chem.Eng. December 28, 1981. pp. 180-82.
25. Hall, R.S., W.M. Vataavuk, J. Matley. Estimating Process Equipment Costs. Chem.Eng. November 21, 1988. pp. 66-75.
26. Ref. 23, p. 572, Figure 13-58.

27. Telecon. Gitelman, A., Research Triangle Institute with Hoyt Corporation. Cost of flame arrestors. September 9, 1986.
28. Richardson Process Plant Construction Estimation Standards: Mechanical and Electrical. Volume 3, Richardson Engineering Services, Inc., Mesa, AZ, 1988. pp. 15-40.
29. Vataavuk, W.M., and R.B. Neveril. Part II: Factors for Estimating Capital and Operating Costs. Chem.Eng. November 3, 1980. pp. 157-182.
30. Memorandum from Peterson, P., Research Triangle Institute, to Thorneloe, S., EPA/OAQPS. January 18, 1988. Basis for steam stripping organic removal efficiency and cost estimates used for the source assessment model (SAM) analysis.
31. Ref. 2, p. 4-27.
32. Memorandum from Watkins, S., Radian Corporation, to Lassiter, P., EPA/CPB. January 20, 1992. Development of a recovery credit for organic compounds removed from wastewater streams by the design steam stripper for the HON.

APPENDIX E

EXAMPLE COSTS FOR THE INSTALLATION OF AN INTERNAL FLOATING ROOF IN A FIXED ROOF TANK

This appendix presents a set of example calculations for estimating the total annual cost for the installation and operation of an internal floating roof in an example fixed roof tank.

The purpose of this appendix is to demonstrate the approach used in the HON analysis. In the calculations below, all significant figures have been retained until the final calculation to make it easier for the reader to follow the calculation and to avoid potential error due to round off of intermediate calculations. It should not be inferred that the intermediate results represent the actual number of significant figures.

The example tank stores styrene and has a capacity of 2,000,000 gallons. Additional design parameters for the example tank are presented in Table E-1 of this appendix. The internal floating roof, installed in this example tank, will have a liquid-mounted primary seal and controlled deck fittings.

1. Before an internal floating roof can be installed in the example tank, the tank must be cleaned and degassed. The cost for cleaning and degassing a tank is based on its capacity.

$$\begin{aligned}\text{Cstdegas} &= 7.61 (\$/\text{gal}) * [\text{tank_size, gal}]^{0.5132} \\ \text{Cstdegas} &= 7.61 * [2,000,000 \text{ gal}]^{0.5132} \\ \text{Cstdegas} &= \$13,034 \text{ per tank}\end{aligned}$$

TABLE E-1. MODEL TANK DESIGN PARAMETERS

Parameter Description	Units	Value
Tank capacity (tank_size)	gal	2,000,000
Annual tank throughput	gal/yr	18,057,775
Tank orientation and type	---	vertical fixed roof
Tank diameter (tank_dia)	feet	85
Tank height	feet	47
Average tank vapor space height	feet	23.6
Adjustment factor for small diameter tanks	dimensionless	1.0
Tank paint factor - white roof and aluminum color shell	dimensionless	1.3
Average ambient diurnal temperature change	°F	20
Product factor - organic liquid other than crude oil	dimensionless	1.0
Stored product - HAP	---	styrene
Product molecular weight	lb/lb mol	104.16
Product specific gravity	---	0.906
Product vapor pressure at 25 °C	mmHg psia	6.591 0.12748
Product Antoine coefficients		
A	dimensionless	7.140
B	dimensionless	1574.51
C	dimensionless	224.09
Product average market price	\$/kg	0.73

2. Determine the installed capital cost for a new internal floating roof having a primary liquid-mounted seal and controlled deck fittings.

$$\text{Cstroof} = [509 (\$/\text{ft}) * \text{tank_dia}(\text{ft})] + 1160 (\$)$$

$$\text{Cstroof} = [509 (\$/\text{ft}) * 85 (\text{ft})] + 1160 (\$)$$

$$\text{Cstroof} = \$44,425$$

3. The total capital investment (TCI) for the installation of an internal floating roof in the example tank is the sum of the tank preparation costs and the installed capital cost for the roof.

$$\begin{array}{rcl} \text{Total} & & \text{Installed} \\ \text{capital} & & \text{capital} \\ \text{investment} & = & \text{and} + \text{cost for} \\ \text{(TCI)} & & \text{degassing cost floating roof} \end{array}$$

$$\text{TCI (July 1989\$)} = \$13,034 + \$44,425$$

$$\text{TCI (July 1989\$)} = \$57,459$$

4. Indirect annual costs (ann_fix) are estimated as a percentage of the total capital investment (TCI) and include capital recovery; maintenance charges (5 percent); inspection charges (1 percent); and taxes, insurance, and administrative charges (4 percent). The capital recovery factor (0.163) is based on 10 percent interest over 10 years.

$$\text{ann_fix (July 1989\$)} = \$57,459 * [0.163 + 0.05 + 0.01 + 0.04]$$

$$\text{Indirect Annual Cost (July 1989\$)} = \$15,112/\text{yr}$$

5. The direct annual cost (ann_var) is a cost savings equal to the value of the recovered product. This product recovery credit is based on the HAP emission reduction achieved by the internal floating roof and the market value of the HAP.

$$\begin{array}{lcl} \text{Direct Annual} & \text{HAP emissions} & \left[\begin{array}{l} \text{-Average market} \\ \text{value of HAP} \end{array} \right] \\ \text{Cost} & = \text{reduction} * & \\ \text{(July 1989 \$)} & \text{(Mg/yr)} & \text{(\$ /Mg)} \end{array}$$

$$\text{ann_var} = 5.17 \text{ (Mg/yr)} * [-0.73 \text{ (\$/kg)}] * 1,000 \text{ (kg/Mg)}$$

$$\text{Direct Annual Cost (July 1989\$)} = -\$3,774/\text{yr}$$

6. The total annual cost (TAC) for the installation and operation of an internal floating roof is the sum of the indirect annual cost and direct annual cost.

$$\text{Total annual cost} = \text{Indirect annual cost} + \text{Direct annual cost}$$

$$\text{Total annual cost} = \$15,112 + [-\$3,774]$$

$$\text{Total annual cost} = \$11,338/\text{yr}$$

APPENDIX F

EXAMPLE COST METHODOLOGY FOR EQUIPMENT LEAKS

The purpose of these calculations is to demonstrate the methodology used for calculating the cost of controlling emissions from equipment leaks. Note that the sources of all data used in the calculations are documented in the memorandum "Final Cost Impacts Analysis for HON Equipment Leaks," April 15, 1991, contained in the equipment leaks docket. The methodology is first described and then applied to an example plant. Note that the costs presented reflect estimated costs associated with implementing control as required in the negotiated regulation. These costs may not be applicable to other potential equipment leak standards nor to any individual process unit. The following steps summarize how the costs are calculated.

1. Determine a base cost for each type of component for which a control equipment requirement is set by the negotiated regulation. Cost are based on information collected from vendors and published sources. Convert all costs to July 1989 dollars.
2. Determine capital costs for process units by multiplying the equipment count by the base cost for each type of component. To this, add the purchase cost of one monitoring instrument for each process unit. (Note that process units controlled at baseline will not incur capital costs for monitoring instruments and caps for open-ended lines.)
3. Determine initial monitoring and leak repair cost estimates for uncontrolled process units, based on

equipment counts, initial leak frequencies, and subsequent repair requirements.

4. Determine annual leak detection and repair cost estimates for all process units based on equipment counts, monitoring frequencies, leak frequencies, and subsequent repair requirements.
5. Determine the annualized capital costs by multiplying the capital cost values for each process unit by the capital recovery factor which is determined based on interest rate and equipment life.
6. Determine annual costs for maintenance and miscellaneous charges associated with equipment leak control.
7. Determine the total annual cost by subtracting the recovery credit from other annual costs (annualized capital costs and annual operating costs). The recovery credit is determined by multiplying the annual emission reduction by the average VOC cost.

F.1 BASE COSTS DEVELOPMENT

The following equations estimate base costs for control equipment including a monitoring instrument, replacement pump seals, compressors, pressure relief devices, open ended lines, and sample connections. Chemical Engineering Plant Cost Index values are used throughout the base cost development to convert all costs to July 1989 dollars. A summary of the base costs for control equipment is presented in Table F-1. Note that total values for control equipment may not exactly add up due to rounding errors.

Monitoring Instrument:

The price of a monitoring instrument is approximately \$6,500 in July 1989 dollars.

Replacement Pump Seals:

The fourth quarter 1978 cost of a replacement seal for a pump, including a 50 percent credit for the old seal, was \$113.

TABLE F-1. BASE COSTS FOR CONTROL EQUIPMENT

Equipment Type	Base Cost (per component) (July 1989 \$)
Monitoring Instrument	6,500
Replacement Pump Seal	180
Compressor	6,240
Pressure Relief Device: ^a	
Rupture Disks	78
Holders, Valves, Installation, etc.	3852
Open-Ended Lines	102
Sample Connections	408

^aPressure Relief Device costs are split into two portions because Rupture Disks have a 2-year equipment life and all other pieces (Holders, Valves, etc.) have a 10-year life.

Using the averaged Chemical Engineering Plant Cost Index Ratio of (356.0/224.7) to convert to July 1989 dollars:

$$\$113 * (356.0/224.7) = \$180.$$

Compressors (Closed-Vent System):

A closed vent system for a compressor consists of piping, plug valves, and a flame arrestor. The costs for equipment and installation of a closed-vent system are described in the following steps.

1. 122 m of 5.1 cm diameter schedule 40 steel pipe. Piping costs are \$261.03/100 ft in 1988 dollars. Using the Chemical Engineering Plant Cost Index Ratio of (356.0/342.5) to convert to July 1989 dollars:

$$(122 \text{ m}) * (3.28 \text{ ft/m}) * (\$261.03/100 \text{ ft}) * (356.0/342.5) = \$1090.$$

2. Installation of this piping will require 66 hrs of labor at a rate of \$22.50/hr:

$$(66 \text{ hrs}) * (\$22.50/\text{hr}) = \$1490.$$

3. Three 5.1 cm cast steel plug valves are required at a 1988 cost of \$783.75 each. Use the Chemical Engineering Plant Cost Index Ratio of (356.0/342.5) to convert to July 1989 dollars.

$$3 * (\$783.75) * (356.0/342.5) = \$2,440$$

4. Installation of these valves will require 2 hours of labor at a rate of \$22.50/hr:

$$3 * (\$22.50/\text{hr}) * (2 \text{ hrs}) = \$135.$$

5. One metal gauze flame arrestor will cost \$869 based a 1990 vendor quote. Use the Chemical Engineering Plant Cost Index Ratio of (356.0/358.7) to convert to July 1989 dollars:

$$(\$869) * (356.0/358.7) = \$860.$$

6. Installation of the flame arrestor will require 10 hrs of labor at a rate of \$22.50/hr:

$$(\$22.50/\text{hr}) * (10 \text{ hrs}) = \$225.$$

Total Compressor Closed Vent Cost:

$$\$1,090 + \$1,490 + \$2,440 + \$135 + \$860 + \$225 = \$6,240.$$

Pressure Relief Devices (Rupture Disk Assembly):

A rupture disk assembly for a pressure relief device consists of a rupture disk, rupture disk holder, pressure gauge, bleed valve, gate valve, offset, and relief valve. The costs for equipment and installation of a rupture disk assembly for a pressure relief device are as follows:

1. Three 7.6 cm stainless steel rupture disks cost \$235 in 1990. Use the Chemical Engineering Plant Cost Index Ratio of (356.0/358.7) to convert to July 1989 dollars:

$$[(\$235)/(3)] * (356.0/358.7) = \$78.$$

2. One 7.6 cm carbon steel rupture disk holder cost \$583 in 1990. Use the Chemical Engineering Plant Cost Index Ratio of (356.0/358.7) to convert to July 1989 dollars:

$$(\$583) * (356.0/358.7) = \$580.$$

3. One 0.6 cm pressure gauge with dial face cost \$15 in last quarter 1978. Use the Chemical Engineering Plant cost Index Ratio of (356.0/224.7) to convert July 1989 dollars:

$$(\$15) * (356.0/224.7) = \$24.$$

4. One 0.6 cm carbon steel bleed valve cost \$25 in last quarter 1978. Use the Chemical Engineering Plant Cost Index Ratio of (356.0/224.7) to convert to July 1989 dollars:

$$(\$25) * (356.0/224.7) = \$40.$$

5. Installation of the pressure gauge and bleed valve will require 16 hours of labor at a rate of \$22.50/hr:

$$(16 \text{ hours}) * (\$22.50/\text{hr}) = \$360.$$

6. One 7.6 cm gate valve cost \$385.60 in 1988. Use the Chemical Engineering Plant Cost Index Ratio of (356.0/342.5) to convert to July 1989 dollars:

$$(\$385.60) * (356.0/342.5) = \$400$$

7. Installation of the gate valve will require 1.5 hours of labor at a rate of \$22.50/hr:

$$(1.5 \text{ hours}) * (\$22.50/\text{hr}) = \$34$$

8. An offset consisting of one 10.2 m tee and one 10.2 cm elbow cost \$47.36 in 1988. Use the Chemical Engineering Plant Cost Index Ratio of (356.0/342.5) to convert to July 1989 dollars:

$$(\$47.36) * (356.0/342.5) = \$49.$$

9. Installation of the offset will require 8.5 hours of labor at a rate of \$22.50/hr:

$$(8.5 \text{ hours}) * (\$22.50/\text{hr}) = \$191.$$

10. Retrofitting of existing pressure relief valves will require a new 7.6 cm stainless steel body and trim relief valve which cost \$1230.80 in the last quarter of 1978. Use the Chemical Engineering Plant Cost Index Ratio of (356.0/224.7) to convert July 1989 dollars:

$$(\$1230.80) * (356.0/224.7) = \$1,950.$$

11. Installation of the relief valve will require 10 hours of labor at a rate of \$22.50/hr:

$$(\$22.50/\text{hr}) * (10 \text{ hrs}) = \$225.$$

Total Pressure Relief Device Cost:

$$= \$78 + \$580 + \$24 + \$40 + \$360 + \$400 + \$34 + \$49 + \$191 + \$1950 + \$225 = \$3,930.$$

Open-Ended Lines:

Control equipment for an open ended line consists of one 2.5 cm gate valve which cost \$67.50 in 1988. Use the Chemical Engineering Plant Cost Index Ratio of (356.0/342.5) to convert to July 1989 dollars:

$$(\$67.50) * (356.0/342.5) = \$70.$$

Installation of the gate valve will require 1.4 hours of labor at a rate of \$22.50/hr:

$$(1.4 \text{ hours}) * (\$22.50/\text{hr}) = \$32.$$

Total Open-Ended Line Cost = \$70 + \$32 = \$102.

Sample Connection:

A closed purge sample connection system consists of piping and three ball valves. The costs for equipment and installation are as follows:

1. 6 m of 2.5 cm diameter schedule 40 steel pipe cost \$120.64/100 ft in 1988. Use the Chemical Engineering Plant Cost Index Ratio of (356.0/342.5) to convert to July 1989 dollars:

$$(6 \text{ m}) * (3.28 \text{ ft/m}) * (\$120.64/100 \text{ ft}) * (356.0/342.5) = \$25.$$
2. Installation of the piping will require 1 hour of labor at a rate of \$22.50/hr:

$$(1 \text{ hour}) * (\$22.50/\text{hr}) = \$23.$$
3. One 2.5 cm carbon steel ball valve cost \$85.60 in 1988. Use the Chemical Engineering Plant Cost Index Ratio of (356.0/342.5) to convert to July 1989 dollars:

$$(3) * (\$85.60) * (356.0/342.5) = \$265.$$
4. Installation of the three valves will require 1.4 hrs of labor each at a rate of \$22.50/hr:

$$(3) * (1.4 \text{ hours}) * (\$22.50/\text{hr}) = \$95.$$

Total Sample Connection Cost:

$$\$25 + \$23 + \$265 + \$95 = \$408$$

F.2 CAPITAL COST DEVELOPMENT

Two sets of equations are used to develop capital costs for the process units, one for controlled units and one for uncontrolled units. Note that open-ended lines are routinely controlled in all process units. If an open-ended line is already controlled with a second valve, blind flange, or cap, then capital costs for that open-ended line are zero dollars. These equations are presented below.

Controlled Units

(Compressor Count) * (\$6,240)	=	Compressor Capital
(Pressure Relief Device Count * (\$78)	=	Rupture Disk Capital
(Pressure Relief Device Count) * (\$3,852)	=	Holder, Valve, Etc. Capital

(Sample Connection Count) * (\$408)	=	Sample Connection Capital
Sum of above	=	Total Capital for Controlled Unit

Uncontrolled Units

(1 Monitoring Instrument) * (\$6,500)	=	Monitoring Instrument Capital
(Compressor Count) * (\$6,240)	=	Compressor Capital
(Pressure Relief Device Count) * (\$78)	=	Rupture Disk Capital
(Pressure Relief Device Count) * (\$3,852)	=	Holder, Valve, Etc. Capital
(Open-Ended Line Count) * (\$102)	=	Open-Ended Line Capital
(Sample Connection Count) * (\$408)	=	Sample Connection Capital
Sum of above	=	Total Capital for Uncontrolled Unit

F.3 INITIAL LEAK DETECTION AND REPAIR COSTS

Leak detection and repair costs are estimated for uncontrolled and controlled process units. Uncontrolled process units have cost associated with implementing the leak detection and repair program. Subsequent to implementing the leak detection and repair program annual costs for uncontrolled and controlled process units are equal.

For process units which are uncontrolled, the costs for initial monitoring and leak repair valves in gas/vapor service, valves in light liquid service, pumps in light liquid

service, and connectors are estimated by the following procedure.

Monitoring Costs:

Monitoring costs are based on a subcontractor's monitoring fee of \$2.50/component/monitoring event. This cost includes a first attempt at repair.

$$(\text{Equipment Type Count}) * (\$2.50/\text{component}) = \text{Monitoring Cost}$$

This is reported for each type of equipment covered.

Initial Repair Cost:

Determination of the cost of initial repairs requires three steps. These steps are to determine the total number of leaks found, determine the number of leaks requiring repair, and determine the cost of repairs. Each step is described below.

1. The total number of leaks found during monitoring is estimated by multiplying the number of components of a specific type by the initial leak frequency. The initial leak frequencies are:

Valves in gas/vapor service	= 0.114
Valves in light liquid service	= 0.065
Pump seals in light liquid service	= 0.20
Connectors	= 0.021

$$(\text{Equipment Type Count}) * (\text{initial leak frequency}) = \text{Total number of leaks}$$

2. The number of leaks requiring further repair (beyond efforts made by the monitoring team) is determined by multiplying the number of leaks found by the percent of leaks requiring further repair. For valves and connectors, this is 25 percent. For pumps this is 75 percent. For example:

$$(\text{Pump leaks}) * (0.75) = \text{Number of pumps requiring further repair}$$

3. The cost of further repairs is determined for each equipment type by the time required for each repair times the labor rate of \$22.50/hr. The time required for each repair is:

Valves = 4 hours/repair
Pumps = 16 hours/repair
Connectors = 2 hours/repair

$$\left(\begin{array}{c} \text{Number of} \\ \text{pumps requiring} \\ \text{further repair} \end{array} \right) * \left(16 \frac{\text{hrs}}{\text{repair}} \right) * \left(\frac{\$22.50}{\text{hr}} \right) = \text{Pump repair cost}$$

The total cost for initial monitoring and repair is the sum of monitoring costs and repair costs.

$$(\Sigma \text{ Monitoring Costs}) + (\Sigma \text{ Repair Costs}) = \begin{array}{l} \text{Total Cost for} \\ \text{initial monitoring} \\ \text{and repair} \end{array}$$

F.4 ANNUAL LEAK DETECTION AND REPAIR COSTS

For all process units, the costs for annual leak detection and repair (LDAR) are estimated by the following procedure.

Monitoring Costs:

Monitoring costs are based on a subcontractor's monitoring fee of \$2.00/component/monitoring event for every monitoring after the first. This cost includes a first attempt at repair by the monitoring team. The price is lower than the initial monitoring fee because of reduced time requirements and lower leak frequency.

$$(\text{Equipment Type Count}) * (\text{Monitoring Frequency}) * (\$2.00/\text{component/event}) = \text{Instrument Monitoring Cost}$$

For pumps, an additional monitoring fee is added to cover weekly visual inspections. This cost is calculated by multiplying the number of pumps by the inspection rate of 1 inspection every 0.5 minutes times the frequency of 52 visual monitoring/year, and then times the labor rate of \$22.50/hr.

$$(\text{Pump count}) * (0.5 \text{ min/inspection}) * (1 \text{ hr}/60 \text{ min}) * (52 \text{ monitoring/yr}) * (\$22.50/\text{hr}) = \text{Visual Monitoring Cost}$$

Repair Costs:

The annual leak repair costs are determined using three steps: determine the total number of leaks found, determine the total number of leaks requiring repair, and determine the cost of repairs. These steps are discussed below.

1. The total number of leaks found during monitoring is estimated by multiplying the number of components of a specific type by the number of times the type of equipment is monitored annually times the initial leak frequency. The leak frequencies were taken from the negotiated regulation as the maximum value which did not require the implementation of a quality improvement plan. These leak frequencies are:

Valves in gas/vapor service	= 0.02
Valves in light liquid service	= 0.02
Pump seals in light liquid service	= 0.10
Connectors	= 0.005

The monitoring frequencies are as follows:

Valves in gas/vapor service	= monthly ^a
Valves in light liquid service	= monthly ^a
Pumps in light liquid service	= monthly
Connectors	= annually

$(\text{Equipment Count}) * (\text{Monitoring Frequency}) * (\text{Leak Frequency}) = \text{Total number of leaks}$

2. The number of leaks requiring further repair (beyond efforts made by the monitoring team) is determined by multiplying the total number of leaks found by the percentage of leaks requiring further repair. For valves and connectors this is 25 percent. For pumps this was determined to be 75 percent.
3. The cost of further repairs is estimated using the same approach as described for the initial repair costs.

^aValves are monitored monthly in this example, but if less than 2 percent are leaking, can be monitored quarterly.

The total cost for annual leak detection and repair is the sum of monitoring and repair costs, plus administrative and support cost. Administrative and support cost are estimated to be 40 percent of the monitoring and repair labor expenses:

$$(\text{Monitoring labor} + \text{Repair labor}) * (0.40) = \begin{matrix} \text{Administrative} \\ \text{and Support} \\ \text{Costs} \end{matrix}$$

Note that the cost for monitoring pressure relief devices once annually is included in the LDAR costs. Further repair of this equipment is not included, however, because it is assumed to be covered by plant maintenance costs not the LDAR program.

F.5 ANNUALIZED CAPITAL COSTS

The annualized capital costs for the process units is determined by multiplying the capital cost for each type of equipment by the Capital Recovery Factor for that type of equipment. The Capital Recovery Factors were calculated using the following equation:

$$CRF = \frac{i * (1 + i)^n}{(1 + i)^n - 1}$$

where: i = interest rate, expressed as a decimal
 n = economic life of the component, in years.

For all components, the interest rate is 10 percent or 0.10. For pump seals and rupture disks, the economic life is 2 years. For monitoring instruments it is 6 years, and for all other equipment it is 10 years. Then the CRF values are:

CRF: 2 years = 0.58
 CRF: 6 years = 0.23
 CRF: 10 years = 0.163

$$\left(\begin{matrix} \text{Component Type} \\ \text{Capital Cost} \end{matrix} \right) * (CRF) = \left(\begin{matrix} \text{Annualized} \\ \text{Capital Costs} \end{matrix} \right)$$

For uncontrolled units, the costs for initial LDAR are annualized. This is done by multiplying the total cost for

labor for the initial leak detection and repair by 1.4 to 40 percent for administration and support, and then multiplying by the CRF for 10 years of 0.163.

$$\left(\begin{array}{c} \text{Initial} \\ \text{Leak} \\ \text{Repair} \\ \text{Cost} \end{array} \right) * (1.4) * (0.163) = \left(\begin{array}{c} \text{Annualized} \\ \text{initial LDAR} \\ \text{Labor, Admin.,} \\ \text{and Support Cost} \end{array} \right)$$

The cost of replacement pump seals associated with the initial LDAR are also annualized. It was calculated by multiplying the number of pumps requiring further repair by the CRF for 2 years times the cost of \$180/pump seal.

$$\left(\begin{array}{c} \text{Number of} \\ \text{Pumps} \\ \text{Requiring} \\ \text{Repair} \end{array} \right) * (0.58) * \left(\frac{\$180}{\text{pump seal}} \right) = \left(\begin{array}{c} \text{Annualized cost} \\ \text{of pump seal} \\ \text{replacement} \end{array} \right)$$

F.6 ANNUAL MAINTENANCE COSTS AND MISCELLANEOUS COSTS

In addition to annual leak detection and repair program costs and annualized capital costs, there are annual maintenance charges and annual miscellaneous fees.

Annual Maintenance Charges:

Annual maintenance charges include maintenance costs associated with the monitoring instrument, control equipment installed on compressors, pressure relief devices, open-ended lines, and sampling connections, and replacement pump seals. These costs are described below.

1. In the last quarter of 1978, the cost for all materials and labor required for maintenance and calibration of the monitoring instrument was \$2,700. Updating this cost to July 1989 dollars is done using the Chemical Engineering Plant Cost Index Ratio of (356.0/224.7)

$$\$2700 * (356.0/224.7) = \$4,280$$

2. For compressors, pressure relief devices, open-ended lines, and sampling connections, the annual maintenance charge was calculated as 5 percent of the capital charge for the same equipment type.

$$(\text{Capital Cost}) * (0.05) = \text{Annual Maintenance Charge}$$

For this analysis, it was assumed that open-ended lines were controlled at baseline and would therefore incur no capital cost. However, in order to estimate the maintenance and miscellaneous charges associated with open-ended lines, it was necessary to calculate a capital cost value for use in the above equation.

3. Replacement of pump seals is considered an annual maintenance cost. The cost is calculated by multiplying the number of pumps found leaking as a result of the annual LDAR program by \$180/pump seal.

Annual Miscellaneous Costs:

The annual miscellaneous costs include taxes, insurance, and administration. For the monitoring instrument, compressors, pressure relief devices, open-ended lines, sampling connections, and pump seals. The annual miscellaneous charges are calculated as 4 percent of the capital cost.

$$(\text{Capital Cost}) * (0.04) = \text{Annual Miscellaneous Charges}$$

For replacement pump seals, the annual miscellaneous charge was calculated as 80 percent of the annual maintenance cost for pump seals.

F.7 TOTAL ANNUAL COSTS

Total annual costs include all annual cost described previously plus recovery credit associated with emission reductions. The recovery credit for VOC's which are not lost to the air through equipment leaks is calculated by multiplying the annual emission reduction by the average VOC cost.

$$\left(\frac{\text{Annual Emission}}{\text{Reduction}} \right) * \left(\frac{\text{Average VOC}}{\text{Cost (\$/Mg)}} \right) = \text{Recovery Credit}$$

Total annual cost for a process unit is determined by subtracting the Recovery Credit from other annual costs. Other annual costs include the sum of the annualized capital cost, and annual operating expenses for the LDAR program, maintenance charges, and miscellaneous costs.

$$\left(\frac{\text{Annualized Capital Costs}}{\text{Costs}} \right) + \left(\frac{\text{Operating Expenses}}{\text{Expenses}} \right) - \left(\frac{\text{Recovery Credit}}{\text{Credit}} \right) = \text{Total Annual Cost}$$

F.8 EXAMPLE CALCULATIONS

Example calculations are provided for estimation of annualized capital costs and operating costs for uncontrolled and controlled hypothetical process units. Total annual costs are not calculated in these examples. BID Volume 1C, Appendix E demonstrates calculation of total annual costs. Note that all summations are rounded to 100's of dollars.

The following calculations are for an uncontrolled process unit which has the following hypothetical equipment counts:

Valves in gas/vapor service	414
Valves in light liquid service	1,179
Pumps in light liquid service	40
Connectors	2,662
Pressure Relief Devices (PRVs)	45
Compressors	2
Open-Ended Lines (OELs)	141
Sample Connection	35

Capital Costs

Monitoring Instrument	=	\$ 6,500
Compressors		
(2 compressors) * (\$6,240/compressor)	=	\$ 12,480
Rupture Disk Assemblies:		
- Disks		
(45 PRVs) * (\$78/PRV)	=	\$ 3,510

- Holders, valves, etc.		
(45 PRVs) * (\$3,852/PRV)	=	\$173,340
Sample Connections		
(35 sample con.) * (\$408/sample con.)	=	<u>\$ 14,280</u>
Total Capital Cost		\$210,100

Initial Monitoring and Leak Repair

Monitoring

Gas/Vapor Valves:		
(414 valves) * (2.50/component)	=	\$ 1,035
Light Liquid Valves:		
(1,179 valves) * (2.50/component)	=	\$ 2,948
Light Liquid Pumps:		
(40 pumps) * (\$2.50/component)	=	\$ 100
Connectors:		
(2,662 connectors) (\$2.50/component)	=	\$ 6,655

Initial Repair, Leaks Found:

Gas/Vapor Valves:		
(414 valves) * (0.114 leaks/valve)	=	47.2 leaks
Light Liquid Valves:		
(1,179 valves) * (0.065 leaks/valve)	=	76.6 leaks
Light Liquid Pumps:		
(40 pumps) * (0.20 leaks/pump)	=	8.0 leaks
Connectors:		
(2,662 connectors) *		
(0.021 leaks/connectors)	=	55.9 leaks

Leaks Requiring Repair:

Gas/Vapor Valves: (47.2 leaks) * (0.25)	=	11.8 leaks
Light Liquid Valves: (76.6 leaks) * (0.25)	=	19.2 leaks
Light Liquid Pumps: (8.0 leaks) * (0.75)	=	6.0 leaks
Connectors: (55.9 leaks) * (0.25)	=	14.0 leaks

Repair Costs:

Gas/Vapor Valves:		
(11.8 leaks) * (4 hrs/leak) * (\$22.5/hr)	=	\$ 1,062
Light Liquid Valves:		
(19.2 leaks) * (4 hrs/leak) * (\$22.5/hr)	=	\$ 1,728
Light Liquid Pumps:		
(6.0 leaks) * (16 hrs/leak) * (\$22.5/hr)	=	\$ 2,160
Connectors:		
(14.0 leaks) * (2 hrs/leak) * (\$22.5/hr)	=	\$ 630

Total Initial LDAR Costs

Gas/Vapor Valves:	\$1,035 + \$1,062	=	\$ 2,097
Light Liquid Valves:	\$2,948 + \$1,728	=	\$ 4,676

Light Liquid Pumps:	\$ 100 + \$2,160	=	\$ 2,260
Connectors:	\$6,655 + \$ 630	=	\$ 7,285
			\$16,300

Annual Leak Detection and Repair

Monitoring Costs:

Gas/Vapor Valves:			
(414 valves) * (12 times/yr) *			
(\$2/time/valve)	=	\$ 9,936	
Light Liquid Valves:			
(1,179 valves) * (12 times/yr) *			
(\$2/time/valve)	=	\$28,296	
Light Liquid Pumps:			
(40 valves) * (12 times/yr) *			
(\$2/time/valve)	=	\$ 960	
Connectors:			
(2,662 valves) * (1 time/yr) *			
(\$2/time/concentration)	=	\$ 5,324	
Pressure Relief Leak:			
(45 PRVs) * (1 time/yr) * (\$2/time/PRV)	=	\$ 90	
Pumps (Visual):			
(40 pumps) * (52 times/yr) * (0.5 min/time)			
(1 hr/60 min) * (\$22.5/hr)	=	\$ 390	
			\$45,000

Annual Leaks Found:

Gas/Vapor Valves:			
(414 valves) * (12 times/yr) *			
(0.02 leaks/valve)	=	99.4 leaks	
Light Liquid Valves:			
(1,179 valves) * (12 times/yr) *			
(0.02 leaks/valve)	=	283.0 leaks	
Light Liquid Pumps:			
(40 pumps) * (12 times/yr) *			
(0.10 leaks/pump)	=	48.0 leaks	
Connectors:			
(2,662 connectors) * (1 time/yr) *			
(0.005 leaks/connectors)	=	13.3 leaks	

Leaks Requiring Repair:

Gas/Vapor Valves:	(99.4 leaks) * (0.25)	=	24.9 leaks
Light Liquid Valves:	(283.0 leaks) * (0.25)	=	70.8 leaks
Light Liquid Pumps:	(48.0 leaks) * (0.75)	=	36.0 leaks
Connectors:	(13.3 leaks) * (0.25)	=	3.3 leaks

Cost of Repairs:

Gas/Vapor Valves:

(24.9 leaks) * (4 hrs/leak) * (\$22.5/hr) = \$ 2,241

Light Liquid Valves:

(70.8 leaks) * (4 hrs/leak) * (\$22.5/hr) = \$ 6,372

Light Liquid Pumps:

(36.0 leaks) * (16 hrs/leak) * (\$22.5/hr) = \$12,960

Connectors:

(3.3 leaks) * (2 hrs/leak) * (\$22.5/hr) = \$ 149

\$21,700

Annualized Capital Charges

Monitoring Instrument: (0.23) * (\$6,500) = \$ 1,495

Compressors: (0.163) * (\$12,480) = \$ 2,034

Rupture Disk Assemblies:

-Disks: (0.58) * (\$3,510) = \$ 2,036

-Holders, Valves, etc: (0.163) * (\$173,340) = \$28,254

Sample Connections: (0.163) * (\$14,280) = \$ 2,328

Initial LDAR: (0.163) * (\$16,300) * (1.4) = \$ 3,720

Pump Seals: (6.0) * (0.58) * (\$180) = \$ 626

\$40,500

Annual Maintenance Costs

Monitoring Instrument: = \$ 4,280

Compressors: (\$12,480) * (0.05) = \$ 624

Rupture Disk Assemblies:

-Disks: (\$3,510) * (0.05) = \$ 176

-Holders, Valves, etc: (\$173,340) * (0.05) = \$ 8,667

Open-Ended Lines:^a (\$14,382) * (0.05) = \$ 719

Sample Connections: (\$14,280) * (0.05) = \$ 714

Replacement Pump Seals:

(36.0 leaks) * (\$180/seal) = \$ 6,480

\$21,700

Miscellaneous Annual Charges

Monitoring Instrument: (\$6,500) * (0.04) = \$ 260

Compressors: (\$12,480) * (0.04) = \$ 499

^aIf open-ended lines were uncontrolled, the capital cost to control them would be 141 times \$102 which equals \$14,280. As described in Section F.6, it is assumed that open-ended lines are controlled at baseline, and the value for capital cost for open-ended lines is calculated solely for the purpose of estimating annual maintenance and miscellaneous charges. This value for capital cost is not included in the total capital cost for the process unit (\$210,100).

Rupture Disk Assemblies:		
-Disks: (\$3,510) * (0.04)	=	\$ 140
-Holders, Valves, etc: (\$173,340) * (0.04)	=	\$ 6,934
Pressure Relief Devices	=	
Open-Ended Lines: ^a (\$14,382) * (0.04)	=	\$ 575
Sample Connections: (\$14,280) * (0.04)	=	\$ 571
Replacement Pump Seals: (6,480) * (0.8)	=	\$ 5,184
		\$ 14,200

Labor Charges

Monitoring Cost:	=	\$ 45,000
Repair Costs:	=	\$ 21,700
Administration and Support:		
(0.4) * (\$45,000 + \$21,700)	=	\$ 26,700
		\$ 93,400

Total Annual Costs

Annualized Capital Charges	=	\$ 40,500
Annual Operating Cost: (\$21,700 + \$14,200)	=	\$ 35,900
Annual Labor Costs	=	\$ 93,400
		\$169,800

The following calculations are for a baseline controlled process unit having identical equipment counts to those used in the previous example calculation.

Capital Costs

Compressors:		
(2 compressors) * (\$6,240/compressor)	=	\$ 12,480
Rupture Disk Assemblies:		
-Disks: (45 PRVs) * (78/PRV)	=	\$ 3,510
-Holders, Valves, etc: (45 PRVs) * (\$3,852 PRV)	=	\$173,340

^aIf open-ended lines were uncontrolled, the capital cost to control them would be 141 times \$102 which equals \$14,280. As described in Section F.6, it is assumed that open-ended lines are controlled at baseline, and the value for capital cost for open-ended lines is calculated solely for the purpose of estimating annual maintenance and miscellaneous charges. This value for capital cost is not included in the total capital cost for the process unit (\$210,100).

Sample Connections:

(35 sample connectors) *
(\$408/sample connector) = \$ 14,280

\$203,600

Annual Leak Detection and Repair

The annual leak detection and repair costs for a baseline controlled unit are exactly the same as for a baseline uncontrolled unit. Total monitoring cost is \$45,000, repair cost \$21,700, and administration and support cost \$26,700.

Annualized Capital Charges

Compressors: (0.163) * (\$12,480) = \$ 2,034

Rupture Disk Assemblies:

-Disks: (0.58) * (\$3,510) = \$ 2,036

-Holders, Valves, etc: (0.163) * (\$173,340) = \$28,254

Sample Connections: (0.163) * (\$14,280) = \$ 2,328

\$34,700

Annual maintenance and miscellaneous costs for a baseline controlled process unit are the same as for a baseline uncontrolled unit. Annual maintenance costs are \$21,700 and miscellaneous costs are \$14,200.

Labor Charge Summary:

Monitoring Cost: = \$ 45,000

Repair Costs: = \$ 21,700

Administration and Support: = \$ 26,700

\$ 93,400

Total Annual Costs

Annualized Capital Charges = \$ 34,700

Annual Operating Cost: = \$ 35,900

Annual Labor Costs = \$ 93,400

\$164,000