United States
Environmental Protection
Agency

Air Pollution Training Institute
MD 20
Environmental Research Center
Research Triangle Park, NC 27711

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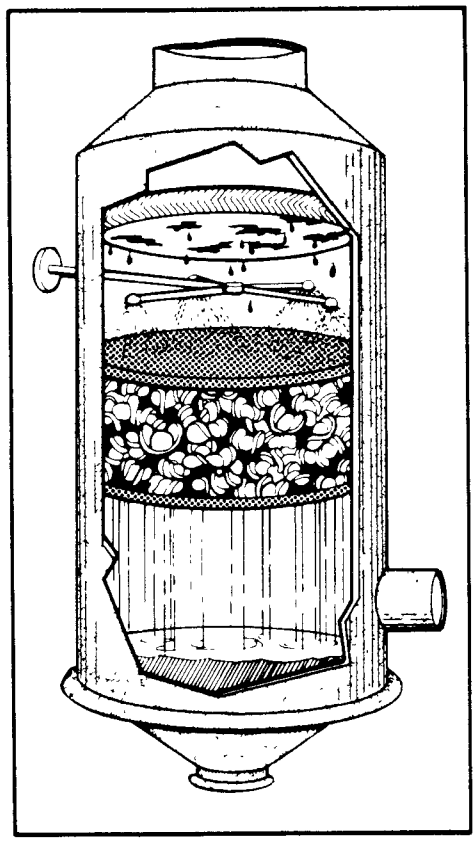
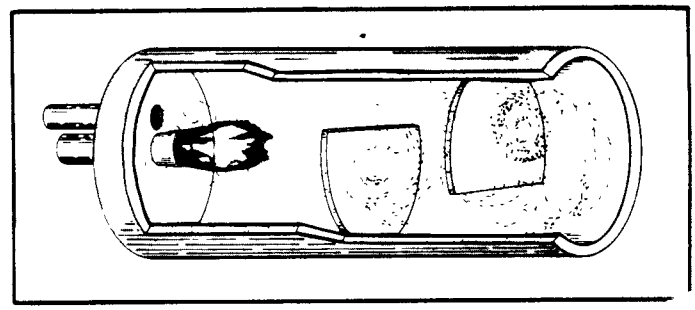
Air

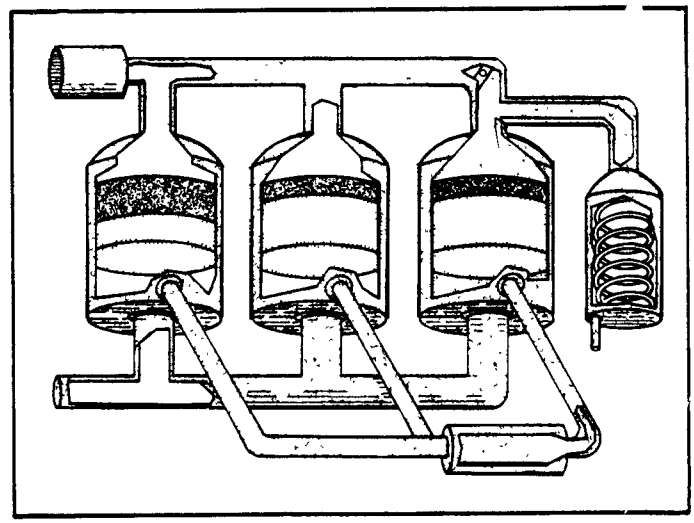
EPA

APTI Course 415 Control of Gaseous Emissions

Student Manual







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Nomenclature

Symbols, metric units (English units)

a	surface area available for absorption per unit volume of tower, m^2/m^3 (ft^2/ft^3)
A	area, m^2 (ft^2)
A_c	area of rising column of a gaseous emission, m^2 (ft^2)
A_s	top surface area of a hot emission source, m^2 (ft^2)
AHP	air horsepower, kW (hp)
BHP	brake horsepower, kW (hp)
c	concentration, g/cm^3 (lb/ft^3)
c_{AI}	concentration of component A at the gas-liquid interface, g/cm^3 (lb/ft^3)
c_{AL}	concentration of component A in the liquid phase, g/cm^3 (lb/ft^3)
C	empirical constant in BET equation, dimensionless
C_B	breakthrough capacity, weight %
C_e	coefficient of air entering an orifice, dimensionless
C_p	specific heat, J/kg•°C (Btu/lb•°F)
$\bar{C}_p$	average specific heat, J/kg•°C (Btu/lb•°F)
C_s	saturation capacity of adsorber, weight %
d	diameter of a pipe or duct, m (ft)
D	adsorber bed depth, m (ft)
d_t	diameter of a tower or column, m (ft)
d_w	diameter of a fan wheel, m (ft)
D_c	diameter of a column of hot rising fume, m (ft)
D_h	diameter of a hood, m (ft)
D_s	diameter of a hot emissions source, m (ft)
E	heat transfer effectiveness, %
$\mathcal{E}$	mechanical efficiency of a fan, %
f	percent of flooding
F	absorber packing factor, m^2/m^3 (ft^2/ft^3)
ΔF	free energy change, J/mol (Btu/mol)
g_c	gravitational constant, $kg \cdot m/N \cdot s^2$ ($lb \cdot ft^2/lb_f \cdot sec$)
G	gas flow rate, kg/s (lb/min)
G′	superficial gas flow rate, $kg/s \cdot m^2$ ($lb/min \cdot ft^2$)
G_m	gas molar flow rate, kg mol/s (lb mol/sec)
	Henry's Law constant, kPa/mol fraction (atm/mol fraction)
H	enthalpy, J/g (Btu/lb)
H_c	heat of combustion, J/m^3 of fuel (Btu/ft^3 of fuel)
H_A	available heat, J/m^3 (Btu/ft^3)
H_{OG}	height of a transfer unit based on overall gas phase, m (ft)
H_{OL}	height of a transfer unit based on overall liquid phase, m (ft)
h_p	perimeter of canopy hood, m (ft)
H_s	sensible heat, J/g (Btu/lb)

H_v	latent heat of vaporization, J/g (Btu/lb)
HV_G	gross heating value, J/m³ (Btu/ft³)
HV_N	net heating value, J/m³ (Btu/ft³)
k_g	individual mass transfer coefficient based on gas phase, kg mol/h•m²•kPa (lb mol/hr•ft²•atm)
k_ℓ	individual mass transfer coefficient based on liquid phase, kg mol/h•m²•kPa (lb mol/hr•ft²•atm)
k_s	individual mass transfer coefficient for combined surface migration and pore diffusion, kg mol/h•m²•kPa (lb mol/hr•ft²•atm)
K_{OG}	overall mass transfer coefficient based on gas phase, kg mol/h•m²•kPa (lb mol/hr•ft²•atm)
K_{OL}	overall mass transfer coefficient based on liquid phase, kg mol/h•m²•kPa (lb mol/hr•ft²•atm)
l	length, m (ft)
L	liquid flow rate, kg/s (lb/sec)
L'	superficial liquid flow rate, kg/s•m² (lb/sec•ft²)
L_m	liquid molar flow rate, kg mol/s (lb/sec)
m	slope of a line
$\dot{m}$	mass flow rate, kg/h (lbs/hr)
M	mass, kg (lb)
M_{at}	mass of theoretical air per unit mass of fuel combusted, kg/kg (lb/lb)
MW	molecular weight, g mol (lb mol)
n	number of moles
N_A	mass flux, kg mol/s•m² (lb mol/sec•ft²)
N_{OG}	number of transfer units based on overall gas coefficient
N_{OL}	number of transfer units based on overall liquid coefficient
N_p	number of theoretical plates
p	partial pressure, kPa (psi)
p^*	partial pressure at equilibrium, kPa (psi)
p°	vapor pressure, kPa (psi)
p_g	gage pressure, kPa (psig)
p_s	static pressure, kPa (psi)
p_v	velocity pressure, kPa (psi)
p_{AI}	partial pressure of component A at the gas-liquid interface, kPa (psi)
p_{AG}	partial pressure of component A in the gas phase, kPa (psi)
P_T	total pressure, kPa (psi)
P_b	barometric or atmospheric pressure, kPa (psi)
P	absolute pressure, kPa (psi)
$\mathscr{P}$	power, kW (hp)
Q	volumetric flow rate, m³/min (cfm)
q	heat rate, J/h (Btu/hr)
r	radius, m (ft)
rpm	fan speed (revolutions per min)

R	ideal gas constant (see page 2-8)
Re	Reynolds Number, dimensionless
T	temperature, °C (°F)
ΔT_{lm}	log mean temperature difference, °C (°F)
ΔT_m	mean temperature difference, °C (°F)
U	heat transfer coefficient, kJ/h/m² (Btu/hr/ft²)
v	velocity, m/s (ft/sec)
V	volume, m³ (ft³)
V_{at}	volume of theoretical air required to combust a unit volume of fuel, m³/m³ (ft³/ft³)
w	width, m (ft)
x	mole fraction of contaminant in liquid phase
x	distance, m (ft)
X	mole fraction of contaminant in pure liquid
X_c	sum of the distance from the hot source to the hood face plus the distance from the top of the heat source to a hypothetical vortex of the fume column, m (ft)
X_s	degree of saturation in the mass transfer zone, percent
y	mole fraction of contaminant in gas phase
Y	mole fraction of contaminant in carrier gas
Y_c	distance from hood face to hot emissions source, m (ft)
Z	height of packing in a tower, m (ft)
Z_c	distance from top of heat source to hypothetical vortex of fume column, m (ft)

Greek letters

ϱ	density, kg/m³ (lb/ft³)
Υ	shearing stress, kPa/m² (lb/ft²)
ν	kinematic viscosity, m²/s (ft²/sec)
μ	viscosity, centipoise, Pa•s (centipoise)
Θ	residence time, s (sec)
ϕ	ratio of specific gravity of water to solvent
ϵ	flooding correlation ordinate, dimensionless
β	void volume between adsorbent granules m³/m³ (ft³/ft³)
ψ	empirical correlation for tray tower diameter, $m^{0.25} \cdot h^{0.25}/kg^{0.25}$ ($ft^{0.25} \cdot hr^{0.25}/lb^{0.25}$)

Chapter 1

Air Pollution Control

Overview

Controlling the emission of pollutants from industrial sources is important in protecting the quality of the air. Air pollutants can exist in the form of particulate matter or as gases. This text will focus on controlling gaseous air pollutants. National Ambient Air Quality Standards (NAAQS) currently exist for five gaseous pollutants. Table 1-1 lists the primary standards; note that these are *ambient* standards and not *source emission limits*.

Table 1-1. National primary ambient air quality standards for gaseous pollutants.

Pollutant	Standards	
	Annual: arithmetic mean	Other
Sulfur dioxide, SO_2	80 $\mu g/m^3$ (0.03 ppm)	365 $\mu g/m^3$ (0.14 ppm) 24 hr max. once a yr.
Carbon monoxide, CO	—	40,000 $\mu g/m^3$ (35 ppm) 1 hr max. once a yr. 10,000 $\mu g/m^3$ (9 ppm) 8 hr max. once a yr.
Ozone, O_3	—	235 $\mu g/m^3$ (0.12 ppm) 1 hr max. once a yr.
Nonmethane hydrocarbons, NMHC	—	160 $\mu g/m^3$ (0.24 ppm) 3 hr max. (6 to 9 a.m.) once a yr.
Nitrogen dioxide, NO_2	100 $\mu g/m^3$ (0.05 ppm)	—

Air cleaning devices have been reducing gaseous emissions from various industrial sources for many years. Originally, air cleaning equipment was used only if the gaseous contaminant was highly toxic or had some recovery value. Now with legislation such as the 1970 Clean Air Act and the 1977 Clean Air Act Amendments, control technologies have been upgraded and more industrial sources are regulated in order to meet the NAAQS. In addition, State and local air pollution agencies have adopted regulations that are in some cases more stringent than Federal emission standards.

The Clean Air Act as amended in 1977 contains basically three classes of control technology which are aimed at reducing emissions from existing sources, new sources, and modified sources. These levels of control are: Reasonably Available Control Technology (RACT), Best Available Control Technology (BACT), and controls that reflect the Lowest Achievable Emission Rate (LAER).

In order to understand the distinctions between these various types of control technologies it is necessary to become familiar with their regulatory definitions.

- First consider the following definitions from the U.S. Code of Federal Regulations:

A *new source* is one which is contracted and installed at a facility after the date emission standards are proposed for that industry.

Modification means any physical change or change in the operation of an existing facility which increases the amount of an air pollutant emitted, or results in the emission of an air pollutant not previously emitted into the atmosphere to which a standard applies.

The term *existing source* refers to an air pollution source which was constructed before the proposal date of the emission standard.

- On this basis, then:

Reasonably Available Control Technology is control technology applied to existing sources to meet ambient air quality standards. This technology is generally designated in the *State Implementation Plan* (SIP) so as to reduce emissions to achieve ambient standards. RACT considers both the cost and technology available for emission control. This control is usually less stringent than both BACT or LAER.

Best Available Control Technology is control technology applied to new and modified sources. *Prevention of Significant Deterioration* (PSD) is the review process for sources in attainment areas and is required for each pollutant subject to regulation under the Act. An *attainment area* is an air quality control region that meets the National Ambient Air Quality Standards.

New Source Performance Standards (NSPS) are set for a number of source categories. Best Available Control Technology must reduce emissions at least as much as the NSPS limit. New Source Performance Standards that have been promulgated are published first in the Federal Register and then in the U.S. Code of Federal Regulations. BACT applied to a source must be technologically feasible and must also reflect considerations of cost and energy usage. The BACT requirement is to be applied on a case-by-case basis by EPA regional engineers.

Lowest Achievable Emission Rate refers to control technology applied to new and modified sources in a nonattainment area. A nonattainment area is an area or region where the National Ambient Air Quality Standard for a particular pollutant is being violated. Note: A nonattainment area might be nonattainment in one pollutant, but attainment in another. LAER is the most stringent emission limitation which is contained in the SIP of any State or the most stringent emission limitation achieved in practice by that source category. LAER may be in some cases considered to be **technology forcing**, involving the transfer of technology from one source category to another. LAER is more stringent than BACT or RACT and **must** be at least as stringent as the control specified by the NSPS.

Control equipment which satisfies the requirements of RACT, BACT or LAER is generally specific to a particular source category. For a given source category one control device may be able to satisfy all three levels of control, or three different control devices may be needed. Once the appropriate gaseous emission control device is identified (absorber, adsorber, or thermal incinerator), its operation usually can be modified to meet the required level of control. For example, a large utility may choose to meet the RACT requirements for SO_2 emissions by burning low sulfur coal. BACT or LAER requires scrubbing the flue gas. The same scrubber system (i.e. venturi, spray, or plate) may be used to meet the appropriate control level by improving its operating efficiency or by adding stages to it.

As illustrated in the above example, gaseous emission control can also be achieved by methods other than add-on control devices. Changing fuel sources, modifying or changing raw materials, or using alternative production procedures are also used to reduce emissions without adding control equipment.

This manual is intended to be used in APTI Course 415 Control of Gaseous Emissions, and it presents the fundamental concepts of the operation of gaseous emission control equipment for stationary sources. Control techniques applied to reducing the emissions of gaseous pollutants are *absorption, adsorption, combustion,* and *condensation*. Controlling SO_2 and NO_x emissions from fossil-fuel-fired steam generators is also discussed. These are the largest stationary sources emitting SO_2 and NO_x pollutants. Sources and control of volatile organic compounds are not included in this manual since this topic is covered in APTI Course 482 Sources and Control of Volatile Organic Compounds. Also, a chapter describing the auxiliary equipment, such as hooding, ductwork, and fans associated with the control equipment is included.

Gaseous Emission Control Techniques

Gaseous pollutants are emitted from a variety of processes. Gaseous pollutants which must be controlled are not limited to those listed in Table 1-1. Most noxious or toxic gases are also controlled. Table 1-2 lists some typical gaseous pollutants and their sources (Parekh, 1975).

Table 1-2. Typical gaseous pollutants and their sources.

Key element	Pollutant	Source
S	SO_2	Boiler flue gas
	SO_3	Sulfuric acid manufacturing
	H_2SO_4 vapors	Sulfuric acid manufacturing, pickling operations
	H_2S	Natural gas processing, pulp and paper mills, sewage treatment
	R-SH (mercaptans)	Petroleum refining, pulp and paper mills
N	NO, NO_2	Nitric acid manufacturing, boiler flue gas
	HNO_3 vapors	Nitric acid manufacturing, pickling operations
	NH_3	Ammonia manufacturing
	Other N compounds (i.e. amines, pyridines)	Sewage, rendering, solvent processes
C	Inorganic CO	Incomplete combustion
	Organics: Volatile Organic Compounds Hydrocarbons Paraffins Olefins Aromatics	Solvent uses, gasoline marketing, petrochemical plants
	Oxygenated hydrocarbons Aldehydes Ketones Alcohols Phenols	Surface coating operations, petroleum processing, plastics manufacturing
	Chlorinated solvents	Dry cleaning, degreasing
Halogen F	HF	Phosphate fertilizer plant, aluminum plant
	SiF_4	Ceramics, fertilizer plant
Cl	HCl	HCl manufacturing, PVC combustion
	Cl_2	Chlorine manufacturing

Techniques used to control gaseous emissions are absorption, adsorption, combustion, and condensation. The applicability of a given technique depends on the physical and chemical properties of the pollutant and the exhaust stream. More than one technique may be capable of controlling emissions from a given source. For example, vapors generated from loading gasoline into tank trucks at large bulk terminals are controlled by using any of these four techniques. Most often, however, one control technique is used more frequently than others for a given source-pollutant combination. For example, absorption is commonly used to remove SO_2 from boiler flue gas.

Absorption

Absorption is a mass transfer operation in which a gas is dissolved in a liquid. A contaminant (pollutant) exhaust stream contacts a liquid, and the contaminant diffuses from the gas phase into the liquid phase. The absorption rate is enhanced by high diffusion rates of the contaminant in both the liquid and gas phase, high solubility of the contaminant, large liquid-gas contact area and, good mixing between liquid and gas phases (turbulence).

The liquid most often used for absorption is water since it is inexpensive, readily available, and can dissolve a number of contaminants. Reagents can be added to the absorbing water to increase the removal efficiency of the system. Certain reagents merely increase the solubility of the contaminant in the water. Other reagents **chemically react** with the contaminant after it has been absorbed (such as a lime slurry used to scrub SO_2 flue gases). In reactive scrubbing the absorption rate is much higher, so in some cases a smaller, more economical system can be used. However, the reactions can form a precipitate which could cause plugging problems in the absorber or associated equipment.

Gas absorbers or wet scrubbers are designed to provide good mixing of the gas and liquid phases. The devices used for gas absorption are often the same as those used in particulate emission scrubbing. These include packed towers, plate towers, spray columns, and venturi scrubbers. Although these devices are identical, a wet particulate emission scrubber is not operated the same way as a gas absorber is operated. In a wet scrubber, particulate matter removal efficiency is a function of the pressure drop—the higher the pressure drop the greater the removal efficiency. For gas absorption, a higher pressure drop results in shorter contact time between the phases which limit the amount of absorption that can occur. Therefore, **optimizing** both gas and particulate pollutant removal in one device is difficult.

If a gaseous contaminant is very soluble, almost any of the wet scrubbers will adequately remove this contaminant. However, if the contaminant is of low solubility, the packed tower or the plate tower is more effective. Both of these devices provide long contact time between phases and have relatively low pressure drops. The packed tower, the most common gas absorption device, consists of an empty shell filled with packing. Liquid flows down over the packing exposing a large film area to the gas flowing up through the packing. Plate towers consist of horizontal plates placed inside the tower. Gas passes up through orifices in these plates, while liquid flows down across the plate providing the desired contact. The major disadvantage in using these devices is that they both are susceptible to plugging when the exhaust stream contains high concentrations of particulate matter. When an exhaust stream contains a high level of particulate matter, venturis, spray devices, or combinations of systems can be used.

Adsorption

Adsorption is a mass transfer process that involves removing a gaseous contaminant by adhering it to the surface of a solid. Adsorption can be classified as *physical* or *chemical*. In physical adsorption, a gas molecule adheres to the surface of the solid

due to an imbalance of natural forces (electron distribution). In chemisorption, once the gas molecule adheres to the surface, it **reacts chemically** with it. The major distinction is that physical adsorption is readily reversible while chemisorption is not.

All solids physically adsorb gases to some extent. Certain solids, called *adsorbents* have a high attraction for specific gases plus a large surface area which gives them a large capacity. By far the most important adsorbent for air pollution control is activated carbon. Due to its surface, activated carbon will preferentially adsorb hydrocarbon vapors and odorous organic compounds from an air stream. Most other adsorbents (molecular sieves, silica gel, and activated aluminas) will preferentially adsorb water vapor thereby rendering them useless to remove other contaminants.

For activated carbon, the amount of hydrocarbon vapors that can be adsorbed depends on the physical and chemical characteristics of the vapors, their concentration in the gas stream, system temperature, system pressure, humidity of the gas stream, and molecular weight of the vapor. Physical adsorption is a reversible process; the adsorbed vapors can be released by increasing the temperature, decreasing the pressure, or using a combination of both. Vapors are normally desorbed by heating the adsorber with steam.

Adsorption can be a very useful removal technique, since it is capable of removing very small quantities (a few ppm) of vapor from an air stream. The vapors are not destroyed, only stored on the adsorbent surface until they can be removed by desorption. The desorbed vapor stream is highly concentrated. Depending on the nature of this material, it can be condensed and recycled or burned as the ultimate disposal technique.

The most common adsorption system is the fixed bed adsorber. These systems consist of two or more adsorber beds operating on a timed adsorbing and desorbing cycle. One or more beds are adsorbing vapors, while the other bed is being regenerated. If particulate matter or liquid droplets are present in the vapor-laden air stream, this stream is sent to pretreatment to remove them. If the temperature of the inlet vapor stream is high (much above 120°F), cooling may also be provided. Also, since all adsorption processes are *exothermic*, cooling coils in the carbon bed itself may be needed to prevent excessive heat buildup. Carbon bed depth is usually limited to a maximum of four feet and vapor velocity through the adsorber is held below 100 fpm to prevent an excessive pressure drop.

Combustion

Combustion is defined as rapid, high temperature gas phase oxidation. Simply, the contaminant (a carbon-hydrogen substance) is burned with air and converted to carbon dioxide and water vapor. The operation of any combustion source is governed by the three T's of combustion: *temperature, turbulence*, and *time*. For complete combustion to occur, each contaminant molecule must come in contact with oxygen (turbulence) at a sufficient temperature, and be held at this temperature for an adequate time. These three variables are totally dependent on each other. For example, if a higher temperature were used, less mixing of the

contaminant and combustion air or shorter residence time may be used. If adequate turbulence cannot be provided, a higher temperature or longer residence time may be required for complete combustion.

Combustion devices can be categorized as *flares, thermal incinerators,* or *catalytic incinerators*. Flares are direct combustion devices used to dispose of large quantities or emergency releases of combustible gases. Flares are normally elevated (from 100 to 400 ft) to protect the surroundings from heat and flames. Flares are often designed for steam injection at the flare tip. The steam provides sufficient turbulence to ensure complete combustion which prevents smoking. Flares are very noisy which can cause problems for adjacent neighborhoods.

Thermal incinerators are also called afterburners, direct flame incinerators, or thermal oxidizers. These are devices in which the contaminant air stream passes around or through a burner and into a refractory-lined residence chamber where oxidation occurs. To ensure complete combustion of the contaminant, thermal incinerators are designed to operate between a temperature of 700 to 800°C (1300 to 1500°F) and a residence time of 0.3 to 0.5 seconds. Ideally, as much of the fuel value as possible is supplied by the waste contaminant stream. This reduces the amount of auxiliary fuel needed to maintain the proper temperature.

In catalytic incineration the contaminant-laden stream is heated and passed through a catalyst bed which promotes the oxidation reaction at a lower temperature. Catalytic incinerators normally operate between 370 and 480°C (700 and 900°F). This reduced temperature represents a continuous fuel savings. However, this may be offset by the cost of the catalyst. The catalyst, which is usually platinum, is coated on a cheaper metal or ceramic support base. The support can then be arranged to expose a high surface area, which provides sufficient active sites on which the reactions occur. Catalysts are subject to both physical and chemical deterioration. Halogens and sulfur containing compounds act as catalyst suppressants and decrease the catalysts' usefulness. Certain heavy metals such as mercury, arsenic, phosphorous, lead and zinc are particularly poisonous.

Condensation

Condensation is a process in which the volatile gases are removed from the contaminant stream and changed to a liquid. Condensation is usually achieved by reducing the temperature of a vapor mixture until the partial pressure of the condensable component equals its vapor pressure. Condensation requires low temperatures to liquify most pure contaminant vapors. Condensation is affected by the composition of the contaminant gas stream. The presence of additional gases which do not condense at the same conditions, such as air, hinders condensation.

Condensers are normally used in combination with primary control devices. Condensers can be located upstream of (before) an incinerator, adsorber or absorber. These condensers reduce the volume of vapors the more expensive equipment must handle. Therefore, the size and the cost of the primary control device can be reduced. Similarly, condensers can be used to remove water vapors from a process stream with a high moisture content upstream of a control system. A prime example is the use of condensers in rendering plants to remove moisture from the cooker

exhaust gas. When used alone, refrigeration is required to achieve the low temperatures required for condensation. Refrigeration units are used to successfully control gasoline vapors at large bulk gasoline dispensing terminals.

Condensers are classified as being either *contact* condensers or *surface* condensers. Contact condensers cool the vapor by spraying liquid directly on the vapor stream. These devices resemble a simple spray scrubber. Surface condensers are normally *shell-and-tube heat exchangers.* Coolant flows through the tubes, while the vapor is passed over and condenses on the outside of the tubes. In general, contact condensers are more flexible, simpler and less expensive than surface condensers. However, surface condensers require much less water and produce nearly 20 times less wastewater that must be treated than contact condensers. Surface condensers also have an advantage in that they can directly recover valuable contaminant vapors.

This chapter presented a brief overview of the regulations which specify a required level of emission reduction. The four methods used to control gaseous emissions were summarized. These methods will be discussed in detail in later chapters in this manual. The chapters on the control methods have been developed in three steps. First, the basic theory of the control method is presented, then a section discusses procedures used to estimate size and operating conditions of the control equipment, and finally, the types and operation of the equipment are described. The equations presented in each chapter are used to estimate important design variables for that particular control device. They are intended to give practical operating values, not rigid design values. The following chapter reviews some of the basic concepts of the behavior of gases. These basic concepts are important for understanding the operation of gaseous emission control equipment.

Reference

Parekh, R. 1975. Equipment for controlling gaseous pollutants. *Chem. Engr.* 82:129 (October 6).

Chapter 2
Basic Concepts of Gases

In order to properly evaluate gaseous control devices, a thorough understanding of the process variables that affect a gas stream is essential. This chapter briefly reviews a few basic concepts of gas behavior. The chapter is divided into two parts. The first half defines important physical and chemical properties of gases: temperature, pressure, density, molecular weight, viscosity and specific heat. Common symbols, units of measurement, and any important values are presented. The second half of this chapter reviews a few scientific laws used to predict the behavior of a gas under varying process conditions including the ideal gas law, Raoult's Law, Dalton's Law and Henry's Law.

The system of measurement used in this manual is the International System of Units (SI). This system is an updated, more restrictive version of the metric system which derives its base units from physical phenomena and scientific relationships. In this manual the SI units are normally listed first with the English units given in parenthesis. Occasionally the units in parenthesis are the commonly used units instead of English units. For some tables, figures, and equations, conversion from English to SI was not practical. For these instances only the English units are presented.

Expression of Gas Temperature

Fahrenheit and Celsius Scales

The range of units on the Fahrenheit scale between freezing and boiling is 180; on the Celsius or Centigrade scale, the range is 100. Therefore, each Celsius-degree is equal to 9/5 or 1.8 Fahrenheit-degree. The following relationships convert one scale to the other:

(Eq. 2-1) $$°F = 1.8°C + 32$$

(Eq. 2-2) $$°C = (°F - 32)/1.8$$

Where: °F = degrees Fahrenheit
°C = degrees Celsius or degrees Centigrade

Absolute Temperature

Experiments with perfect gases have shown that, under constant pressure, for each change in Fahrenheit-degree below 32°F the volume of gas changes 1/491.67. Similarly, for each Celsius-degree, the volume changes 1/273.16. Therefore, if this change in volume per temperature-degree is constant, the volume of gas would, theoretically, become zero at 491.67 Fahrenheit-degrees below 32°F or at a reading

of −459.67 °F. On the Celsius or Centigrade scale, this condition occurs at 273.16 Celsius-degrees below 0 °C, or at a temperature of −273.16 °C. Relationships of the various temperatures are shown in Figure 2-1.

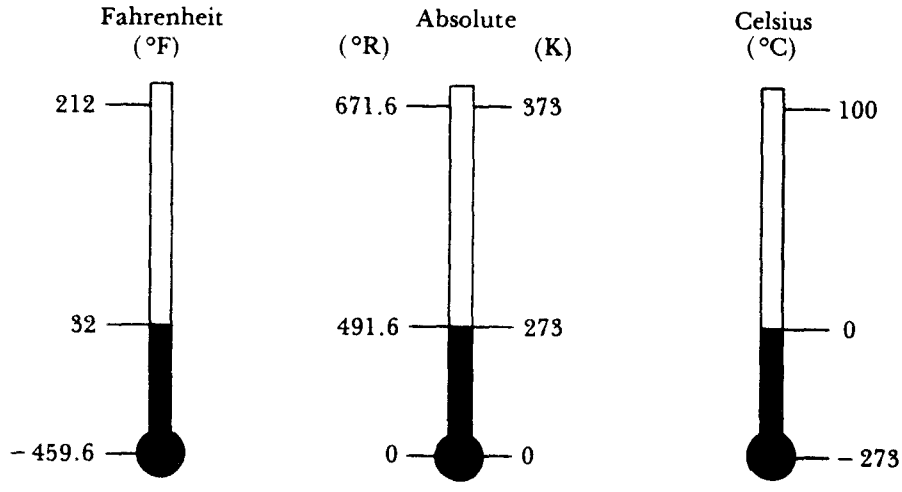


Figure 2-1. Comparison of Fahrenheit, Celsius, and absolute temperature scales.

Absolute temperatures determined by using Fahrenheit units are expressed as degrees Rankine (°R); those determined by using Celsius units are expressed as degrees Kelvin (K). The following relationships convert one scale to the other:

(Eq. 2-3) $$°R = °F + 459.67$$

(Eq. 2-4) $$K = °C + 273.16$$

In the International System of Units (SI) the degree mark (°) is not used with the units in the Kelvin scale. Throughout this manual degrees Celsius will be °C, degrees Kelvin will be K.

Expression of Gas Pressure

Definition of Pressure

A body may be subjected to three kinds of stress: shear, compression, and tension. Fluids are unable to withstand tensile stress; hence, they are subject to shear and compression only.

Unit compressive stress in a fluid is termed pressure and is expressed as force per unit area (e.g. N/m^2 or $Newton/meter^2$, lb/in^2 or psi). For the SI system of units, 1 $Newton/meter^2$ is defined as a Pascal (Pa).

Pressure is equal in all directions at a point within a volume of fluid, and acts perpendicular to a surface.

Barometric Pressure

Barometric pressure and atmospheric pressure are synonymous. These pressures are measured with a barometer and are usually expressed as inches, or millimeters, of mercury (Hg). Standard barometric pressure is the average atmospheric pressure at sea level, 45° north latitude at 35°F. It is equivalent to a pressure of 14.696 pound-force per square inch exerted at the base of a column of mercury 29.92 inches high. In the SI system this pressure is equal to 101.325 kPa. In the metric system, the average atmospheric pressure is 760 mm Hg or 1 atm. Weather and altitude are responsible for barometric pressure variations.

Gage Pressure

Measurements of pressure by ordinary gages are indications of difference in pressure above, or below, that of the atmosphere surrounding the gage. Gage pressure, then, is ordinarily the pressure of the system. If greater than the pressure prevailing in the atmosphere, the gage pressure is expressed as a positive value; if smaller, the gage pressure is expressed as negative. The term *vacuum* designates a negative gage pressure.

The abbrevation *g* (in p_g) is used to specify a gage pressure. For example, psig means pound-force per square inch gage pressure.

Absolute Pressure

Because gage pressure (which may be either positive or negative) is the pressure relative to the prevailing atmospheric pressure, the gage pressure, added algebraically to the prevailing atmospheric pressure (which is always positive), provides a value that has a datum of *absolute zero pressure*. A pressure calculated in this manner is called absolute pressure. The mathematical expression is:

(Eq. 2-5) $$P = P_b + p_g$$

Where: P = absolute pressure, kPa
P_b = barometric pressure (atmospheric), kPa
p_g = gage pressure, kPa

The abbrevation *a* is sometimes used to indicate that the pressure is absolute. For example, psia means pounds per square inch absolute pressure. The symbol *P* by itself, will also be used in this manual to indicate absolute pressure.

The absolute pressure allows conversion of one pressure system to the other. Relationship of the pressure systems are shown graphically in Figure 2-2 using two typical gage pressures, p_{g1} and p_{g2}. Gage pressure p_{g1} is above the zero from which gage pressures are measured, and, hence, is expressed as a positive value; gage pressure p_{g2} is below the gage pressure zero, and, therefore, is expressed as a negative value.

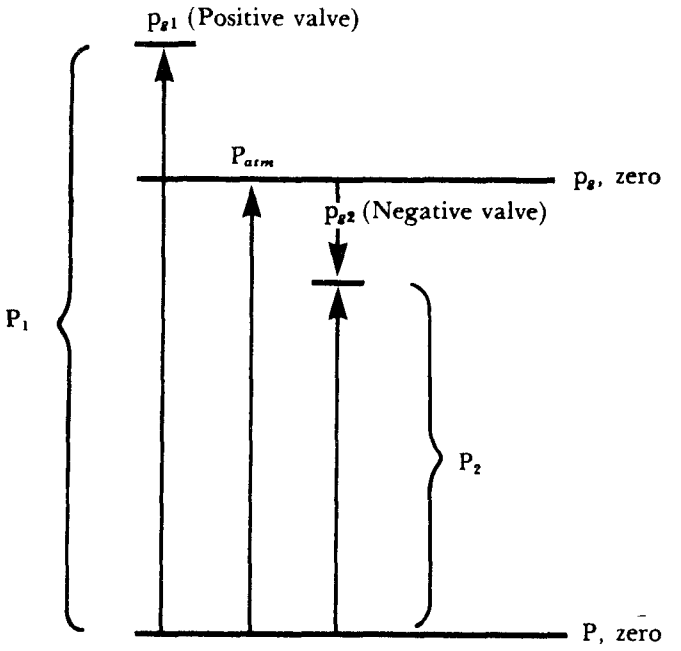


Figure 2-2. Gas-pressure relationship.

Density

Density is defined as a mass per unit volume:

(Eq. 2-6)

$$\varrho = \frac{M}{V}$$

Where: ϱ = density, kg/m^3
M = mass of a substance, kg
V = volume, m^3

The density of a solid or a liquid is a function of the temperature at which the density is measured. For example, water has a density of 1.0 g/cm^3 (62.42 lb/ft^3) at 0°C. Gas densities are referenced at a given temperature and pressure. For example, the density of dry air at 101.325 kPa (760 mm Hg) and 25°C is 1.1844 mg/cm^3 (0.0739 lb/ft^3). Densities of specific compounds may be found in tables listed in handbooks such as Perry's *Chemical Engineering Handbook.*

Appendices G and H give some of the physical properties of air and water.

Molal Units

The molecular weight of a compound or an element is simply the sum of all the atomic weights (from the Periodic Chart) of all the atoms in the molecule. A mole of any pure substance is defined as the amount of substance that numerically equals the molecular weight of that compound. Therefore, by definition, a gram mole or pound mole of oxygen weighs 32 grams or pounds depending on the units used. Molal units are used extensively in air pollution control for gaseous removal calculations as they greatly simplify material balances where chemical reactions (removing pollutants) are occurring.

For mixtures of substances (gas, liquid and solids) it is also convenient to express compositions in mole fraction or mole percent instead of a mass fraction. The mole fraction is the ratio of the number of moles of one component to the total number of moles in the mixture. For control of gaseous pollutants, these mixtures involve an exhaust stream (and sometimes a liquid stream) with a certain concentration of pollutant.

Viscosity

Origin and Definition of Viscosity

Viscosity is associated with the fluid resistance to flow. Viscosity is the result of two physical phenomena: intermolecular cohesive forces, and momentum transfer between flowing strata caused by molecular agitation perpendicular to the direction of motion. Between adjacent strata of a moving fluid, a shearing stress (Υ) that is directly proportional to the velocity gradient occurs (Figure 2-3).

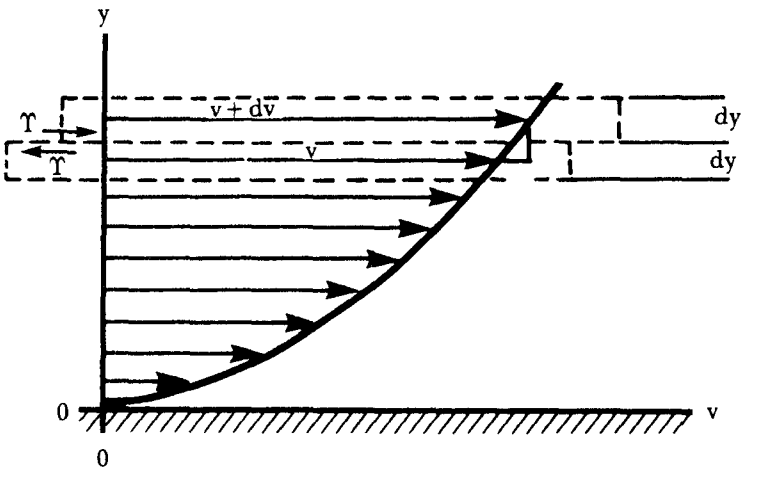


Figure 2-3. Shearing stress in a moving fluid.

This is expressed in the equation:

(Eq. 2-7) $$\Upsilon = \mu \frac{dv}{dy}$$

Where: Υ = unit shearing stress between adjacent layers of fluid
dv/dy = velocity gradient, m/s
μ = viscosity proportionality constant, centipoise

The proportionality constant, μ, is called the coefficient of viscosity, or merely, viscosity. It should be noted that the pressure does not appear in Equation 2-7 which indicates that the shear (Υ) and the viscosity (μ) are independent of pressure. (Viscosity actually increases very slightly with pressure but this variation is negligible in most engineering problems.)

Kinematic Viscosity

The ratio of the absolute viscosity to the density of a fluid often appears in dimensionless numbers such as the Reynolds Number. The expression for kinematic viscosity is used to simplify calculations. Kinematic viscosity is defined as:

(Eq. 2-8) $$\nu = \frac{\mu}{\varrho}$$

Where: ν = kinematic viscosity, m^2/s
μ = viscosity of the gas, Pa•s
ϱ = density of the gas, g/cm^3

Liquid Viscosity

In a liquid, transfer of momentum between strata having a relative velocity is small compared to the cohesive forces between molecules. Hence, shear stress is predominantly the result of intermolecular cohesion. Because forces of cohesion decrease rapidly with an increase in temperature, the shear stress decreases with an increase in temperature. Equation 2-7 shows that shear stress is directly proportional to the viscosity. Therefore, liquid viscosity decreases when the temperature increases.

Gas Viscosity

In a gas, the molecules are too far apart for intermolecular cohesion to be effective. Thus, shear stress is predominantly the result of an exchange of momentum between flowing strata caused by molecular activity. Because molecular activity increases as temperature increases, the shear stress increases with a rise in the temperature. Therefore, gas viscosity increases as the temperature rises.

Determination of Viscosity of Gases

The viscosity of a gas for the given conditions may be found accurately from the following formula:

(Eq. 2-9)

$$\frac{\mu}{\mu^{\circ}} = \left(\frac{T}{273.2}\right)^{n}$$

Where: μ = viscosity of the gas for prevailing conditions, Pa•s
μ° = viscosity at 0°C and prevailing pressure, Pa•s
T = absolute prevailing temperature, K
n = an empirical exponent (n = 0.768 for air)

The viscosity of air and other gases at various temperatures and at a pressure of 1 atmosphere may be found in engineering tables. One unit commonly used to describe viscosity has been the centipoise which equals 0.01 g/cm•s (6.72×10^{-4} lb/ft•s). In the SI system this now becomes a millipascal second (1 cp = 1 mPa•s).

Specific Heat

The specific heat of a gas is the amount of heat required to change the temperature of a unit-mass of gas one temperature-degree. Units of specific heat are calories/g•°C or Btu/lb•°F, depending upon the dimensional system used.

Heat may be added while the volume or pressure of the gas remains constant. Hence, there may be two values of specific heat; the specific heat at constant volume (C_v) and the specific heat at constant pressure (C_p).

Because the heat energy added at constant pressure is used in raising the temperature and doing work against the pressure as expansion takes place, C_p is greater than C_v. The most commonly encountered heat capacity is C_p and unless otherwise noted will be the value used in this manual when referring to specific heat. Since C_p varies with temperature, an average specific heat between two temperature values is often used. The average specific heat will be denoted as $\bar{C}_p$. Values for the specific heats of common gases and liquids are given in Appendix F of this manual.

Relationships of Ideal Gases

Laws of Boyle and Charles

Boyle's Law states that, when the temperature (T) is held constant, the volume (V) of a given mass of a perfect gas of a given composition varies inversely as the absolute pressure varies, i.e.:

$$V \propto \frac{1}{P} \text{ at constant } T$$

Where: $\propto$ = proportional to

Charles' Law states that, when the volume is held constant, the absolute pressure of a given mass of a perfect gas of a given composition varies directly as the absolute temperature varies, i.e.:

$$P \propto \frac{1}{T} \text{ at constant } V$$

Where: $\propto$ = proportional to

Law of Ideal Gases

Both Boyle's and Charles' Law are satisfied in the following equation:

(Eq. 2-10)

$$PV = \frac{MRT}{MW}$$

Where: P = absolute pressure
V = volume of a gas
M = mass of gas
MW = molecular weight of a gas
T = absolute temperature
R = universal gas constant

The unit of R depends upon the units of measurement used in the equation. Some useful values are:

$$1544 \frac{(\text{lb})(\text{ft})}{(\text{lb mole})(°\text{R})}$$

$$21.83 \frac{(\text{in. Hg})(\text{ft}^3)}{(\text{lb mole})(°\text{R})}$$

$$554.6 \frac{(\text{mm Hg})(\text{ft}^3)}{(\text{lb mole})(°\text{R})}$$

$$82.06 \frac{(\text{cm}^3)(\text{atm})}{(\text{g mole})(\text{K})}$$

$$0.0083 \frac{(\text{m}^3)(\text{Pa})}{(\text{g mole})(\text{K})}$$

Any value of R can be obtained by utilizing the fact, with appropriate conversion factors, that there are 22.414 liters per g mole (359 ft^3/lb mole) at 0°C (32°F) and 101.3 kPa (29.92 in. Hg).

The gas law applies to mixtures of gases as well as a pure gas. Thus a gram mole of ideal gas, or mixture of ideal gases, occupies 22.41 liters at 0°C and 101.3 kPa. This volume is called the gram-molal volume. Similarly, a pound mole of ideal gas occupies 359 ft^3 at 32°F and 1 atm. The pound-molal volume at any other combination of temperature and pressure may be calculated by Equation 2-11.

(Eq. 2-11)
$$V_2 = V_1 \frac{P_1}{P_2} \frac{T_2}{T_1}$$

Dalton's Law of Partial Pressure

When gases, or vapors (having no chemical interaction), are present as a mixture in a given space, the pressure exerted by a component of the gas mixture at a given temperature is the same as it would exert if it filled the whole space alone. The pressure exerted by one component of a gas mixture is called its partial pressure. The total pressure of the gas mixture is the sum of the partial pressures. For component A, Dalton's Law is expressed as:

(Eq. 2-12)
$$p_A = \frac{M_A RTV}{MW_A}$$

Where: p_A = partial pressure of component air mixture, Pa
M_A = mass of A, g
MW_A = molecular weight of A, g mole
V = volume of the mixture, m^3
T = absolute prevailing temperature, K

A useful concept from this and the ideal gas law is *pressure percent = volume percent = mole percent*.

For example, if a mixture of air is 21% oxygen and 79% nitrogen by volume at 101.3 kPa (1 atm) and 0°C, then this mixture contains 0.21 mole fraction oxygen and 0.79 mole fraction nitrogen. Also, the partial pressure of oxygen is equal to 21.27 kPa (0.21 atm) and the partial pressure of nitrogen is 80.03 kPa (0.79 atm) at 0°C. In mathematical terms the mole fraction in the gas phase is related to the partial pressure by:

(Eq. 2-13)
$$y_A = \frac{p_A}{P_T}$$

Where: p_A = partial pressure of A, Pa
P_T = total pressure of the system, Pa

Gas-Liquid Relationships

Raoult's Law states that the partial pressure of a component over a solution is the product of the vapor pressure of that component and the mole fraction of that component. Raoult's Law is expressed as:

(Eq. 2-14)

$$p_A = p_A{}^{\circ} x_A$$

Where: p_A = partial pressure of A, Pa
$p_A{}^{\circ}$ = vapor pressure of A, Pa
x_A = mole fraction of A in solution

Raoult's Law applies over the entire concentration range from 0 to 1.0, but is applicable only if the solution behaves in an ideal manner. Unfortunately few solutions behave ideally. A special case of Raoult's Law is Henry's Law. Henry's Law states that the concentration of a gas dissolved in solution is directly proportional to its partial pressure at a constant temperature.

(Eq. 2-15)

$$p_A = \mathscr{H} x_A$$

Where: $\mathscr{H}$ = Henry's Law constant, Pa/mole fraction

Henry's Law applies only if a plot of the partial pressure versus the mole fraction of component A is a straight line (discussed in more detail in Chapter 4, Absorption). This relationship holds true for more solutions in the dilute concentration range. The value and concentration range over which Henry's Law constant applies can be found only by experiment.

Reynolds Number

Definition of Reynolds Number

A typical inertial force per unit of fluid is $\varrho v^2/l$; a typical viscous force per unit volume of fluid is $\mu v/l^2$. The first expression divided by the second provides the dimensionless ratio known as Reynolds Number:

(Eq. 2-16)

$$Re = \frac{l v \varrho}{\mu}$$

Where: ϱ = density of the fluid, g/cm^3
v = velocity of the fluid, m/s
l = linear dimension, m
μ = viscosity of the fluid, Pa/s
Re = Reynolds Number (dimensionless)

The larger the Reynolds Number, the smaller is the effect of viscous forces; the smaller the Reynolds Number, the greater the effect of the viscous forces.

The linear dimension, l, is a length characteristic of the flow system. It is equal to four times the mean hydraulic radius, which is the cross sectional area divided by the wetted perimeter. Thus, for a circular pipe, l = diameter of the pipe; for a particle settling in a fluid medium, l = diameter of the particle; for a rectangular duct, l = twice the length times the width divided by the sum; and for an annulus such as a rotameter system, l = outer diameter minus the inner diameter.

Laminar and Turbulent Flow

In **laminar** flow, the fluid is constrained to motion in layers (or laminae) by the action of viscosity. The layers of fluid move in parallel paths that remain distinct from one another; any agitation is of a molecular nature only. Laminar flow occurs when Reynolds Number for circular pipes is less than 2000 and less than 0.1 for particles settling in a fluid medium.

In **turbulent** flow, the fluid is not restricted to parallel paths but moves forward in a haphazard manner. Fully turbulent flow occurs when the Reynolds Number is greater than 2500 for circular pipes and greater than 1000 for settling particles.

Chapter 3
Combustion

Introduction

The process of combustion is most often used to control the emissions of volatile organic compounds from process industries. At a sufficiently high temperature and adequate residence time, any organic vapor can be oxidized to carbon dioxide and water by the combustion process. Combustion systems are often relatively simple devices capable of achieving very high removal efficiencies. They consist of: burners, which ignite the fuel and organic vapors; and a chamber, which provides appropriate residence time for the oxidation process. Due to the high cost and decreasing supply of fuels, combustion systems are designed to include some type of heat recovery. If heat recovery can be used, combustion can be a very effective control technique. For example, pollutant emissions from paint bake ovens can be reduced by 99.9+ % using incineration while heat recovered from the incinerator flue gases can be fed back to the oven. Combustion can also be used for serious emission problems which require high destruction efficiencies, such as odor problems or the emission of toxic gases.

There are, however, some problems that may occur when using combustion to control gaseous pollutants. Incomplete combustion of many organic compounds results in the formation of aldehydes and organic acids which may create an additional pollution problem. Oxidizing organic compounds containing sulfur or halogens produce unwanted pollutants such as sulfur dioxide, hydrochloric acid, hydrofluoric acid or phosgene. If present, these pollutants would require a scrubber to remove them prior to release into the atmosphere.

Four basic combustion systems can be used to control combustible gaseous emissions. They are *flares, thermal oxidizers, catalytic oxidizers,* and *process boilers.* Although these devices are physically similar, the parameters under which they operate are markedly different. Choosing the proper device depends on many factors including: concentration of combustibles in the gas stream, process flow rate, control requirements, presence of contaminants in the waste stream, and an economic evaluation. This chapter will examine the principles of combustion, simple combustion calculations, and the design and operating parameters of each combustion device.

Combustion Principles

Combustion is a chemical process occurring from the rapid combination of oxygen with various elements or chemical compounds, resulting in the release of heat. The process of combustion is also referred to as *oxidation*. Most fuels used for combus-

tion are composed essentially of carbon and hydrogen, but can include other elements such as sulfur. Simplified reactions for the oxidation of carbon and hydrogen are given as:

(Eq. 3-1) $$C + O_2 \rightarrow CO_2 + \text{energy}$$

(Eq. 3-2) $$2H_2 + O_2 \rightarrow 2H_2O + \text{energy}$$

Equations 3-1 and 3-2 show that the final products of combustion from an organic fuel are carbon dioxide and water vapor. Although combustion seems to be a very simple process that is well understood, in reality it is not. The exact manner in which a fuel is oxidized does not occur exactly as given in the above equations, but rather in a series of complex, free radical chain reactions. The precise set of reactions by which combustion occurs is termed the *mechanism* of combustion. By analyzing the mechanism of combustion, the rate at which the reaction proceeds and the variables affecting the rate can be predicted. For most combustion devices, the rate of reaction proceeds extremely fast compared to the mechanical operation of the device. Maintaining efficient and complete combustion is somewhat of an art rather than a science, as anyone who has built a campfire can attest to. Therefore, this chapter will focus on the factors which influence the completeness of combustion, rather than analyzing the mechanisms involved.

To achieve complete combustion once the air (oxygen) and fuel have been brought into contact, the following conditions must be provided: a *temperature* high enough to ignite the air and fuel mixture; *turbulent mixing* of the air and fuel; and sufficient *residence time* for the reaction to occur. These three conditions are referred to as the "three T's of combustion". Time, temperature, and turbulence govern the speed and completeness of reaction. They are not independent variables since changing one affects the other two.

Temperature

The rate at which a combustible compound is oxidized is greatly affected by temperature. The higher the temperature, the faster the oxidation reaction will proceed. The chemical reactions involved in the combination of a fuel and oxygen can occur even at room temperature, but very slowly. For this reason, a pile of oily rags can be a fire hazard. Small amounts of heat are liberated by the slow oxidation of the oils. This in turn raises the temperature of the rags and increases the oxidation rate, liberating more heat. Eventually a full-fledged fire can break out.

For combustion processes, ignition is accomplished by adding heat to speed up the oxidation process. Heat is needed to combust any mixture of air and fuel until the *ignition temperature* of the mixture is reached. By gradually heating a mixture of fuel and air, the rate of reaction and energy released will gradually increase until the reaction no longer depends on the outside heat source. More heat is being generated than is lost to the surroundings. The ignition temperature must be reached or exceeded to ensure complete combustion.

The ignition temperature of various fuels and compounds can be found in combustion handbooks such as the North American Combustion Handbook (1965). These temperatures are dependent on combustion conditions and therefore should be used only as a guide. Ignition depends on (EPA, 1972):

1. concentration of combustibles in the waste stream
2. inlet temperature of the waste stream
3. rate of heat loss from combustion chamber
4. residence time and flow pattern of the waste stream
5. combustion chamber geometry and materials of construction.

Most incinerators operate at higher temperatures than the ignition temperature which is a minimum temperature. Thermal destruction of most organic compounds occurs between 590 and 650°C (1100 and 1200°F). However, most incinerators are operated at 700 to 820°C (1300 to 1500°F) to convert CO to CO_2; which occurs only at these higher temperatures.

Time

Time and temperature affect combustion in much the same manner as temperature and pressure affect the volume of a gas. When one variable is increased, the other may be decreased with the same end result. With a higher temperature, a shorter residence time can achieve the same degree of oxidation. The reverse is also true, a higher residence time allows the use of a lower temperature. In describing incinerator operation, these two terms are always mentioned together. One has little meaning without specifying the other.

The choice between higher temperature or longer residence time is based on economic considerations. Increasing residence time involves using a larger combustion chamber resulting in a higher capital cost. Raising the operating temperature increases fuel usage which also adds to the operating costs. Fuel costs are the major operating expense for most incinerators. Within certain limits, **lowering** the temperature and **adding volume** to increase residence time can be a cost effective alternative method of operation.

The residence time of gases in the combustion chamber may be calculated from:

(Eq. 3-3)

$$\Theta = \frac{V}{Q}$$

Where: Θ = residence time, s
V = chamber volume, m^3
Q = gas volumetric flow rate at combustion conditions, m^3/s

Q is the total flow of hot gases in the combustion chamber. Adjustments to the flow rate must include any outside air added for combustion. Example 3-1 shows the determination of residence time from the volumetric flow rate of gases.

Turbulence

Proper mixing is important in combustion processes for two reasons. First, for complete combustion to occur, every particle of fuel must come in contact with air (oxygen). If not, unreacted fuel will be exhausted from the stack. Second, not all of the fuel or waste gas stream is able to be in direct contact with the burner flame. In most incinerators a portion of the waste stream bypasses the flame and is mixed at some point downstream of the burner with the hot products of combustion. If the two streams are not completely mixed, a portion of the waste stream will not react at the required temperature and incomplete combustion will occur.

A number of methods are available to improve mixing the air and combustion streams. Some of these include the use of refractory baffles, swirl fired burners, and baffle plates. These devices are discussed in more detail in the equipment section of this chapter. The problem of obtaining complete mixing is not easily solved. Unless properly designed, many of these mixing devices may create "dead spots" and reduce operating temperatures. Merely inserting obstructions to increase turbulence is not the answer. According to one study of afterburner systems, "the process of mixing flame and fume stream to obtain a uniform temperature for decomposition of pollutants is the most difficult part in the design of the afterburner" (EPA, 1972).

Oxygen Requirements

Oxygen is necessary for combustion to occur. To achieve complete combustion of a compound, a sufficient supply of oxygen must be present to convert all of the carbon to CO_2. This quantity of oxygen is referred to as the *stoichiometric* or *theoretical amount*. The stoichiometric amount of oxygen is determined from a balanced chemical equation summarizing the oxidation reactions. For example, from Equation 3-4, 1 mole of methane requires 2 moles of oxygen for complete combustion.

(Eq. 3-4) $$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$$

If an insufficient amount of oxygen is supplied, the mixture is referred to as *rich*. There is not enough oxygen to combine with all the fuel so that incomplete combustion occurs. This condition results in black smoke being exhausted. If more than the stoichiometric amount of oxygen is supplied, the mixture is referred to as *lean*. The added oxygen plays no part in the oxidation reaction and passes through the incinerator.

Oxygen for combustion processes is supplied by using air. Since air is essentially 79% nitrogen and 21% oxygen, a larger volume of air is required than if pure oxygen were used. To balance Equation 3-4, 9.53 moles of air would be required to completely combust the 1 mole of methane. The stoichiometric calculations for this are presented in Appendix D. A listing of theoretical air requirements is given in Table 3-1.

Table 3-1. Combustion constants and approximate limits of flammability of gases and vapors in air.

Substance	Lb/ft³	Ft³/lb	Heat of combustion (Btu/ft³) Gross (high)	Heat of combustion (Btu/ft³) Net (low)	Heat of combustion (Btu/lb) Gross (high)	Heat of combustion (Btu/lb) Net (low)	For 100% total air (mol/mol of combustible) (ft³/ft³ of combustible) Required for combustion O_2	Required N_2	Required Air	Flue products CO_2	Flue products H_2O	Flue products N_2	For 100% total air (lb/lb of combustible) Required for combustion O_2	Required N_2	Required Air	Flue products CO_2	Flue products H_2O	Flue products N_2	Flammability limits (% by volume) Lower	Upper
Carbon, C*	—	—	—	·	14,093	14,093	1.0	3.76	4 76	1.0	—	3.76	2.66	8.86	11.53	3.66	—	8.86	—	—
Hydrogen, H_2	0.0053	187.723	325	275	61,100	51,623	0.5	1.88	2.38	—	1.0	1 88	7.94	26.41	34.34	—	8.94	26.41	4.00	74 20
Oxygen, O_2	0.0846	11.819	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—
Nitrogen (atm), N_2	0.0744	13.443	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—
Carbon monoxide, CO	0.0740	13.506	322	322	4,347	4,347	0 5	1.88	2.38	1 0	—	1.88	0.57	1.90	2.47	1.57	—	1 90	12.50	74 20
Carbon dioxide, CO_2	0 1170	8.548	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—
Paraffin series																				
Methane, CH_4	0.0424	23.565	1013	913	23,879	21,520	2.0	7.53	9.53	1.0	2 0	7.53	3.99	13.28	17.27	2.74	2.25	13.28	5.00	15.00
Ethane, C_2H_6	0 0803	12 455	1792	1641	22,320	20,432	3 5	13.18	16.68	2.0	3.0	13.18	3 73	12.39	16.12	2.93	1.80	12.39	3.00	12.50
Propane, C_3H_8	0 1196	8.365	2590	2385	21,661	19,944	5.0	18.82	23.82	3.0	4 0	18.82	3.63	12.07	15.70	2.99	1.68	12 07	2.12	9.35
n-Butane, C_4H_{10}	0.1582	6.321	3370	3113	21,308	19,680	6.5	24.47	30.97	4.0	5.0	24.47	3.58	11.91	15.49	3.03	1.55	11.91	1.86	8.41
Isobutane, C_4H_{10}	0.1582	6.321	3363	3105	21,257	19,629	6.5	24.47	30.97	4.0	5.0	24.47	3 58	11.91	15.49	3.03	1.55	11.91	1.80	8.44
n-Pentane, C_5H_{12}	0.1904	5.252	4016	3709	21,091	19,517	8.0	30 11	38.11	5.0	6.0	30 11	3.55	11.81	15.35	3.05	1.50	11.81	—	—
Isopentane, C_5H_{12}	0 1904	5.252	4008	3716	21,052	19,478	8.0	30.11	38.11	5.0	6 0	30.11	3.55	11.81	15.35	3.05	1.50	11.81	—	—
Neopentane, C_5H_{12}	0.1904	5.252	3993	3693	20,970	19,396	8.0	30.11	38.11	5.0	6.0	30.11	3.55	11.81	15.35	3.05	1.50	11.81	—	—
n-Hexane, C_6H_{14}	0.2274	4.398	4762	4412	20,940	19,403	9.5	35.76	45.26	6.0	7.0	35.76	3.53	11.74	15.27	3.06	1.46	11.74	1.18	7 40
Olefin series																				
Ethylene, C_2H_4	0.0746	13.412	1614	1513	21,644	20,295	3.0	11.29	14.29	2.0	2.0	11.29	3.42	11.39	14.81	3.14	1.29	11.39	2 75	28 60
Propylene, C_3H_6	0.1110	9 007	2336	2186	21,041	19,691	4.5	16.94	21.44	3.0	3 0	16 94	3.42	11 39	14.81	3.14	1.29	11.39	2.00	11.10
n-Butene, C_4H_8	0 1480	6.756	3084	2885	20,840	19,496	6.0	22.59	28.59	4.0	4 0	22.59	3 42	11.39	14 81	3.14	1.29	11.39	1.75	9 70
Isobutene, C_4H_8	0 1480	6.756	3068	2869	20,730	19,382	6.0	22.59	28.59	4.0	4.0	22.59	3.42	11.39	14.81	3.14	1.29	11.39	—	—
n-Pentene, C_5H_{10}	0 1852	5.400	3836	3586	20,712	19,363	7.5	28.23	35.73	5.0	5.0	28.23	3.42	11.39	14.81	3.14	1.29	11.39	—	—
Aromatic series																				
Benzene, C_6H_6	0.2060	4.852	3751	3601	18,210	17,480	7.5	28.23	35.73	6.0	3.0	28.23	3.07	10.22	13.30	3.38	0.69	10.22	1.40	7.10
Toluene, C_7H_8	0.2431	4.113	4484	4284	18,440	17,620	9.0	33.88	42.88	7.0	4.0	33.88	3.13	10.40	13.53	3.34	0.78	10.40	1.27	6.75
Xylene, C_8H_{10}	0.2803	3.567	5230	4980	18,650	17,760	10.5	39.52	50.02	8.0	5.0	39.52	3.17	10.53	13.70	3.32	0.85	10.53	1.00	6.00
Miscellaneous gases																				
Acetylene, C_2H_2	0.0697	14.344	1499	1448	21,500	20,776	2.5	9.41	11 91	2.0	1.0	9.41	3.07	10.22	13.30	3.38	0.69	10.22	—	—
Napthalene, $C_{10}H_8$	0.3384	2.955	5854	5654	17,298	16,708	12.0	45.17	57.17	10.0	4.0	45.17	3.00	9.97	12.96	3.43	0.56	9.97	—	—
Methyl alcohol, CH_3OH	0.0846	11.820	868	768	10,259	9,078	1.5	5.65	7.15	1.0	2.0	5.65	1.50	4.98	6.48	1.37	1.13	4.98	6.72	36.50
Ethyl alcohol, C_2H_5OH	0.1216	8.221	1600	1451	13,161	11,929	3.0	11.29	14.29	2.0	3.0	11.29	2.08	6.93	9.02	1.92	1.17	6.93	3.28	18.95
Ammonia, NH_3	0.0456	21.914	441	365	9,668	8,001	0.75	2.82	3.57	—	1.5	3.32	1.41	4.69	6.10	—	1.59	5.51	15.50	27.00
	—	—								SO_2						SO_2				
Sulfur, S*	0.0911	10.979	—	—	3,983	3,983	1.0	3.76	4.76	1.0	—	3.76	1.00	3.29	4.29	2.00	—	3.29	—	—
Hydrogen sulfide, H_2S	0.1733	5.770	647	596	7,100	6,545	1.5	5.65	7.15	1.0	1.0	5.65	1.41	4.69	6.10	1.88	0.53	4.69	4.30	45.50
Sulfur dioxide, SO_2	0.0476	21.017	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—
Water vapor, H_2O	0.0766	13.063	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—
Air	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—
Gasoline	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	1.40	7.60

*Carbon and sulfur are considered as gases for molal calculations only.

Sources: Adapted from *Fuel Flue Gases*, American Gas Association.
Combustion Flame and Explosions of Gases, 1951.

In industrial applications, more than the stoichiometric amount of air is used to ensure complete combustion. This extra volume is referred to as *excess air*. If ideal mixing were achievable, no excess air would be necessary. However, most combustion devices are not capable of achieving ideal mixing of the fuel and air streams. The amount of excess air is held to a minimum in order to reduce heat losses. Excess air takes no part in the reaction but does absorb some of the heat produced. To raise the excess air to the combustion temperature, additional fuel must be used to make up for this loss of heat. Operating at a high volume of excess air can be very costly in terms of the added fuel required. Equations to calculate excess air are also listed in Appendix D.

In addition to the theoretical air required, Table 3-1 also lists the *volume of combustion products produced* from oxidizing a substance. This is an important term used to determine the size of the combustion chamber. Example 3-1 illustrates how these values are used. The values in Table 3-1 are given in *volume percent* and *weight percent*. Volume percent is the more important term for combustion of gaseous substances since gas flows are measured in cubic meters per second instead of weight units. For example, from Table 3-1, when 1 m^3 of methane is combusted with the theoretical amount of air, 10.53 m^3 of flue gas is produced. Table 3-1 is given in English units. Since these are volume **ratios** (ft^3/ft^3), metric units and English units are interchangeable.

Natural gas is not listed in Table 3-1 since its chemical composition can vary. When 1 m^3 of natural gas is burned with a stoichiometric amount of air, it produces approximately 11.5 m^3 (average value) of flue gas.

Combustion Limits

Not all mixtures of fuel and air are able to support combustion. The flammable or explosive limits for a mixture are the maximum and minimum concentrations of fuel in air that will support combustion. The *upper explosive limit* (UEL) is defined as the concentration of fuel which produces a nonburning mixture due to a lack of oxygen. The *lower explosive limit* (LEL) is defined as the concentration of fuel below which combustion will not be self-sustaining. Table 3-1 lists the flammability limits (LEL and UEL) for common fuels and solvents.

For example, consider that a mixture of gasoline vapors and air is at atmospheric conditions. From Table 3-1 the LEL is 1.4% by volume of gasoline vapors and the UEL is 7.6%. Any concentration of gasoline in air within these limits will support combustion. That is, once a flame has been ignited it will continue to burn. Concentrations of gasoline in air below or above these limits will not burn and can quench the flame. The lower explosive limit is the more important of the two terms in describing gas streams with combustible contaminants. Industrial processes which handle combustible vapors, such as paint or solvent vapors, are usually required by insurance companies to operate at less than 25% of the LEL of the vapor in the ducts to minimize fire hazards. By using gas analyzers and an alarm system, the concentration of vapor may be allowed to be as high as 50% of the LEL by an insurance company covering the plant.

Flammability limits are not absolute values, but are affected by temperature,

pressure, geometry of the chamber, and presence of other contaminants. The higher the temperature, the greater the activation energy of combustion is. Thus, there is a greater probability that the flame will propagate. At high enough temperatures even vapor concentrations within 25% of their LEL can propagate a flame. This high degree of flammability limits the temperature to which some waste streams can be preheated prior to oxidation. Pressure also changes vapor flammability by its effect on gas densities. The higher the pressure, the closer the gas molecules are to one another. This decreases the distance a flame or spark must jump from one combustible point to another (Bethea, 1978).

The size and geometry of the combustion chamber also affect the flammability limits. The smaller the diameter of the chamber, the narrower the flammability range will become. Increased surface-to-volume ratio promotes rapid cooling and flame quenching, limiting flammability. Some flame arresters are designed on this principle of large surface-to-air volume ratio to eliminate fire hazards. The presence of other compounds can also affect flammability limits. The flammability range of a mixture can be either increased or decreased depending on the properties of the contaminant.

Flame Combustion

When mixing fuel and air, two different mechanisms of combustion can occur. A luminous (yellow) flame results when air and fuel flowing through separate ports are ignited at the burner nozzle. The yellow flame results from thermal cracking of the fuel. Cracking occurs when hydrocarbons are intensely heated before they have a chance to combine with oxygen. The cracking releases both hydrogen and carbon which diffuse to the flame to form CO_2 and H_2O. The carbon particles give the flame the yellow appearance. If incomplete combustion occurs from flame temperature cooling or if there is insufficient oxygen, soot and black smoke will form.

Blue flame combustion occurs when the fuel and air are premixed in front of the burner nozzle. This produces a short, intense, blue flame. The reason for the different flame is that the fuel-air mixture is gradually heated. The hydrocarbon molecules are slowly oxidized, going from aldehydes and ketones to CO_2 and H_2O. No cracking occurs and no carbon particles are formed. Incomplete combustion results in the release of the intermediate, partially oxidized compounds. Blue haze and odors are emitted from the stack.

Example 3-1

Emissions from a paint baking oven are controlled by a thermal oxidizer. The unit has a diameter of 1.5 m (5 ft) and is 3.5 m (12 ft) long. The exhaust from the oven is 3.8 m^3/s (800 scfm). The oxidizer uses 0.14 m^3/s (300 scfm) of natural gas and operates at a temperature of 760°C (1400°F). If all the oxygen necessary for combustion is supplied from the process stream (no outside air added), what is the residence time of the gases in the chamber?

Solution:

To solve the problem one can use the approximation given previously that 11.5 m^3 of combustion products are formed for every 1.0 m^3 of natural gas burned at stan-

dard conditions (16 °C and 101.3 kPa). Also, 10.33 m^3 of theoretical air is required to combust 1 m^3 of natural gas at standard conditions.

1. First determine the volume of **combustion products** from burning the natural gas.

$$0.14\ \frac{m^3}{s} \times \frac{11.5\ m^3 \text{ of product}}{1.0\ m^3 \text{ of gas}} = 1.61\ \frac{m^3}{s}$$

2. Then determine the air **required** for combustion.

$$0.14\ \frac{m^3}{s} \times \frac{10.33\ m^3 \text{ of air}}{1.0\ m^3 \text{ of gas}} = 1.45\ \frac{m^3}{s}$$

Sum up the volumes.

flow from paint bake oven =	3.8 m^3/s
products of combustion =	1.61 m^3/s
minus the air from exhaust used in combustion =	(− 1.45 m^3/s)
total volume =	3.96 m^3/s

Convert to actual m^3/s.

$$3.96\ \frac{m^3}{s} \times \frac{(273\,°C + 760\,°C)}{(273\,°C)} = 14.98\ m^3/s$$

3. The volume of the chamber is given by:

$$\begin{aligned} V &= \pi r^2 \ell \\ &= \pi(0.75\ m)^2(3.5\ m) \\ &= 6.18\ m^3 \end{aligned}$$

4. The residence time from Equation 3-3 is:

$$\Theta = \frac{6.18\ m^3}{14.98\ m^3/s} = 0.41\ sec$$

Combustion Calculations

Definitions

In describing any combustion process, there are numerous terms used to define *heat*. These terms, describing the heat of a system, can be broken into two categories: combustion and thermodynamic. Combustion terms were initiated to aid in standardizing fuel usage calculations. These terms are applied to heat that is produced by combustion methods. Thermodynamic terms apply to all systems. Thermodynamic terms define the energy level or potential (in terms of heat) that is present in any substance. Since the combustion terms are specific examples of the thermodynamic terms, some overlap is involved in defining them.

The following are important terms describing heat *thermodynamically*:

Sensible heat (H_s): Heat which when added or removed causes a change in temperature.

Latent heat (H_v): Heat given off by a vapor condensing to a liquid or gained by a liquid evaporating to a vapor, without a change in temperature. The latent heat of vaporization of water at 212°F is 970.3 Btu/lb.

Heat content or *enthalpy* (H): The sum total of latent and sensible heat present in a substance (gas, liquid or solid) **minus** that contained at an arbitrary set of conditions chosen as the base or zero point. Values for various gases are listed in Table 3-2.

Table 3-2. Heat contents of various gases.

Temp (°F)	Heat content, H (Btu/lb at 1 atm)								
	O_2	N_2	Air	CO	CO_2	SO_2	H_2	CH_4	H_2O
60	0	0	0	0	0	0	0	0	0
100	8.8	9.9	9.6	10.0	8.0	5.9	137	21.0	–
200	30.9	34.8	33.6	34 9	29.3	21 4	484	76.1	–
300	53.3	59.9	57.7	59.9	52.0	37 5	832	136.4	1165
400	76.2	85.0	81.8	85 0	75.3	54.4	1182	202.1	1212
500	99.4	110.3	106.0	110.6	99.8	71.8	1532	272.6	1259
600	123.1	136.1	130.2	136.3	125 1	89.8	1882	347.8	1307
700	147.2	161.7	154 5	162.4	149.6	108.2	2233	427.4	1355
800	171.7	187.7	178.9	188.7	177 8	127.0	2584	511.2	1404
900	196.6	213.9	203.4	215.6	205.6	146 1	2935	599.2	1454
1000	221.7	240.7	235.0	242.7	233.6	165.5	3291	691 1	1505
1200	272.5	294.7	288.5	297.8	290.9	205.1	4007	886.2	1609
1400	324.3	350.8	343.0	354.3	349 7	245.4	4729	1094.1	1717
1600	377.3	407.3	398.0	407.5	416.3	286.4	5460	1313.0	1829
1800	430.7	465.0	455.0	465.3	470 9	327 8	6198	1542.6	–
2000	484.0	523.8	513.0	523.8	532.8	369.1	6952	–	–
2200	539.3	583.2	570.7	583.3	596.1	411 1	7717	–	–
2400	594.4	642.3	628.5	643.0	659 2	452.7	8490	–	–
2600	649.0	702.8	687.3	703.2	723.2	495.2	9272	–	–
2800	702.8	763.1	746 6	771.3	787.4	557 5	10060	–	–
3000	758.6	824 1	806.3	832.6	852.0	580.0	10870	–	–
3200	816.4	885 8	866.0	894.0	916 7	622.5	11680	–	–
3400	873.4	947.6	925.9	956.0	981.6	665.0	12510	–	–
3600	931.0	1010.3	986 1	1018 3	1047.3	707 5	13330	–	–

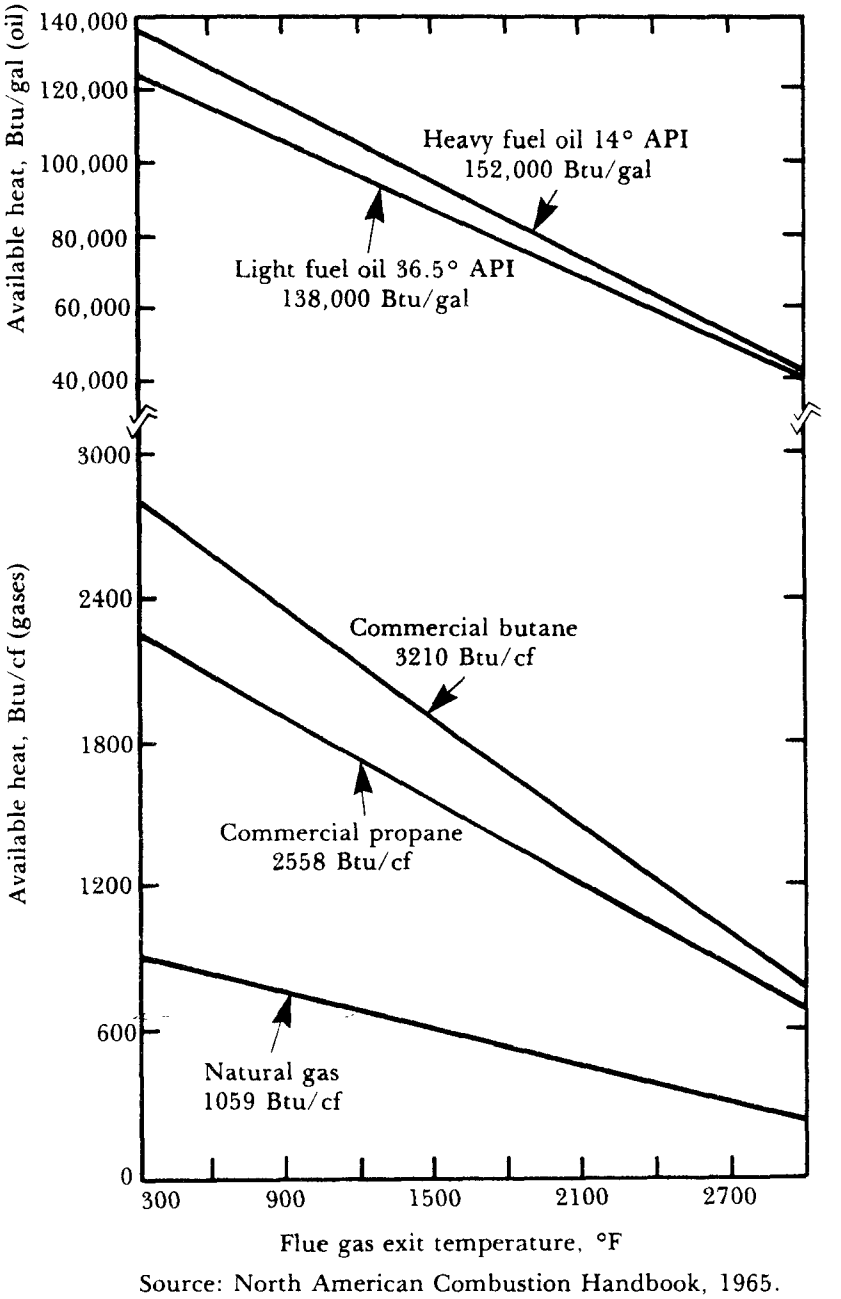
Source: North American Combustion Handbook, North American Manufacturing Co., Cleveland, OH, 1st ed. (1952).

Some useful terms describing the heat **produced** by the *combustion* of a fuel are:

Gross heating value (HV_G): The total heat obtained from the complete combustion of a fuel which is at 60°F when combustion starts, and the combustion products of which are cooled to 60°F before the quantity of heat released is measured. Constant pressure, normally 101.3 kPa (1 atmosphere) is maintained throughout the entire combustion process. Gross heating values are also referred to as *total* or *higher heating values*.

Net heating value (HV_N): The gross heating value minus the latent heat of vaporization of the water formed by the combustion of the hydrogen in the fuel. For a fuel containing no hydrogen, the net and gross heating values are the same.

Available heat (H_A): The gross quantity of heat released within a combustion chamber minus (1) the sensible heat carried away by the dry flue gases and, (2) the latent heat and sensible heat carried away in water vapor contained in the flue gases. The available heat represents the net quantity of heat remaining for useful heating. Figure 3-1 shows the available heat from the complete combustion (no excess air) of various fuels at various flue gas temperatures.



Source: North American Combustion Handbook, 1965.

Figure 3-1. Available heat for some typical fuels (referred to 60°F).

Since all of the previous terms describe heat, they all are expressed in units of Joules per kilogram (1 Btu/lb is equal to 2324 J/kg). Figure 3-2 illustrates the interrelation of these terms.

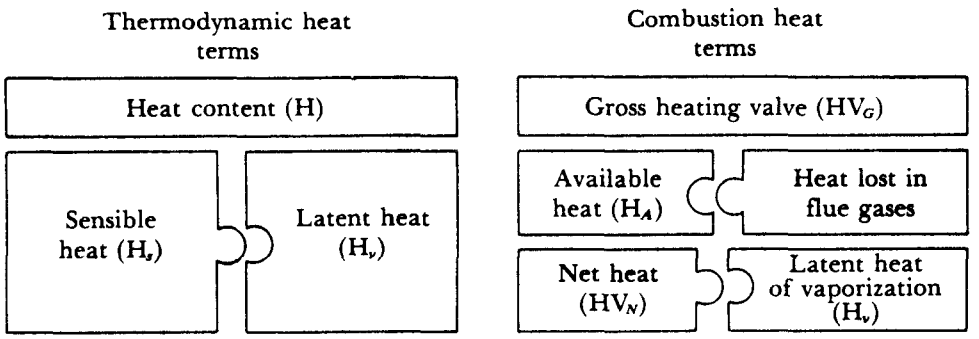


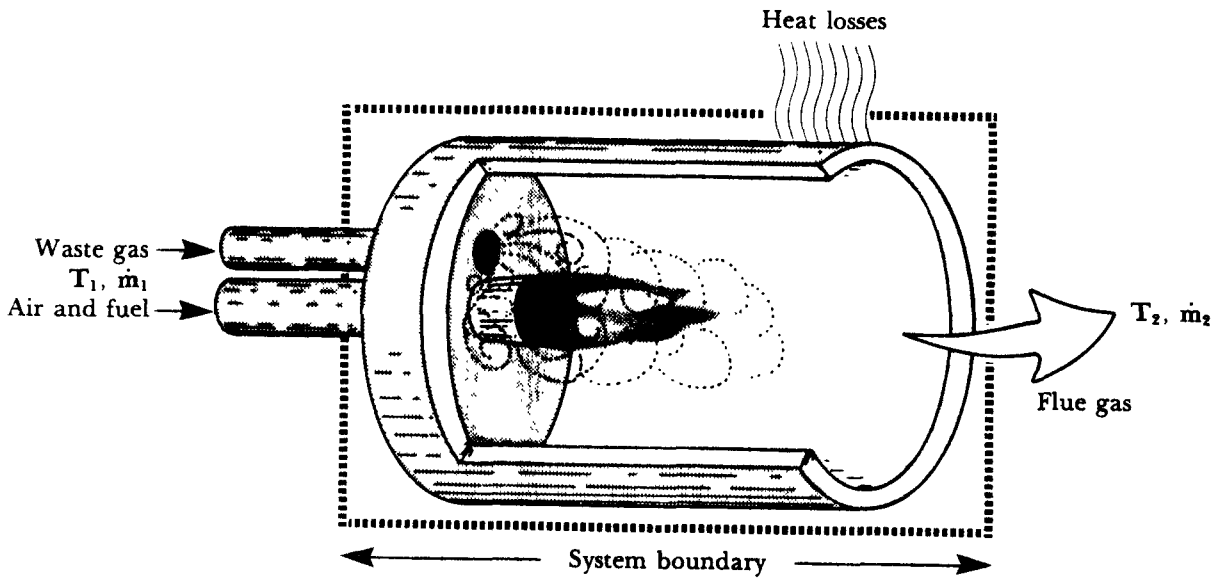
Figure 3-2. Heat terms.

Depending on the user, the above terms can also have more than one definition. For example, a laboratory chemist may describe latent heat as the energy used in the chemical combustion of a fuel to carbon dioxide and water; while a boiler operator may describe latent heat as the difference between the gross and net heating values.

Another important term used in performing combustion calculations is the *specific heat*, C_p, of a substance. Specific heat is defined as the amount of heat required to raise 1 gram of a substance 1 degree centigrade. Specific heat is given as J/kg•K in SI units and is Btu/lb•°F in English units. Specific heat depends on temperature. Appendix F gives a nomograph that can be used to compute the specific heats of common gases at various temperatures.

Heat Balance

A main area of concern in vapor incineration deals with the amount of fuel required to raise the temperature of the waste stream to the temperature required for complete oxidation. The first step in computing the heat required is to perform a heat balance around the oxidation system. Figure 3-3 shows a typical incineration system illustrating the heat entering and leaving the incinerator.



Where: T_1 = waste gas inlet temperature
T_2 = incinerator temperature
$\dot{m}_1$ = mass flow rate of waste gas plus air and fuel
$\dot{m}_2$ = mass flow rate of flue gases

Figure 3-3. Heat balance.

From the law of conservation of energy:

(Eq. 3-5) $$\text{Heat in} = \text{Heat out} + \text{Heat lost}$$

Heat is a **relative** term which is compared at a reference temperature. The heat content of a substance is arbitrarily taken as zero at a specified reference temperature. In the gas industry (natural gas), the reference temperature is normally 16°C (60°F). This is the reference temperature used throughout this chapter. Thus, the heat content can be computed from Equation 3-6 as well as found by using tables such as Table 3-2.

(Eq. 3-6) $$H = C_p(T - T_o)$$

Where: H = enthalpy, J/kg
C_p = specific heat at temperature T, J/kg•°C
T = temperature of the substance, °C
T_o = reference temperature, °C

Subtracting the heat going in from the heat leaving the system in Figure 3-3, gives the heat which must be supplied by the fuel. This is referred to as a change in enthalpy or heat content. Using Equation 3-6, the enthalpy going in (T_1) is subtracted from the enthalpy leaving (T_2) giving:

(Eq. 3-7) $$\Delta H = [C_{p2}(T_2 - T_o) - C_{p1}(T_1 - T_o)]$$

Where: ΔH = change in enthalpy

To simplify this calculation, an average specific heat value ($\overline{C}_p$) can be used over the temperature range involved. This reduces Equation 3-7 to:

(Eq. 3-8) $$\Delta H = \overline{C}_p(T_2 - T_1)$$

The specific heat varies with temperature and composition of the gas stream. Therefore, Equation 3-8 is used to obtain an approximate value. For most incineration systems, the waste gases are considered to be essentially air. For air, an average specific heat value, $\overline{C}_p$, is 1088 J/kg•°R (0.26 Btu/lb•°F) for typical temperature ranges normally encountered in fume incinerators.

Equation 3-8 depicted the amount of heat required to raise a set quantity of gas from T_1 to T_2. Of more importance is the *heat rate*, q, that is required. The heat rate is easily determined by multiplying either side of Equation 3-8 by the *mass flow rate* ($\dot{m}$) of the process gas stream. The heat rate required is given by:

(Eq. 3-9) $$q = \dot{m}\Delta H = \dot{m}\overline{C}_p(T_2 - T_1)$$

Where: q = heat rate, watts (Btu/hr)
$\dot{m}$ = mass flow rate of gases, kg/hr (lb/hr)

In either form, Equation 3-9 can be used to compute the heat rate required to raise the gas temperature from T_1 to T_2. Example 3-2 illustrates the use of Equation 3-9.

These equations are simple heat balances, equating heat in to heat out. They do not account for any heat losses in the system. Heat losses from refractory or ducting are usually accounted for by assuming a fixed percent of the total theoretical heat is lost. For example, if an afterburner is required to supply heat at the rate of 3×10^5 watts ($\approx 1 \times 10^6$ Btu/hr) and there is a 10% heat loss from the combustion chamber, the total heat rate would have to be 3.3×10^5 watts to account for the losses. Heat is also lost to the system from the latent heat associated with vaporization, sublimination and fusion of various components in the waste gas stream. This is the heat which produces a change of state at a constant temperature.

The following example illustrates a method that can be used to determine the fuel requirement for an incinerator. The example has been worked using English units since figures and tables needed to solve the problem are also in English units.

Example 3-2

The exhaust from a meat smoke house contains obnoxious odors and fumes. The company plans to incinerate the 5000 acfm exhaust stream. What quantity of natural gas is required to raise the waste gas stream from a temperature of 90 °F to the required temperature of 1200 °F? The gross heating value of natural gas is 1059 Btu/scf and assume no heat losses.

Solution:

All calculations are based on a 1 hour time period.

First, the volume of waste gas must be corrected to standard conditions (60 °F and 1 atm).

$$Q = 5000\frac{\text{ft}^3}{\text{min}}\left(\frac{460+60}{460+90}\right)\left(\frac{60\text{ min}}{\text{hr}}\right) = 2.84\times10^5\ \text{ft}^3/\text{hr}$$

The volumetric flow rate is then converted to a mass flow rate by multiplying by the density.

$$\dot{m} = Q\rho$$

(assume waste gas molecular weight is the same as air)

$$= \left(\frac{2.84\times10^5\text{ft}^3}{\text{hr}}\right)\left(\frac{\text{lb mole}}{379\ \text{ft}^3}\right)\left(\frac{29\ \text{lb}}{\text{lb mole}}\right)$$

(conversion factor for an ideal gas at 60 °F)

$$= 2.17\times10^4\ \text{lb/hr}$$

The heat rate can be determined by two methods using Equation 3-9.

1. By using the enthalpy values in Table 3-2:

   H for air at 1200 °F = 288.5 Btu/lb

   H for air at 90 °F is obtained by interpolating $= 9.6 - \left[9.6\left(\frac{10}{40}\right)\right] = 7.2$ Btu/lb

   $$q = \dot{m}\Delta H$$
   $$= \dot{m}\ (H_{air_{1200°}} - H_{air_{90°}})$$
   $$= (2.17\times10^4\ \text{lb/hr})(288.5 - 7.2\ \text{Btu/lb})$$
   $$= 6.10\times10^6\ \text{Btu/hr}$$

2. An alternative method is to use an average $\overline{C}_p$.

   For air $\overline{C}_p = 0.26$ Btu/lb•°F

   $$q = \dot{m}\overline{C}_p\Delta T$$
   $$= \left(2.17\times10^4\ \frac{\text{lb}}{\text{hr}}\right)\left(0.26\frac{\text{Btu}}{\text{lb °F}}\right)(1200\text{ °F} - 90\text{ °F})$$
   $$= 6.26\times10^6\ \text{Btu/hr}$$

   This value is not as accurate, since it was calculated using an average value for $\overline{C}_p$.

To compute the amount of natural gas required from the heating rate, the available heat of the fuel (H_A) must be computed using Figure 3-1.

Figure 3-1 is used by locating the flue gas temperature on the x axis (1200°F). Read up from this point to the line for natural gas with a heating value of 1059 Btu/scf. The heat available is read from the y axis as 690 Btu/scf.

The amount of natural gas needed is:

$$Q_{gas} = \frac{q}{H_A}$$

$$= 6.10 \times 10^6 \frac{\text{Btu}}{\text{hr}} \times \frac{\text{scf}}{690 \text{ Btu}}$$

$$= 8841 \frac{\text{scf}}{\text{hr}}$$

Combustion Equipment Used for Control of Gaseous Emissions

Introduction

Afterburning, incineration, or thermal oxidation are interchangeable terms used to describe the combustion process used to control gaseous emissions. Although not technically correct, this nomenclature is generally accepted. Afterburners and incinerators perform different functions even though both are forms of *thermal oxidation.* The term afterburner is appropriate only to describe a thermal oxidizer used to control gases coming from a process where combustion was not complete (Ross, 1977). An example of an afterburner in use would be an application on a copper wire reclaiming source which **incompletely** burns off the rubber coating. Incineration or thermal oxidation are **not** necessarily afterburning. Each is a primary burning process that is installed to control effluent pollutants which are combustible. Incinerators are used to combust solid, liquid and gaseous materials. When used in this manual, the term incinerator will refer to controlling gaseous emissions of organic vapors. Examples of sources using incineration are curing and drying ovens, chemical processes, petroleum refining processes, and food processing plants. In this chapter the terms incinerator and thermal oxidizer will be used interchangably.

Equipment used to control waste gases by combustion can be divided into three categories: *direct combustion* or *flaring, thermal oxidation,* and *catalytic oxidation.* A *direct combustor* or *flare* is a device in which air and all the combustible waste gases react at the burner. Complete combustion must occur instantaneously since there is no residence chamber. Therefore, the flame temperature is the most important variable in flaring waste gases. In contrast, in *thermal oxidation,* the combustible waste gases pass over or around a burner flame into a residence chamber where oxidation of the waste gases is completed. *Catalytic oxidation* is

very similar to thermal oxidation. The main difference is that after passing through the flame area, the gases pass over a catalyst bed which promotes oxidation at a lower temperature than does thermal oxidation.

Direct Combustion or Flaring

Direct combustion or flaring is used for the disposal of intermittent or emergency emissions of combustible gases from industrial sources. Safety and health hazards at or near the plant can be eliminated by using flares to prevent the direct venting of these emissions. Flares have been used mainly at oil refineries and chemical plants which handle large volumes of combustible gases.

Flares are simply burners that have been designed to handle varying rates of fuel while burning smokelessly. In general, flares can be classified as either *elevated* or *ground level*. The reason for elevating a flare is to eliminate any potential fire hazard at ground level. Ground level flares must be completely enclosed to conceal the flame. Either type of flare must be capable of operating over a wide range of flow rates in order to handle all plant emergencies. The range of waste gas flows within which a flare can operate and still burn efficiently is referred to as the turndown ratio. Flares are expected to handle turndown ratios of 1000:1, while most industrial boilers seldom handle more than a 10:1 turndown ratio (Gottschlich, 1977). For example, a flare should be capable of maintaining complete combustion for waste gas flow rates ranging from 20,000 m^3/h to 20 m^3/h.

Although the flare is designed to eliminate waste gas stream disposal problems, it can present safety and operational problems of its own. Some of the problems associated with operation of a flare system are:

1. Thermal radiation: Heat given off to the surrounding area may be unacceptable.
2. Light: Luminescence from the flame may be a nuisance if the plant is located in an urban area.
3. Noise: Mixing at the flare tip is done by jet venturis which can cause excess noise levels in nearby neighborhoods.
4. Smoke: Incomplete combustion can result in toxic or obnoxious emissions.
5. Energy consumption: Flares waste energy in two ways. First by keeping the pilot flame constantly lit and secondly, by the potential recovery value of the waste gas being flared.

For flaring to be economically feasible, the waste gas usually must supply at least 50% of the fuel value to combust the mixture. When the heat content of the waste gas is below 4.28×10^3 kJ/m^3 (115 Btu/ft^3), injecting an additional gas with a higher heating value is necessary to achieve complete combustion. This type of system is referred to as a *fired* or *endothermic* flare. The cost of the additional fuel for endothermic flaring can be considerable. To conserve energy, some companies have used other waste gases in place of conventional fuels.

Elevated Flares

A typical elevated flare is composed of a system which first collects the waste gases and passes them through a knockout drum to remove any liquids. Water seals or other safety devices are placed between the knockout drum and the flare stack to prevent a flashback of flames into the collection system. The flare stack is essentially a hollow pipe that may extend to a height of 100 meters or more. The diameter of the flare stack determines the volume of waste gases that can be handled. At the top of the stack is the flare tip which is comprised of the burners and a system to mix the air and fuel.

Smokeless combustion is accomplished by proper design of the flare tip. Smokeless flares require a system allowing for intimate mixing of waste gases and air. A number of flare tip designs provide good mixing characteristics over a wide range of waste gas flow values and still have excellent flame holding capabilities. Steam jets have proved to be one of the most effective ways to mix air and waste gases. An example of a flare tip designed for steam injection is illustrated in Figure 3-4. In addition to increasing turbulence, steam injection reduces the partial pressure of the waste gas, thus reducing polymerization reactions. The steam also reacts with the gases to produce oxygenated compounds which readily burn at lower temperatures. Other devices used to reduce smoke are high pressure fuel gases, water sprays, high velocity swirl fired burners, and electric air blowers (Straitz, 1980). In addition to these devices, shields are also used on elevated flares to protect the flame zone from atmospheric conditions. This shield also helps reduce noise and visibility problems associated with flares.

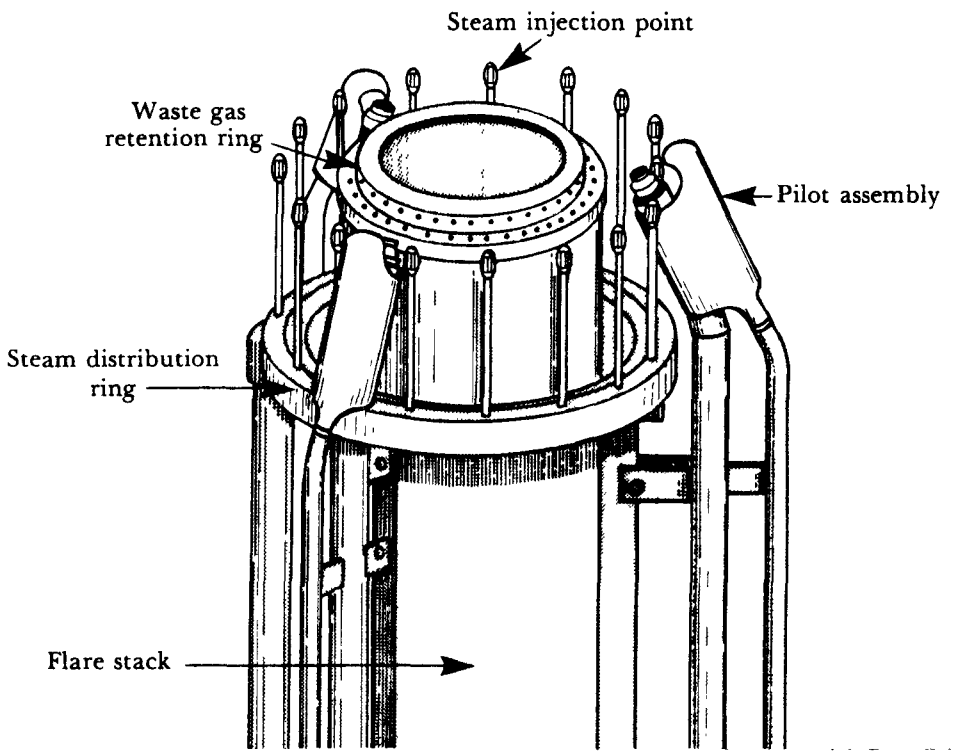


Figure 3-4. Smokeless flare tip.

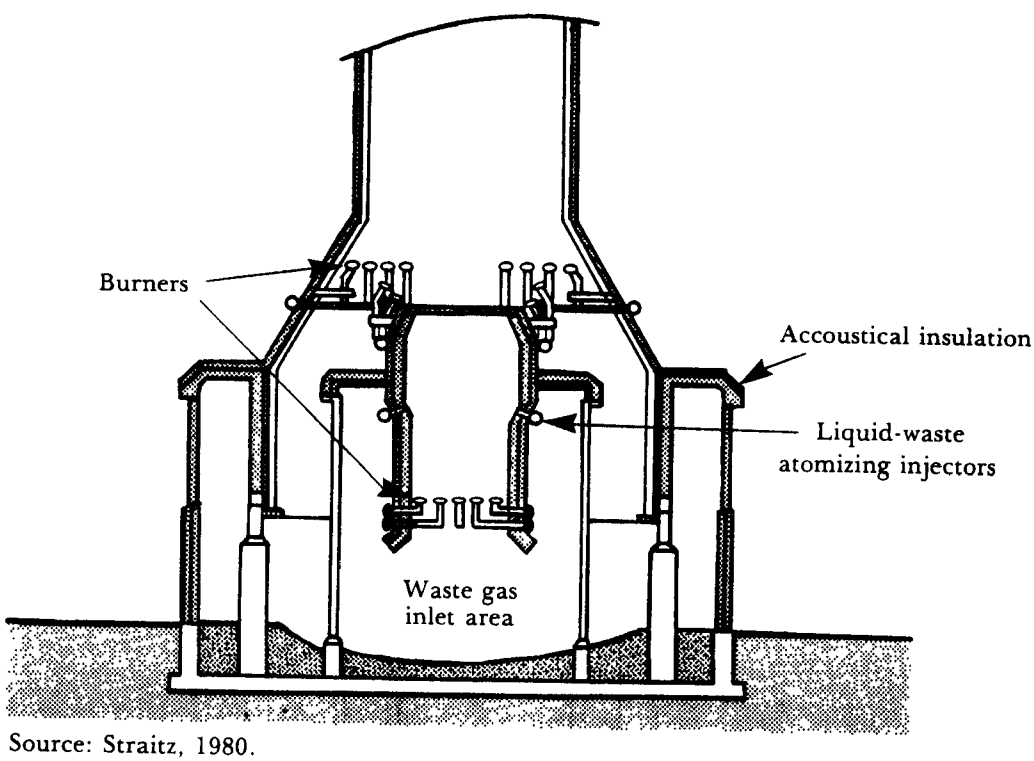
As previously mentioned, steam reacts with intermediate combustion products to form compounds that burn readily at lower temperatures. One of the main reactions taking place is a hydrolysis reaction referred to as the *water-gas reaction*. This reaction produces hydrogen gas which helps flare operation. The simplified water-gas shift can be written as:

$$CO + H_2O \rightleftharpoons H_2 + CO_2$$

Steam-to-hydrocarbon mass ratios are usually determined by the molecular weight and concentration of the unsaturated organic compounds (alkenes and alkynes) in the waste gas. Steam requirements generally range between 0.05 and 0.30 kg of steam per kg of waste gas (Gottschlich, 1977). The steam is automatically injected proportionally to the flow rate of the waste gases.

Enclosed Ground Flares

Manufacturers design a number of different ground flares. Most ground flares consist of multiple burners enclosed within a refractory shell. The shell encloses the flame to eliminate noise, luminescence, and safety hazards. The waste gas is introduced through a jet or venturi to provide turbulent mixing. The term *ground flare* refers to locating the flare tip at ground level. The flare system still requires a stack for proper release of the effluent gases. Figure 3-5 shows a ground flare installed at Nippon Steel Company in Oita City, Japan (Straitz, 1980). The flare is composed of two chambers designed to combust eight different gases and a liquid-waste stream.



Source: Straitz, 1980.

Figure 3-5. Ground flare.

Ground level flares normally have a much higher capital cost than elevated flares. The cost of a ground flare depends on the size of the enclosure. The size of a ground flare is directly proportional to the volume of vapors it must handle. Ground flares can be designed for efficient combustion without steam injection. Eliminating the use of steam injection can significantly lower operating costs as compared to elevated systems (Straitz, 1980). Some plants have incorporated both designs. A ground flare is used for normal or intermittent operation and a large elevated flare is only used to control emergency releases of large quantities of gases.

Incinerators

Thermal oxidizers or organic vapor incinerators refer to any device that uses a flame (temperature) combined with a chamber (time and turbulence) to convert combustible material to carbon dioxide and water. An incinerator usually consists of a refractory-lined chamber that is equipped with one or more sets of burners. A typical incinerator is depicted in Figure 3-6. The contaminant-laden stream is passed through the burners where it is heated above its ignition temperature. The hot gases then pass through one or more residence chambers where they are held for a certain length of time to ensure complete combustion. Depending on the particular needs of the system, additional fuel and/or excess air can be added through the burners. Also, since the flue gases are discharged at elevated temperatures, a system to recover the heat may be included. With the rising costs of fuel and solvents, heat recovery devices are becoming an integral part of many incineration systems.

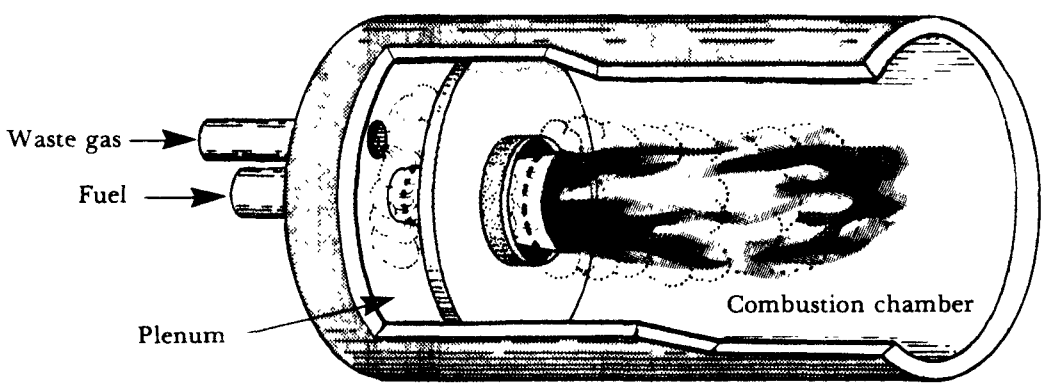


Figure 3-6. Typical thermal incinerator (UOP raw gas burner).

Incinerators on industrial processes are most often used to control gas streams with a **low concentration** of organic vapors. The concentration of vapors delivered to an incinerator through process ductwork is limited by the LEL of the mixture. Safety codes usually limit combustible vapor concentrations in ducts to a maximum of 25% of the LEL. By using process control monitors and alarm systems, some plants have been allowed to go up to 50% of the LEL. Supplemental fuel is required to ignite the waste gas stream and bring it up to the proper operating

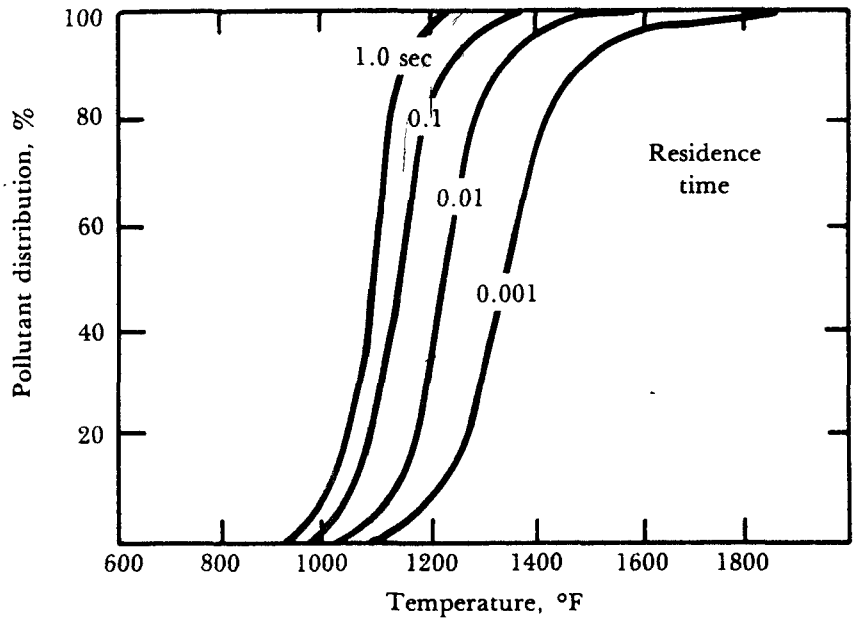
temperature. The amount of fuel is then decreased proportionally to the heating value supplied by the waste gas stream. Ideally, after initial startup, little or no auxiliary fuel would be needed.

Incinerators operate at temperatures between 700 and 800°C (1300 and 1500°F) with a residence time of 0.1 to 0.5 seconds. The residence time is determined by the size of the combustion chamber and is measured after the required temperature has been reached. Typical operating temperatures of thermal oxidizers for various processes are listed in Table 3-3 (EPA, 1972). As noted previously, temperature and residence must be discussed together since changing one affects the other. By raising the temperature, the residence time for complete combustion is reduced and vice versa. However, temperature is the more important process variable. Figure 3-7 illustrates the effect of both temperature and residence time on the percent destruction of pollutants (oxidation rate). Depending on the initial temperature, small increases in temperature can bring dramatic increases in pollutant destruction. For example, for a 0.01 second residence time increasing the temperature from 1200 to 1400°F, the percent destruction is doubled from approximately 50 to 100%. At 1200°F, the residence time must be increased ten times (from 0.01 second to over 0.1 second) for the same increase in percent pollutant destruction.

Table 3-3. Typical waste gas incinerators' operating temperatures (°F).

Industry	LAAPCD** recommendation	Survey data	Literature
Asphalt blowing	–	1000-2000	–
Biological control, fermentation	–	110-1250	–
Carpet laminating	1200-1400	–	–
Coffee roasting	–	1200-1500	1050
Coil coating, sheet coating, metal decorating	–	1200-1500	1300
Core ovens, foundry	1400	2000	–
Coating, engraving	–	1000*-1450	–
Cloth carbonization	1800	–	–
Deep fat fryers	1200	–	–
Gum label drying oven	–	–	1250
Mineral wool, fiberglass curing	–	1000*	1310
Odor control (general) sludge off-gas	1300-1500	1100-1250	1300-1425
Hardboard tempering	–	1200	1200
Oil and grease smoke (metal chip recovery, heat quench baths, tempering)	1200-1400	900*-1600	–
Paint bake ovens	1200-1500	1100-1500	1240
Paper manufacture—sulfite digester off-gas	–	–	1350
Pipe wrapping	1400	–	–
Rendering plants	1200	1200-1300	1200
Rubber products	–	1300-1400	1300
Petroleum refining and products	–	1300-2000	–
Printing, lithographing	–	1300-1500	–
Smelting, refining, metal recovery, wire burnoff	–	1300-1650	–
Smokehouse operation	1200	800*-1200	–
Solvent control	1300-1500	–	–
Varnish cookers, resin kettles	1200	1000*-1500	1200-1400
Vinyl plastisol curing	1200-1400	1200	900*
Wood milling	–	–	1200
Wire enameling	–	–	1300-1350, 1400
Phthalic anhdyride	–	1250-1440	–
Textile drying oven	–	1350	–

*Low temperature generally for odor, smoke control, not true organic vapor destruction.
**LAAPCD is the Los Angeles Air Pollution Control District.



Source: Rolke, 1972.

Figure 3-7. Effects of temperature and residence time on rate of pollutant oxidation.

Besides temperature and residence time, the concentration of the pollutant in the waste gas stream also affects operation of the incinerator. First, the concentration of the contaminant dictates the amount of supplemental fuel required. Initially, supplemental fuel must be added to start the oxidation reaction. As the temperature rises, the rate of oxidation of the contaminant increases until the reaction becomes self-sustaining. The amount of fuel can then be decreased. A certain amount of fuel is normally burned to ensure stable operation. Secondly, there is a problem of flame quenching. To avoid noncombustible mixtures, the entire amount of waste gas and fuel cannot usually be mixed at the burner. It would require an inordinate amount of fuel to bring the entire waste gas stream within combustible limits. Only part (about one-half) of the waste gas stream is mixed with the fuel at the burner. The remainder of the waste gas stream must be mixed with the hot products of combustion downstream of the flame to avoid quenching the flame. Unless adequate mixing of the waste gas stream and hot products of combustion is accomplished, incomplete combustion products will be exhausted.

Types of Burner Arrangements

The design of an incinerator must allow adequate mixing of the fuel and waste gas streams. A number of different burner designs provide the required mixing without quenching the flame. Burners are classified as either *discrete* or *distributed* (EPA, 1973). The main distinction between the two types of burners is the number and size of flames in each type. The discrete burner produces a **single large flame** plume into which the waste gas stream must be blended. The distributed burner system produces an **array of smaller flames**. The waste gas stream is divided and then flows around these flames. In distributed burners more of the waste gas stream is passed over or around the flame. This allows distributed burners to complete mixing of the waste gases and hot products of combustion in shorter distances than in discrete burners. Distributed burners also utilize the oxygen in the waste gas stream more efficiently, reducing the amount of oxygen supplied through the burner. This in turn reduces the need for additional fuel to heat the air/waste gas mixture to the required oxidation temperature.

Three types of distributed burner systems are the *line burner*, the *multijet*, and the *grid* burner. All three of these systems are comprised of a manifold system, to inject the fuel, and a mixing or profile plate which regulates air flow to the flame area. Fuel (usually natural gas) enters the burner area of the line burner (Figure 3-8) through holes in manifold pipes. The pipes are placed across the duct at 0.3 meters (1.0 foot) intervals (EPA, 1972). Waste gas enters the burner area through holes in the profile plate. The plate allows only a portion of the waste gas, usually 50%, to directly contact the flame. The remaining portion of the gas bypasses the flame area and is mixed with the hot combustion gases downstream of the flame. Figure 3-9 illustrates an example of one type of mixing plate. The plate is attached to the manifold pipes and forms a "v" shaped trough; part of the waste gas enters the flame area through the holes, while the remaining portion of waste gas passes around the plate.

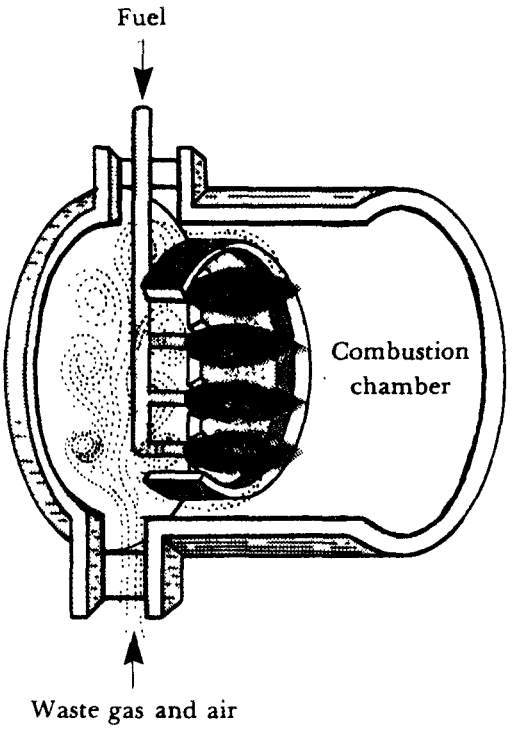


Figure 3-8. Distributed (line) burner.

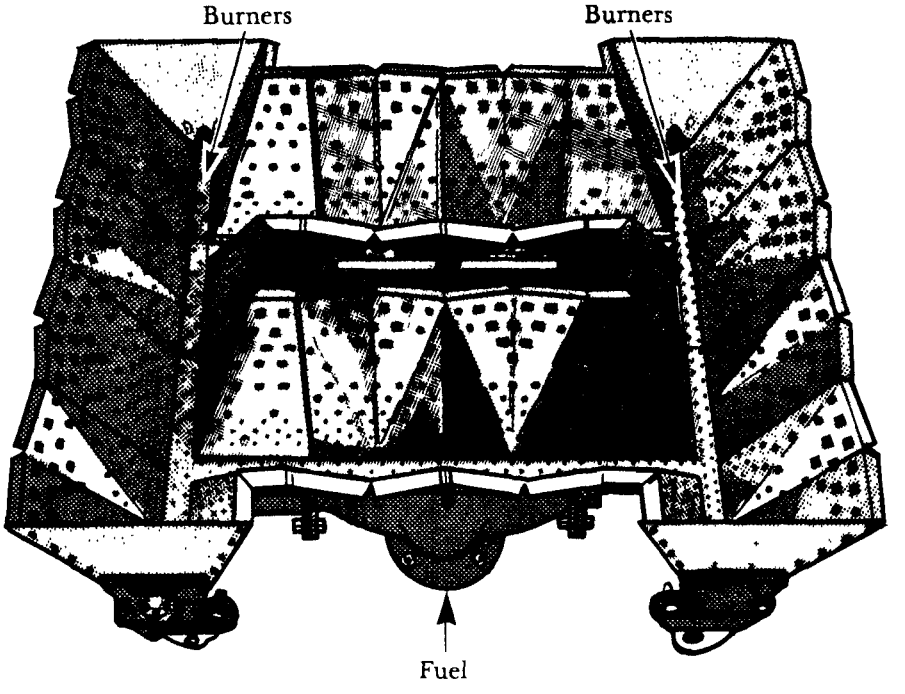


Figure 3-9. Mixing plate for waste gas and hot combustion products.

Multijet burners are very similar to line burners. Multijet burners have a single fuel inlet which is manifolded to produce numerous individual flames. Figure 3-10 shows the Hirt multijet burner system. In this system, waste gases flowing to the flame area are controlled by a profile or mixing plate. The profile plate allows a portion of the waste gas stream to flow behind it and mix with the fuel to supply the combustion area. The remaining gases must pass around in front of the profile plate. Air flow to the burner flame area is adjusted by moving the profile plate either closer or further away from the burners. Adjusting the profile plate on the multijet system is relatively simple and can be accomplished while the unit is still in service. Adjusting the mixing plate on the line burner is not as simple, because the unit must be taken off line. The line burner, however, has the advantage of being much smaller. The line burner accomplishes mixing over a relatively short distance, a few inches, as compared to the multijet which requires several feet.

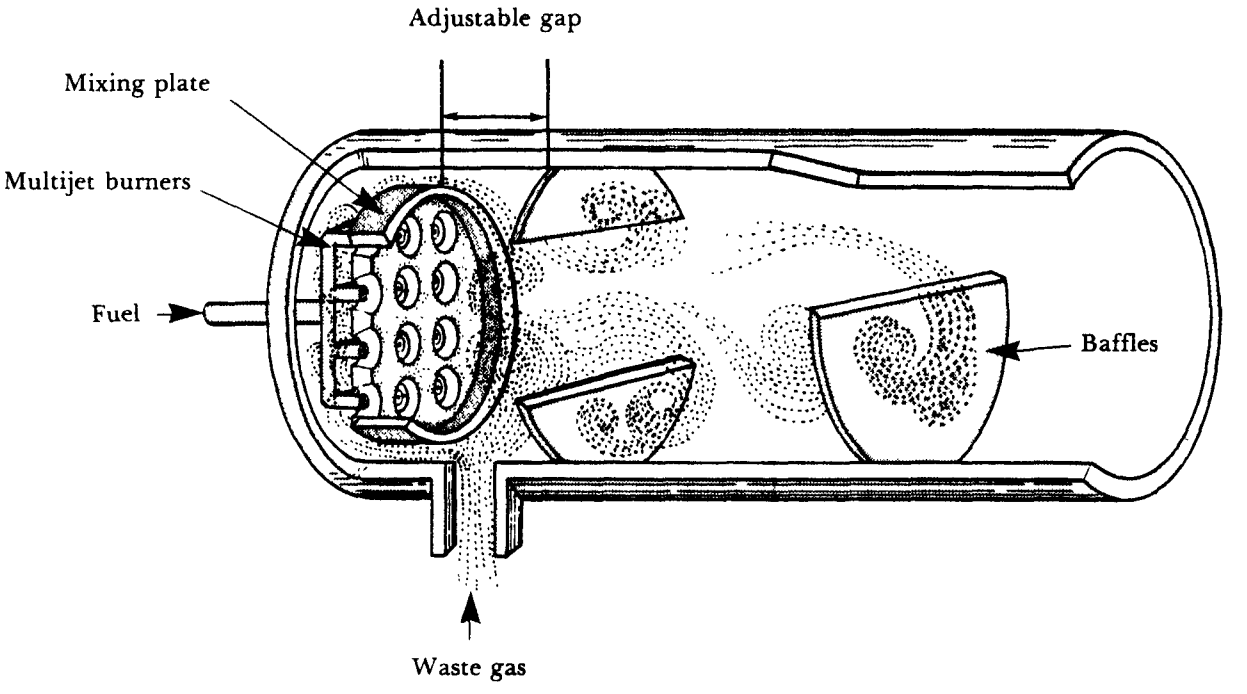


Figure 3-10. Multijet burner.

The grid burner also uses a manifold system to introduce the fuel. Mixing is accomplished by passing the waste gas stream through a slotted grid as shown in Figure 3-11. The slots are approximately 2.5 by 7.5 cm (1 by 3 in.) (EPA, 1972). The grid allows for extremely fast mixing of the fuel and waste gas. There is, however, very little control to allow for adjusting flow rates since the grid is stationary. Grid burners are best suited for applications where waste gas flow ratios are relatively constant.

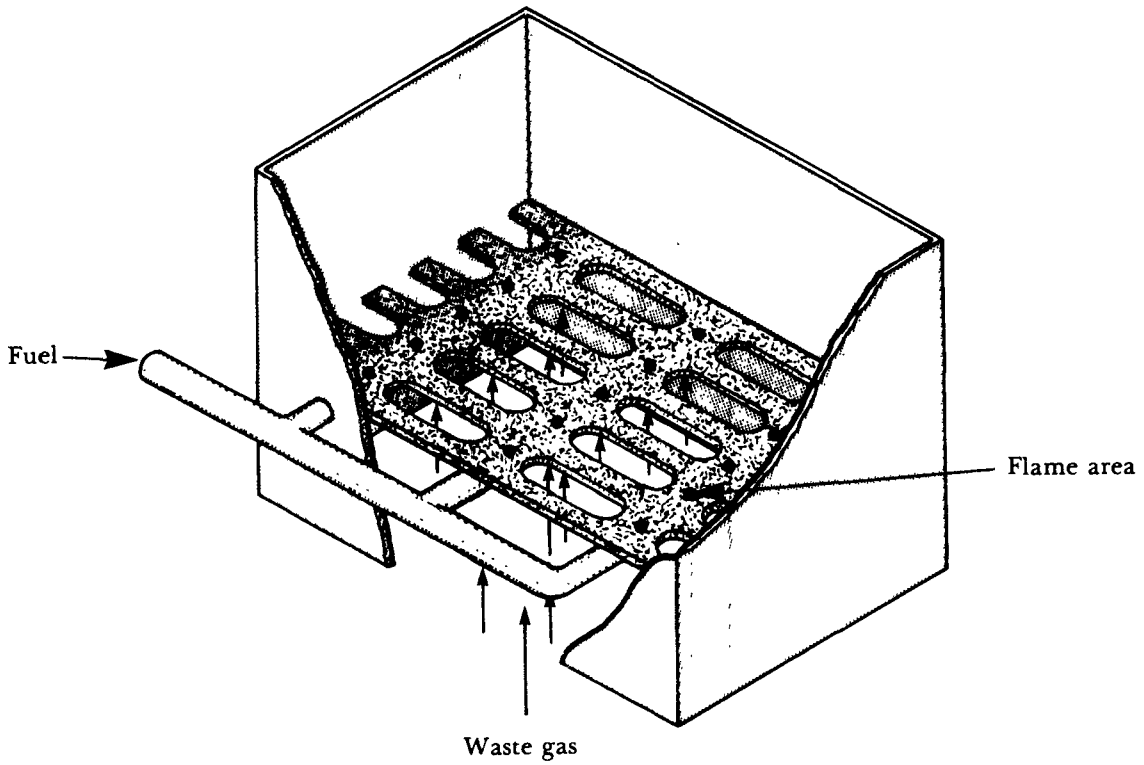


Figure 3-11. North American flame grid burner.

Problems can arise using any of the three types of distributed systems. Some of these are:

1. Natural gas is the only fuel source that can be used. If solid or liquid particles are present or formed during combustion, they can plug the tiny holes in the fuel manifold system. This is known as *burner fouling*.
2. The oxygen content of the waste gas stream must be at least 16%. Oxygen for combustion is supplied by the waste gas stream. If the oxygen content is not 16% or greater, incomplete combustion can occur.
3. Mixing, profile, and grid plates must be able to withstand high temperatures. This adds to the capital equipment costs.

To avoid some of the problems associated with distributed burners, especially fouling, discrete burners are sometimes used. The discrete burner is simply one large burner (Figure 3-12). Many other burner arrangements are classified as discrete burners. They can range from a simple gas ring, comparable to those used on home stoves, to a torch burner (Figure 3-6). In this illustration, mixing the process waste gas and hot products of combustion is accomplished by using a conical mixing plate.

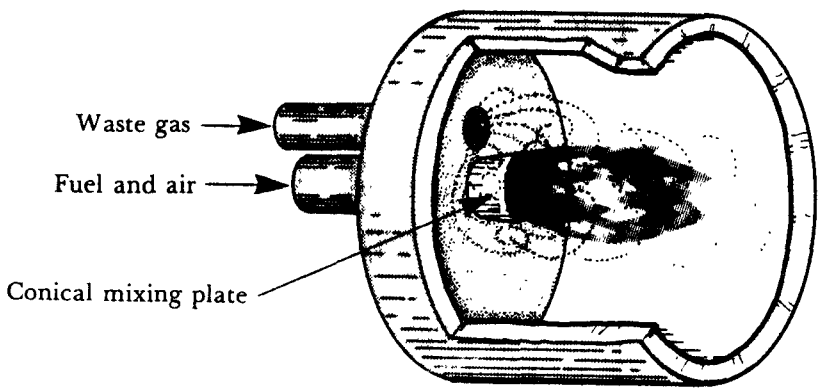


Figure 3-12. Discrete burner.

In most discrete burners, mixing must usually be provided downstream of the flame. Baffles are most commonly used to provide the required turbulence. However, baffles increase the pressure drop across the incinerator. Unless installed correctly, baffles can cause *dead zones* which actually decrease the degree of mixing. Two main types used in incinerators are the *bridge wall baffle* and the *ring and disc baffle* illustrated in Figure 3-13 and Figure 3-14. The bridge wall baffle is a wall placed across the combustion chamber. The bridge wall technique is most effective when pairs of baffles are used. The ring and disc baffles consist of a series of rings inserted in the middle along with discs on the side walls of the chamber.

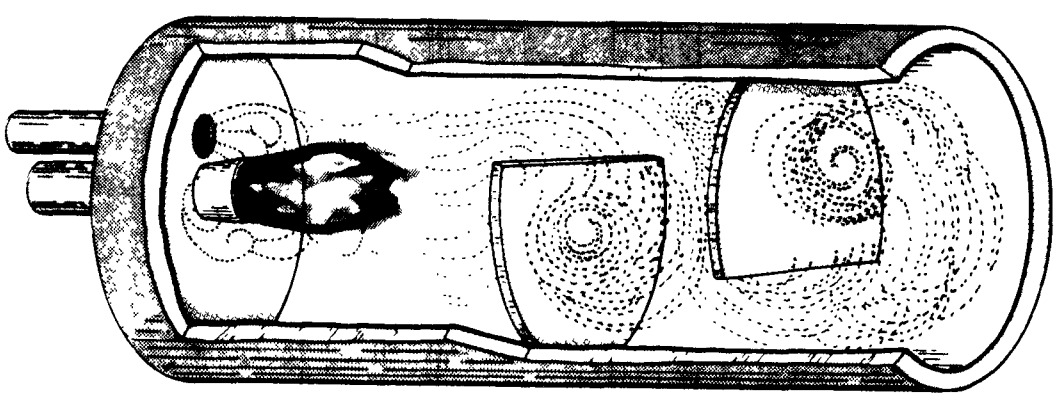


Figure 3-13. Bridge wall baffle.

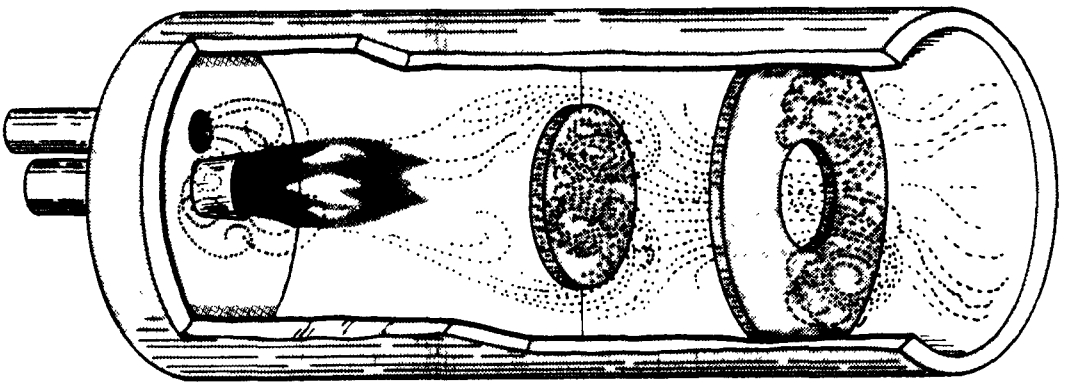


Figure 3-14. Ring and disc baffle.

Catalytic Oxidation

A *catalyst* is a substance which causes or speeds a chemical reaction without itself undergoing a change. In catalytic incineration, a waste gas is passed through a layer of catalyst known as the *catalyst bed*. The catalyst causes the oxidation reaction to proceed at a **faster rate** and **lower temperature** than is capable in thermal oxidation. Catalytic incinerators operating in a 370 to 480°C (700 to 900°F) range can achieve the same efficiency as a thermal incinerator operating between 700 and 820°C (1300 and 1500°F). This can result in a 40 to 60% fuel savings.

Catalytic reactions can be classified as either homogeneous or heterogeneous. Homogeneous reactions occur throughout the bulk of the catalyst, while heterogeneous reactions occur only on the surface of the catalytic material. In air pollution control applications, all reactions are heterogeneous. The oxidation reaction of the organic vapors occurs only on the surface of the catalyst. It should be noted that catalytic reactions produce the same end products (CO_2 and H_2O) and liberate the same heat of combustion as does thermal incineration.

A heterogeneous catalyst reaction proceeds through a series of five basic steps:

1. Organic compounds in the waste gas must first *diffuse* from the bulk of the vapor to the surface of the catalyst.
2. Organic compounds then *adsorb* onto the surface of the catalyst.
3. Organic compounds then *react* (oxidize).
4. New compounds then *desorb* after reacting.
5. New compounds then *diffuse* and mix back into the bulk of the exhaust air stream.

The most effective and commonly used catalysts for oxidation reactions come from the noble metals group. *Platinum* either alone or in combination with other noble metals is by far the most commonly used. Desirable characteristics of platinum are that it gives a high oxidation activity at low temperatures, is stable at high temperatures, and is chemically inert. *Palladium* is another noble metal which exhibits these properties and is sometimes used in catalytic incinerators.

Since catalytic oxidation is a surface reaction, the noble metal is coated onto the surface of a cheaper support material. The support material can be made of a ceramic or a metal such as *alumina, silica-alumina* or *nickel-chromium*. The support material is arranged in a matrix shape to provide: high geometric surface area; low pressure drop; uniform flow of the waste gas through the catalyst bed; and a structurally stable surface (EPA, 1972). Structures which provide these characteristics are pellets, a honeycomb matrix, or a mesh matrix. Figure 3-15 shows a typical honeycomb catalyst module which is the most common. The support material for these is usually ceramic, but can be metal.

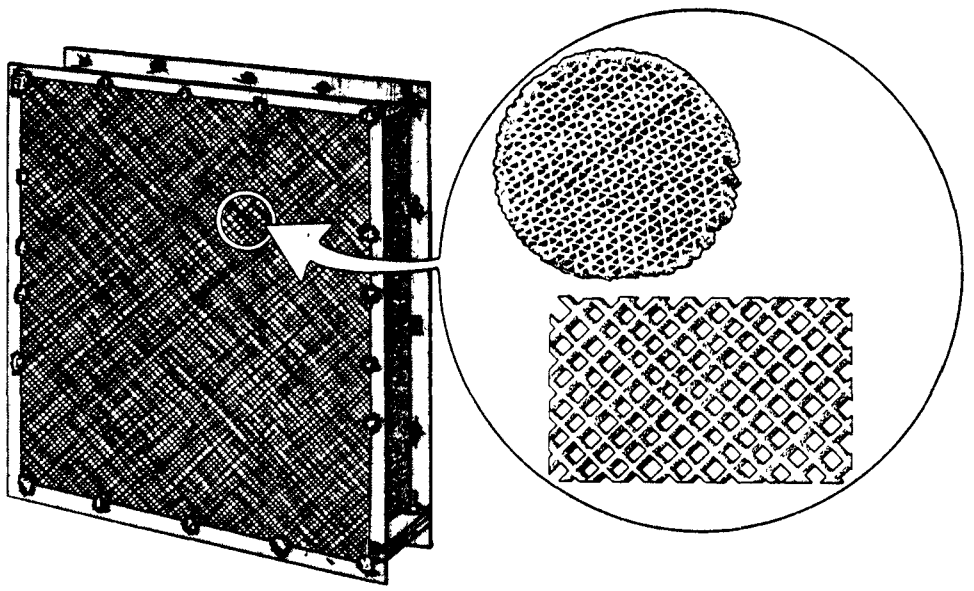


Figure 3-15. Typical honeycomb catalysts (metallic or ceramic).

A schematic of a catalytic incinerator is shown in Figure 3-16. Catalytic incinerators consist of a preheat section (burner area), where part of the waste gas is raised to operating temperature. The burners are the same as those used for thermal incineration, with the majority being distributed burners. The remaining portion of the waste gas is mixed with the hot products of combustion before passing over the catalyst bed. This ensures a homogenous waste gas and temperature mixture as it passes over the bed. After passing over the bed, the hot flue gases may be sent to a heat recovery system.

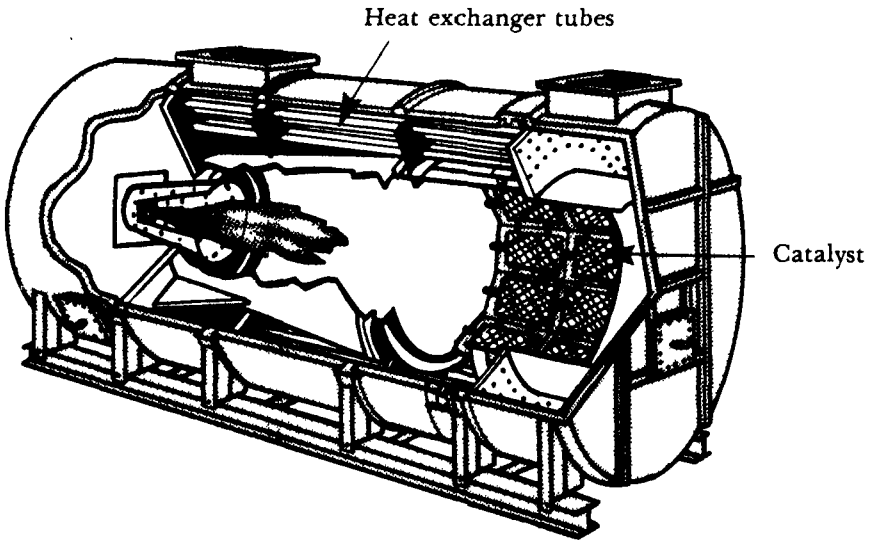


Figure 3-16. Catalytic incinerator.

Catalyst bed depth controls the pressure drop across the incinerator. Typically, the volume of catalyst required for 85 to 95% conversion of all organic compounds in 0.01 to 0.05 seconds is between 0.03 to 0.14 m^3 of catalyst per 1000 m/s of waste gas (0.5 to 2.0 ft^3 of catalyst per 1000 cfm of waste gas). As a general rule, the higher the molecular weight of the compound, the more readily it is oxidized. Normal pressure drops through catalytic incinerators are between 62 and 125 Pa (0.25 and 0.5 in. H_2O). If the exhaust does not have sufficient draft to overcome this pressure drop, a blower may have to be installed.

To ensure proper operation, the inlet and outlet temperatures to the catalyst section are monitored. This verifies that the temperature is sufficient to achieve the required conversion and also ensures protection of the catalyst from excessive temperatures. Since catalytic incinerators operate at lower temperatures than thermal units, less refractory brick and small chamber volumes can be used. This reduction in size of the device reduces installation costs.

Operating Limitations of Catalytic Incinerators

Catalytic incinerators usually cannot be used effectively on waste gas streams which contain particulate matter. Particulate matter that deposits on the surface of the catalyst prohibits the organic compounds from being adsorbed. Coating of the catalyst surface in this manner is referred to as *fouling* of the catalyst. Oil droplets can also foul the catalyst bed unless they are vaporized in the preheat section. By periodically cleaning and washing the catalyst, 90% of its activity can be restored (EPA, 1972). Maintenance of this type, however, adds significantly to the cost of operating the unit.

Certain metals can chemically combine with or alloy to the catalyst, thereby making it useless. Deactivation in this manner is called catalyst *poisoning*. Catalyst poisons can be divided into two categories: fast acting poisons which include

phosphorus, bismuth, arsenic, antimony, and mercury; and slow acting poisons which include zinc, lead and tin. Catalysts are more tolerant of the slow acting poisons, particularly at temperatures below 540°C (1000°F). At sufficiently high temperatures (>590°C), even copper and iron are capable of alloying to platinum, reducing its activity.

Sulfur and halogen compounds act as poisons, but not to the extent of the metals. They are essentially reaction inhibitors called *suppressants*. Their chemical interaction with the catalyst is reversible. Once the halogen or sulfur compound is removed, catalyst activity is restored to normal.

All catalysts deteriorate with normal use. Gradual loss of the catalyst material can occur from erosion, attrition, and vaporization. High temperatures can also accelerate catalyst deactivation. Loss of catalyst activity due to high temperature is known as *thermal aging*, and causes very rapid catalyst deterioration. Even short term temperatures above 820°C (1500°F) can cause a near total loss of catalyst activity. With proper monitoring of operating temperatures, a catalyst bed can be expected to last from three to five years before it must be replaced. The percent destruction of pollutants decreases with increasing thermal aging. Adjustment to operating conditions (temperature, residence time, etc.) may need to be made to ensure that the exhaust continues to meet emission limits.

Comparison of Thermal vs. Catalytic Incinerators

The major difference between thermal and catalytic incinerators is that complete combustion can be achieved at much lower temperatures in a catalytic incinerator. The reduced temperatures cut the cost of fuel usage by 40 to 60%. Operation at lower temperatures also decreases the construction costs. Lighter materials of construction can be used in catalytic units. This also makes installation easier and less costly. In terms of overall equipment costs, for small capacity units, up to 4.7 or 5.7 m^3/s (10,000 or 12,000 cfm), the purchase costs for either unit is essentially equal (EPA, 1972). For larger units that must be custom designed, thermal incinerators are usually less expensive, but this depends on the type of heat recovery system.

The main problem in catalytic incineration is the reduction or loss of catalyst activity. Loss of catalyst activity occurs due to fouling by particulate matter or suppression or poisoning due to other contaminants in the waste gas stream. In order to effectively use catalytic incineration, these contaminants must be removed from the waste gas stream. Removing these contaminants would require additional equipment which adds greatly to the cost of the system. Finally, all catalysts must periodically be replaced due to thermal aging.

Process Boilers Used as Incinerators

An alternative to installing a thermal or catalytic incinerator would be to combust the waste gases in an existing plant or process boiler. This would avoid the capital cost of new equipment and may help to reduce present fuel consumption. Process and plant boilers are normally designed to operate in excess of 980°C (1800°F)

with a flue gas residence time of 0.5 to 3.0 seconds. These conditions exceed those recommended for thermal incinerators. However, a number of additional conditions must be satisfied before waste gases can be properly disposed of in this manner.

The following criteria must be considered before a process boiler can be used as an incinerator (EPA, 1977):

1. The waste gases must be almost completely combustible. If solid particles are present in the waste gases, or formed by incomplete combustion, they can foul heat exchanger surfaces in the boiler, thus reducing boiler efficiency. Particulate matter may also cause boiler emissions to exceed applicable emission regulations. The costs of increased maintenance of the boiler and/or control of particulate matter may well exceed the purchase price of an incinerator.
2. The waste gas should, preferably, constitute only a small fraction of the air requirements of the boiler. If the volume of the waste gas is large, special attention must be paid to the oxygen balance, mixing, and continuation of the air flow in the boiler when the process is shut down.
3. The oxygen concentration of the contaminated gas stream should be close to that of air to avoid incomplete combustion. Incomplete combustion produces tars that coat heat exchanger surfaces, reducing boiler efficiency.
4. The boiler must operate at all times when incineration is required.
5. The waste gas must be free of compounds, such as halogenated hydrocarbons, that accelerate corrosion of the boiler.
6. Baffling may be required in the combustion chamber to ensure adequate mixing and combustion of the waste gas.
7. If the boiler-firing rate varies greatly, it may be necessary to install a small auxiliary boiler that will operate under steady load conditions.

To date, not many industries have been successful in using plant or process boilers as incinerators. Petroleum refineries, which have numerous waste gas streams and process boilers, are one of the few industries which have been able to incinerate waste gases in process boilers.

Heat Recovery Systems

Due to the rising cost and limited availability of fuels, heat recovery systems are becoming an integral part of most incinerators. Heat recovered from hot flue gas can be used in two ways to reduce energy consumption. One way is to preheat the cooler waste gas from the process entering the incinerator, referred to as *primary heat recovery*. The second way is to use the recovered heat in another process in the plant, such as a drying oven, referred to as *secondary heat recovery*. A number of systems are designed to recover heat in either one or in some cases both manners. To date, however, no single heat recovery system works universally well on all processes.

Specific terms are used in this chapter to describe heat recovery. *Waste gas* will always refer to the contaminant gas stream entering the incinerator. *Flue gas* refers to the hot products of combustion that are leaving the incinerator. Also, note that the terms *hot* and *cold* are relative, describing the direction that heat is being transferred. For example, hot flue gas may be used to heat the cold waste gas stream (even though the waste gas may be at 150°C).

Heat recovery devices are classified by their heat transfer method. Heat can be transferred either directly or indirectly. Mixing all or part of the hot flue gases with the waste gas is an example of *direct heat recovery*. *Indirect heat recovery* devices use heat absorbing or heat transferring materials to recover heat. Indirect heat recovery devices are either *recuperative* or *regenerative*. Recuperative heat recovery devices use heat transferring material, such as a tube or bank of tubes with hot gas on one side and cold gas on the other. A home hot water heater is an example of this system. Regenerative heat recovery devices use heat absorbing materials, such as stones or ceramic packing, to store heat from a hot gas stream then give it up to a cooler stream.

To be effective, heat recovery devices must take as much heat as possible from the hot flue gas and transfer it to the cooler waste gas stream. There are two temperature limiting factors to be aware of when trying to maximize this heat transfer. First, the degree of heat recovery is limited by the dew point of any gases in the hot flue gas stream. If the hot flue gas is cooled to below the dew point of any constituent (gas), corrosive acids can be formed. These acids can cause excessive damage to any heat recovery device or associated ductwork. One solution is to use construction materials resistant to acid corrosion. However, these are expensive and add greatly to system cost. The practical solution is to avoid excess cooling of the hot flue gas (below the dew point). For example, the dew point of sulfur-free fuels (i.e. natural gas) is approximately 60°C (140°F). Therefore, this temperature sets the lower limit of heat that is possible to be recovered.

Secondly, heat recovery can also be limited by the autoignition temperature of the waste gas stream. If the waste gas is preheated to its autoignition temperature an explosion or fire could occur. The autoignition temperatures of most industrial hydrocarbons fall between 370 and 540°C (700 and 1000°F). Heat recovery systems are capable of preheating waste gas streams above these temperatures. Unless other uses for the heat are found, safety restrictions will not permit utilizing all of the available heat energy.

Recuperative Heat Recovery Devices

Recuperative heat recovery devices are simply tubular heat exchangers (Figure 3-17). In these devices the cold waste gas usually passes on the inside of the tube, while the hot flue gas passes over the outside of the tubes. The effectiveness of these devices is limited by the thermal resistance to heat transfer through the tubes. Using circular tubes maximizes the amount of surface area available for heat transfer. These are currently the most frequently used heat recovery devices since they contain no moving parts.

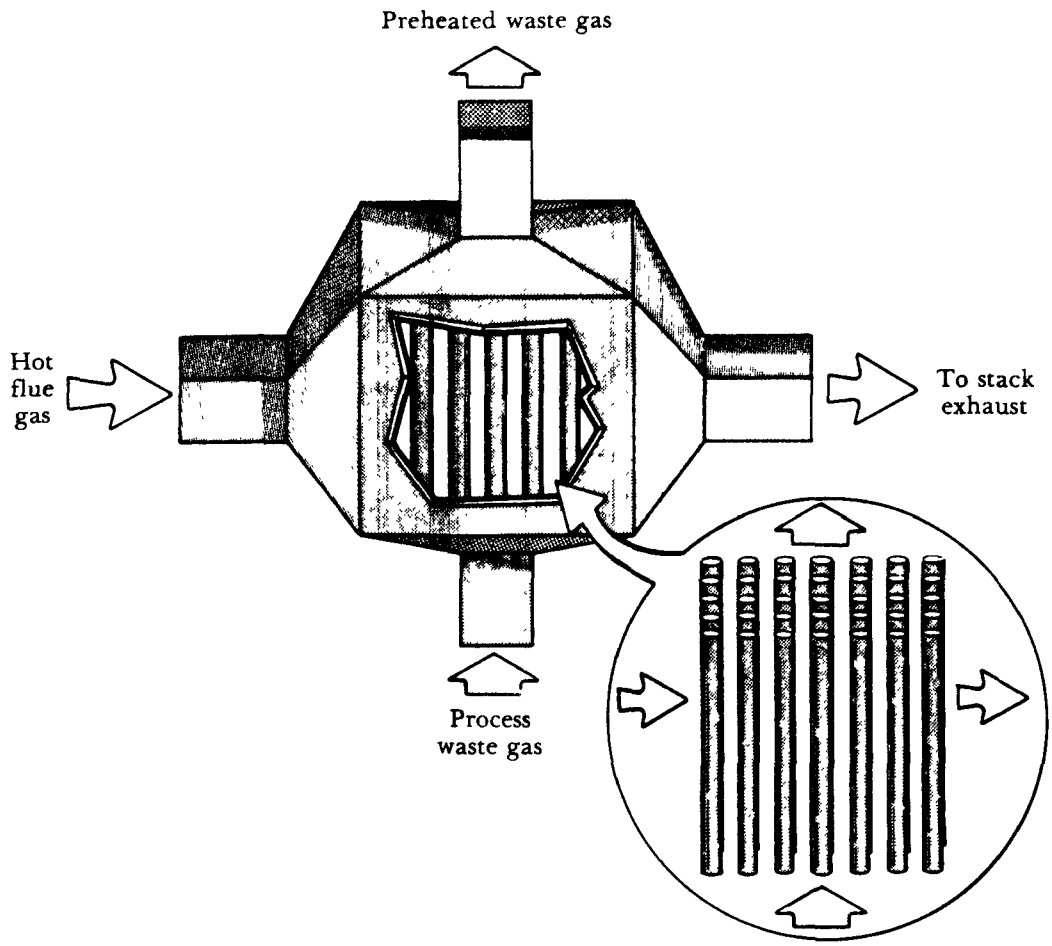


Figure 3-17. Recuperative heat exchanger (shell-and-tube).

In comparing different heat recovery devices, it is mandatory to have a common basis for judging performance. One method used to judge the performance of heat recovery systems is by the *heat transfer effectiveness* (E) measured by (EPA, 1972):

(Eq. 3-10)

$$E = \frac{q}{q_{max}} = \frac{\dot{m}\bar{C}_p(T_{waste\ gas\ leaving} - T_{waste\ gas\ entering})}{\dot{m}\bar{C}_p(T_{flue\ gas\ entering} - T_{flue\ gas\ leaving})}$$

Heat transfer effectiveness is measured as the *ratio* of the *amount of heat actually recovered (q) to the maximum amount of heat (q_{max}) that potentially could be recovered*. Examining the right side of Equation 3-10, it can normally be assumed that the mass flow ($\dot{m}$) and specific heat ($\bar{C}_p$) of both the waste gas and flue streams are equivalent. For this case, the heat recovered depends only on the temperature difference of the streams entering and leaving the device.

For single pass, tubular heat exchangers as shown in Figure 3-17, the effectiveness is usually between 40 and 50% (EPA, 1972). By arranging multiple passes of either hot or cold streams, heat exchanger effectiveness can be increased. By passing the cold waste gas over the tubes twice, the effectiveness can be increased to approximately 60%; while for a three pass system, the effectiveness can be raised

to 70% (Mueller, 1977). Figure 3-18 shows a typical double pass heat exchanger system. Although the effectiveness increases by adding extra passes or stages, it also adds to the total pressure drop. The pressure drop across a typical tubular heat exchanger is between 1 to 2 kPa per stage (4 to 8 inches of H_2O per stage).

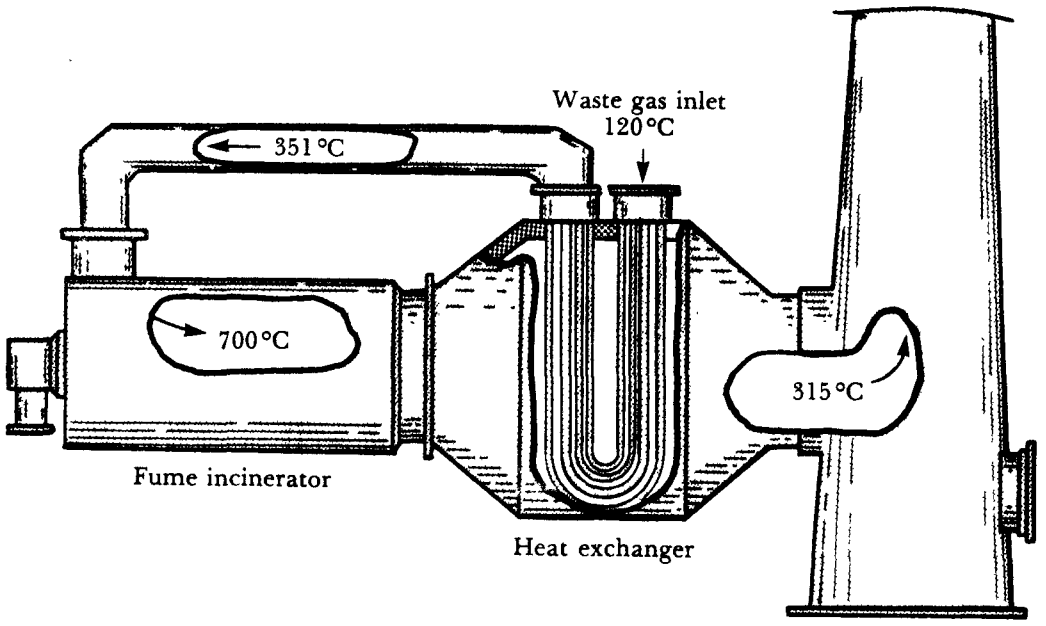


Figure 3-18. Double pass recuperative heat exchanger.

The main disadvantage of the recuperative heat recovery devices are that they are easily fouled by heavy smoke or sticky particulate matter. Fouling occurs when material coats the tubes, reducing the rate of heat transfer. If fouling is a problem, frequent cleaning of the tubes may be required. When contaminants are present, the contaminated stream is usually passed through the inside of the tubes. Cleaning the inside of a tube is much easier than cleaning the outside. Materials of construction for heat exchangers must be able to withstand constant operation at high temperatures. Heavy duty, high quality materials of construction must be used or thermal expansion can cause warpage and thus cause welds to separate. If this occurs, contaminants can bypass the incinerator and be exhausted directly to the atmosphere. In addition, the waste gas must not be preheated above its auto-ignition temperature or cooled below its dew point to avoid damage to the device.

Regenerative Heat Recovery Devices

Regenerative devices use absorbing material that stores heat from hot flue gas for later use. These devices operate on a cycle, part of the time the absorbing material is being heated, then that portion of the bed is moved to regeneration or heat recovery cycle. Regenerative devices can recover a greater percent of heat than recuperative devices. Regenerative devices have only one gas-solid heat resistance

boundary (between hot gas and absorbing material) to pass through. In recuperative devices heat must pass through two transfer boundaries (the inside and outside of the tube). The two most prevalent regenerative devices are the *rotary* exchanger and the *packed bed.*

The rotary or wheel heat exchanger consists of a wheel with a large surface area (Figure 3-19). Hot flue gas flows through one portion of the wheel while the cool waste gas flows in the remaining portion. The surface of the wheel is constructed of corrugated sheet metal plates. The plates are arranged in a honeycomb matrix form to give both maximum heat transfer and air flow between the plates. Rotating the wheel allows one of its sides to heat up, then this heat is transferred to the cool side. Rotary wheels of this type are capable of achieving 80% effectiveness and have very low pressure drops (Bethea, 1978).

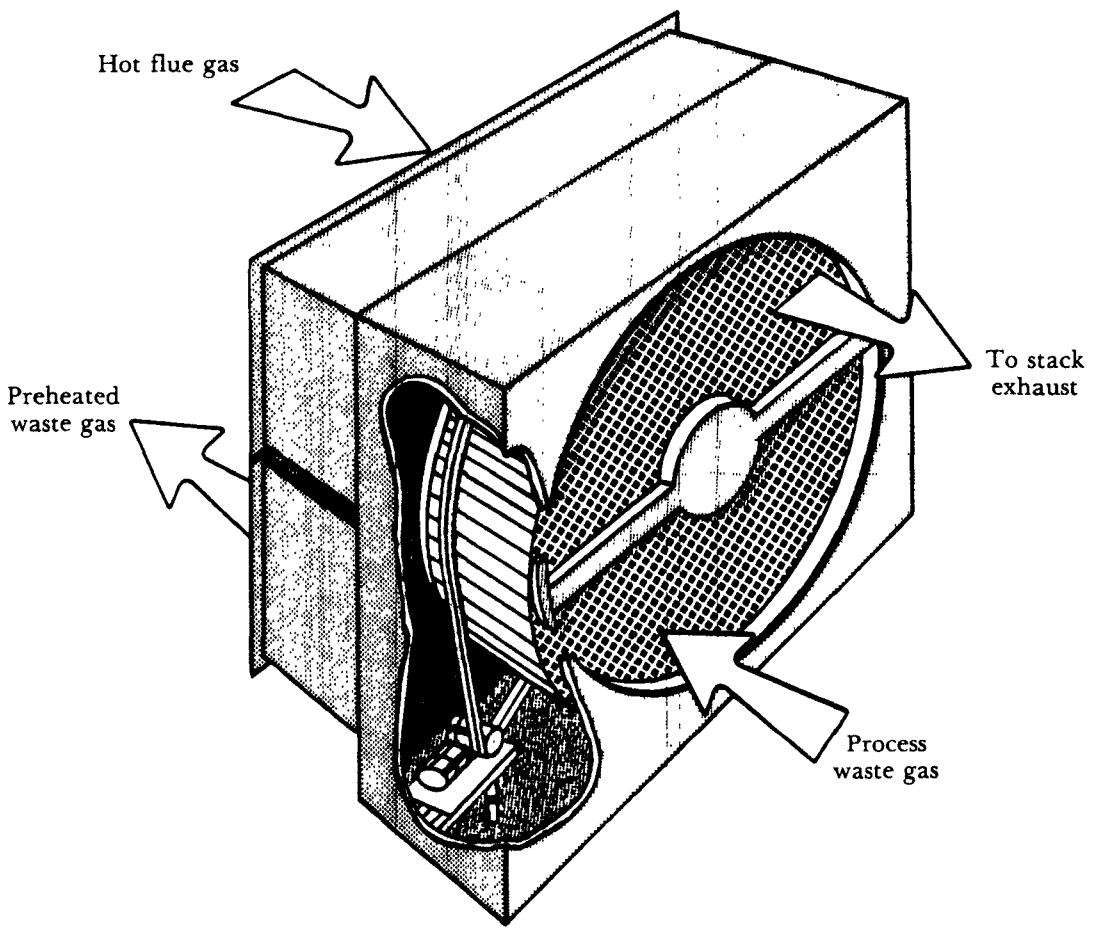
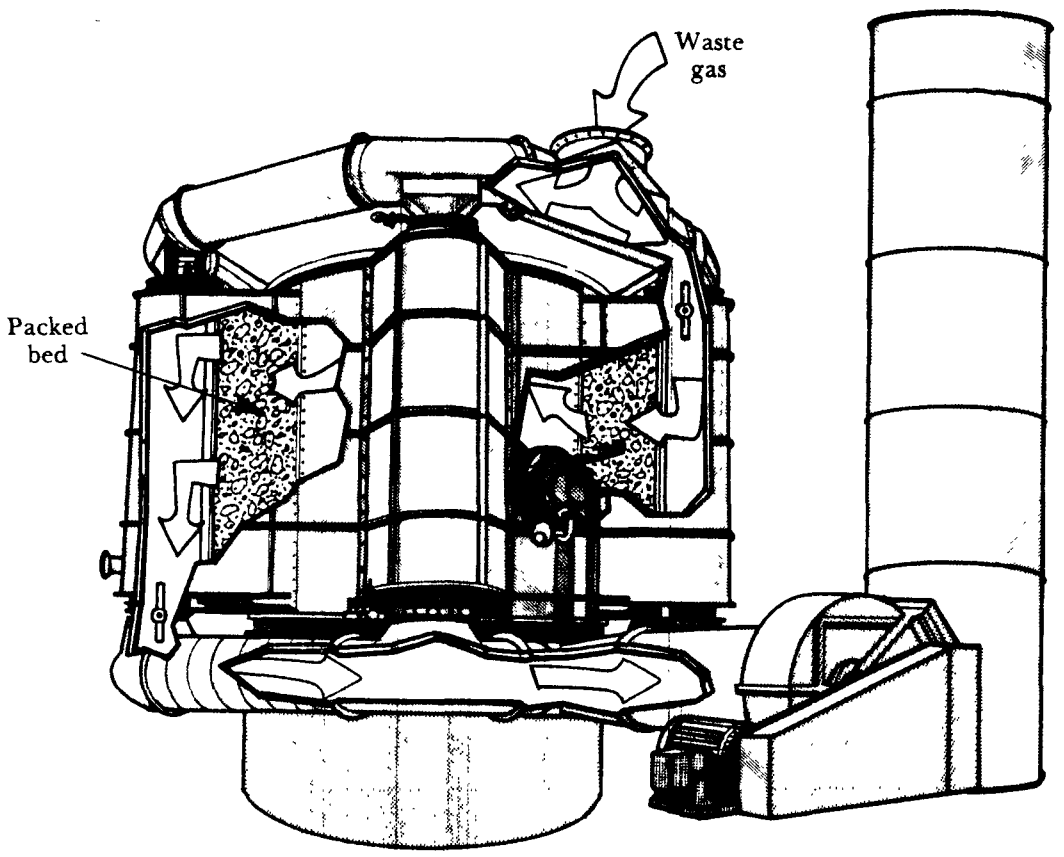


Figure 3-19. Regenerative heat wheel.

Disadvantages associated with use of the heat wheel are clogging and thermal wear. The small passageways between the plate of the wheel can become plugged by heavy smoke or sticky particulate matter in the waste gas stream. Isolating seals control air leakage between the hot and cold sections. These seals are susceptible to wear and warpage due to high temperatures and constant movement. Seal failure

allows contaminants to bypass the incinerator. Leakage from these seals can be minimized in some instances by arranging blowers to provide a higher pressure on the flue gas side of the wheel so that any leaks will be directed into the waste gas stream.

The other regenerative heat recovery device is a packed bed. One design consists of a number of towers packed with a heat absorbing material. The towers surround a central combustion chamber. A typical regenerative heat recovery device manufactured by REECO, Inc. is shown in Figure 3-20. Initially the waste gas passes into the combustion chamber where it is incinerated. Hot flue gas passes from the combustion chamber through a packed bed, heating the bed. Once a packed bed is heated, flow switching valves direct the cold waste gas through the hot bed which preheats it to combustion temperatures. The cycle continues with the hot flue gas heating another packed section. At steady state conditions it is possible to operate the REECO heat recovery device without any auxilary fuel. This capability depends on the concentration and heating value of the combustible compounds in the waste gas.



Source: REECO, Inc.

Figure 3-20. Regenerative heat recovery device.

The REECO device has been marketed only for the past few years. Although there is not a lot of data on these systems, the installed units seem to be operating well. These are large units, approximately 10 meters high by 7 meters in diameter. Therefore, they are economical only for large industrial applications. They are designed for high effectiveness, between 70 and 90%. Due to their unique design, they are self-cleaning and they also eliminate the hazard of autoignition in preheating the waste gas.

Economic Considerations of Heat Recovery

A recent EPA study of ten heat recovery systems noted that the payback time for these systems ranged from a low of 0.8 to a high of 10.9 years (EPA, 1978). Most of the data was grouped between one and two years. With the increasing cost of fuel, the financial benefit of heat recovery devices will increase until it is uneconomical to operate without them.

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Chapter 4
Absorption

Introduction

Absorption refers to the transfer of a gaseous component from the gas phase to a liquid phase. More specifically, in air pollution control, absorption involves the removal of objectionable gaseous contaminants from a process stream by dissolving them in a liquid.

Some common terms used when discussing the absorption process follow:

- *Absorbent*: the liquid, usually water, into which the contaminant is absorbed.
- *Absorbate* or *solute*: the gaseous contaminant being absorbed, such as SO_2, H_2S, etc.
- *Carrier gas*: the inert portion of the gas stream, usually air, from which the contaminant is to be removed.
- *Interface*: the area where the gas phase and the absorbent contact each other.
- *Solubility*: the capability of a gas to be dissolved in a liquid.

Absorption is a *mass transfer operation*. Mass transfer can be compared to heat transfer in that both occur because a system is trying to reach equilibrium conditions. For example, in heat transfer, if a hot slab of metal is placed on top of a cold slab, heat will be transferred from the hot slab to the cold slab until both are at the same temperature (equilibrium). In heat transfer, the process continues as long as a temperature differential or gradient exists. In absorption **mass** instead of heat is transferred, and instead of occurring due to a temperature change, absorption occurs as a result of a **concentration** difference. Absorption continues as long as a concentration differential exists between the absorbent and the gas from which the contaminant is being removed. In absorption, equilibrium is not as easily defined as in heat transfer, since the concentration difference depends on the solubility of the solute.

Absorption devices used to remove gaseous contaminants are referred to as absorbers or wet scrubbers. Wet scrubbers are also used to remove particulate matter from gas streams. Wet scrubbers usually cannot be operated to optimize **simultaneous** removal of both gases and particulate matter. In designing absorbers for gaseous emissions, optimum mass transfer can be accomplished by:

- providing a large interfacial contact area.
- providing good mixing between gas and liquid phases.
- allowing sufficient residence or contact time between the phases.
- ensuring a high degree of solubility of the contaminant in the absorbent.

Mechanism of Absorption

To remove a gaseous contaminant by absorption, the contaminant-laden exhaust stream must be passed through (contacted with) a liquid. Figure 4-1 illustrates the steps involved in the absorption mechanism. In the first step of the absorption process, the solute diffuses from the bulk area of the gas phase to the gas-liquid interface. In the second step, the solute transfers across the interface to the liquid phase. In the third step, the solute diffuses into the bulk area of the liquid, making room for additional gas molecules to be absorbed.

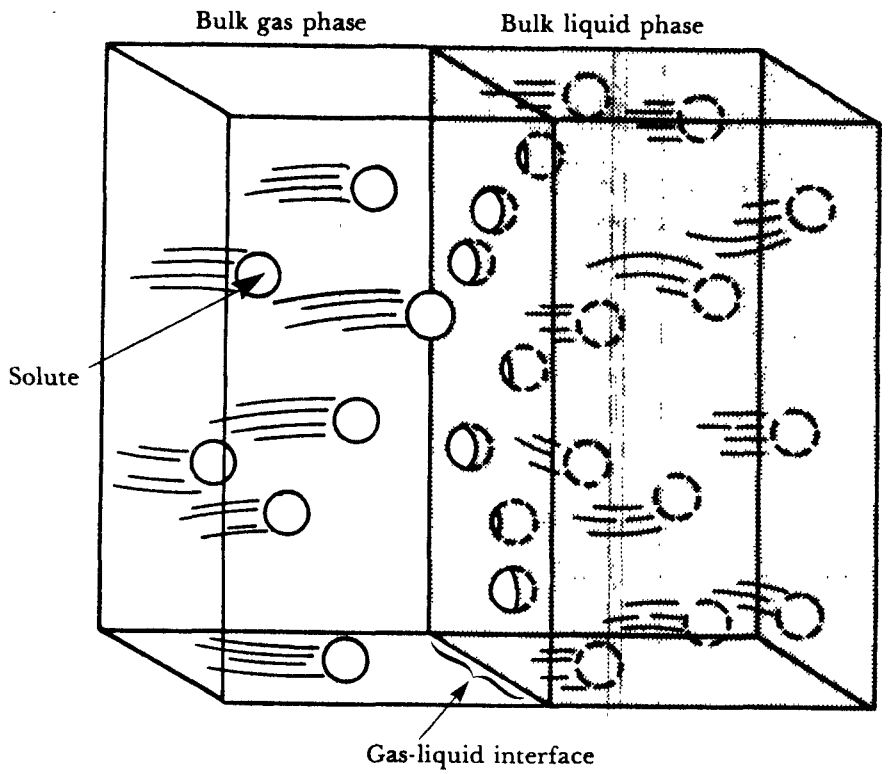


Figure 4-1. Absorption process.

The purpose of analyzing these three steps is to determine which variables control the process. The most efficient system can be designed by knowing these variables. It is assumed that once the solute arrives at the interface area, transfer across it occurs instantaneously. This second step in the absorption mechanism is extremely rapid. Therefore, it does not need to be considered when deriving absorption efficiency equations. The rate of mass transfer (absorption) is dependent upon the diffusion rate in either the gas phase or liquid phase.

Two terms used to describe the mass transfer rate are: *gas phase controlled* absorption and *liquid phase controlled* absorption. Each mechanism depends on the *rate* of diffusion in both phases and upon the *solubility* of the pollutant in the liquid.

The diffusion rate of a gaseous pollutant molecule through a gas is always faster than its diffusion rate through a liquid because gas molecules in a gas are further apart than are liquid molecules in a liquid. The pollutant molecule will move faster through the gas phase than through the liquid phase since in a gas, obstructions such as other molecules are farther apart than in a liquid. However, the mass transfer rate depends primarily upon the solubility of the pollutant in the liquid.

Consider two gaseous pollutants, A and B, that are approximately equal in size. Assume that A is not very soluble in the liquid while B is very soluble in the liquid. Assume also that both A and B will move through the gas phase at the same rate. The rates at which the pollutants will be absorbed in the liquid phase are different because B is very soluble while A is not. A is absorbed very slowly and B is absorbed very quickly.

The transfer rate of A (slowly absorbed) is greatly affected by its absorption and diffusion in the liquid phase and is therefore referred to as liquid phase controlled. The transfer rate of B (quickly absorbed) is also greatly affected by its rate of movement through the liquid phase but is absorbed so rapidly that it doesn't influence the overall transfer rate. The mass transfer rate for B depends only on how fast it diffuses through the gas phase towards the gas-liquid interface and is referred to as gas phase controlled.

The gas phase controlled systems absorb pollutants more readily than do the liquid phase controlled systems. Thus, absorption systems used in the field of air pollution control are usually designed to be gas phase controlled.

The previously mentioned three step mechanism of absorption occurs on two levels: a micro or molecular level and a macro or bulk level. Molecular diffusion of individual molecules occurs due to a concentration difference because all systems try to achieve an equilibrium state. Molecules migrate from areas of high concentrations to areas of low concentrations. Macro or bulk diffusion occurs as a mass of liquid or gas moves to or from the interface. In an absorber, bulk diffusion is accomplished by turbulent mixing of the gas and liquid phases. To be efficient, an absorber must provide both turbulent mixing of the gas and liquid phases and sufficient residence time to allow pollutant molecules to be absorbed.

Solubility

A very important factor affecting the amount of a contaminant that can be absorbed is the solubility of the contaminant. Solubility is a function of both the temperature and to a lesser extent the pressure of the system. As temperature increases, the amount of gas that can be absorbed by a liquid decreases. From the ideal gas law: as temperature increases, the volume of a gas also increases; therefore, at the higher temperature less gas is absorbed due to the increased volume it occupies. Pressure affects the solubility of a gas in the opposite manner. By increasing the pressure of a system the amount of gas absorbed generally increases.

The solubility of a specific gas in a given liquid is defined at a designated temperature and pressure. Table 4-1 presents data on the solubility of SO_2 gas in water at 101 kPa or 1 atm and various temperatures. In determining solubility data, the partial pressure (in mm Hg) is measured with the concentration (in grams of solute per 100 g of liquid) of the solute in the liquid. The data in Table 4-1 was taken from *The International Critical Tables*, a good source of information concerning gas-liquid systems.

Table 4-1. Partial pressure of SO_2 in aqueous solution, mm Hg.

Grams SO_2 per 100 g H_2O	10°C	20°C	30°C	40°C	50°C	60°C	70°C
0.0	—	—	—	—	—	—	—
0.5	21	29	42	60	83	111	144
1.0	42	59	85	120	164	217	281
1.5	64	90	129	181	247	328	426
2.0	86	123	176	245	333	444	581
2.5	108	157	224	311	421	562	739
3.0	130	191	273	378	511	682	897
3.5	153	227	324	447	603	804	—
4.0	176	264	376	518	698	—	—
4.5	199	300	428	588	793	—	—
5.0	223	338	482	661	—	—	—

Solubility data is obtained at equilibrium conditions. This involves putting measured amounts of a gas and a liquid into a closed vessel and allowing it to sit for a period of time. Eventually the amount of gas that is being absorbed into the liquid will equal the amount which is coming out of the solution. At this point, there is no net transfer of mass to either phase and the concentration of the gas in both the gaseous and liquid phases remains constant. The gas-liquid system is at equilibrium.

Equilibrium conditions are important in operating an absorption tower. If equilibrium were to be reached in the actual operation of an absorption tower, the collection efficiency would fall to zero at that point since no net mass transfer would occur. The equilibrium concentration, therefore, limits the amount of solute that can be removed by absorption. The most common method of analyzing solubility data is to use an equilibrium diagram. An equilibrium diagram is a plot of the **mole fraction of solute in the liquid phase**, denoted as x, versus the **mole fraction of solute in the gas phase**, denoted as y. Equilibrium lines for the SO_2 and water system given in Table 4-1 are plotted in Figure 4-2. Figure 4-2 also illustrates the temperature dependence of the absorption process. At a constant mole fraction of solute in the gas (y), the mole fraction of SO_2 in the liquid (x) increases as the temperature decreases.

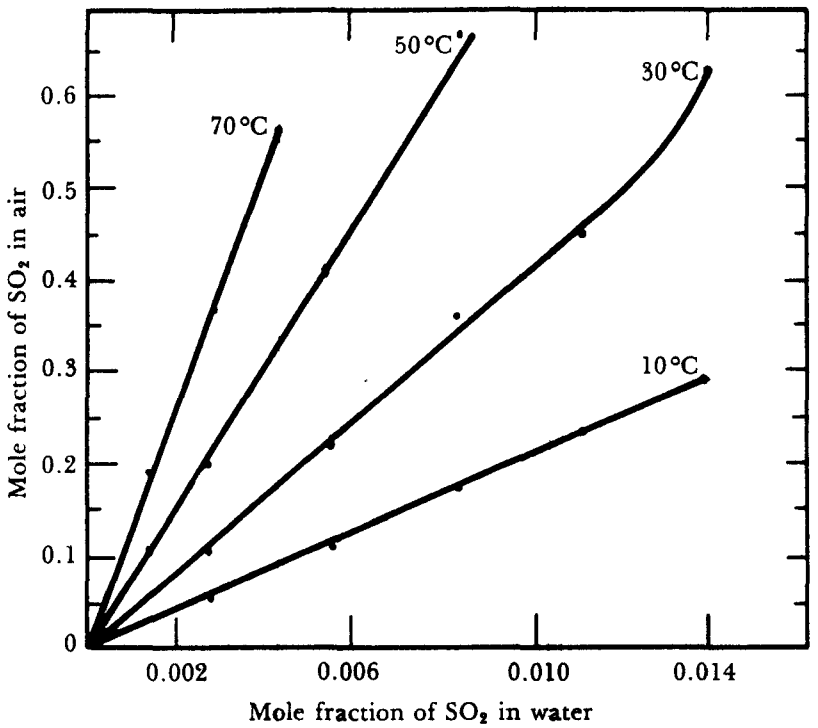


Figure 4-2. Equilibrium lines for SO_2—H_2O system.

Under certain conditions, *Henry's Law* may also be used to express equilibrium solubility of gas-liquid systems. Henry's Law is expressed as:

(Eq. 4-1) $$p^* = \mathscr{H} x$$

Where: p^* = partial pressure of solute at equilibrium, Pa
x = mole fraction of solute in the liquid
$\mathscr{H}$ = Henry's Law constant, Pa/mole fraction

Henry's Law can be written in a more useful form by dividing both sides of Equation 4-1 by the total pressure, P_T, of the system. The left side of the equation becomes the partial pressure divided by the total pressure which equals the mole fraction in the gas phase, y^*. Equation 4-1 now becomes:

(Eq. 4-2) $$y^* = \mathscr{H}' x$$

Where: y^* = mole fraction in gas phase in equilibrium with liquid
$\mathscr{H}'$ = Henry's Law constant, mole fraction in vapor/mole fraction in liquid
Note: $\mathscr{H}'$ is now dependent on the total pressure

Equation 4-2 is the equation of a straight line, where the slope (m) is equal to $\mathscr{H}'$. Henry's Law can be used to predict solubility **only** when the equilibrium line is **straight**. Equilibrium lines are usually straight when the solute concentrations are very dilute. In air pollution control applications this is usually the case. For example, an exhaust stream that contains a 1000 ppm SO_2 concentration corresponds to a mole fraction of SO_2 in the gas phase of only 0.001. Figure 4-2 demonstrates that the equilibrium lines are still straight at this low concentration of SO_2.

Another restriction on using Henry's Law is that it does not hold true for gases that react or dissociate upon dissolution. If this happens, the gas no longer exists as a simple molecule. For example, scrubbing HF or HCl gases with water causes both compounds to dissociate in solution. In these cases, the equilibrium lines are curved rather than straight. Data on systems that exhibit curved equilibrium lines must be obtained from experiments.

Henry's Law constants for the solubility of several gases in water are listed in Table 4-2. The units of Henry's Law constants are atm/mole fraction. The smaller the constant, the more soluble the gas. Table 4-2 demonstrates that SO_2 is approximately 100 times more soluble in water than CO_2 is in water. The following example illustrates how to develop an equilibrium diagram from solubility data.

Table 4-2. Henry's Law constants for gases in H_2O*.

Gas	20°C	30°C
N_2	80.4	92.4
CO	53.6	62.0
H_2S	48.3	60.9
O_2	40.1	47.5
NO	26.4	31.0
CO_2	1.42	1.86
SO_2	0.014	0.016

*Expressed in $\mathscr{H}' \times 10^{-3}$, atm/mole fraction.

Source: Marchello, 1976.

Example 4-1

Given the following data for the solubility of SO_2 in pure water at 303 K (30 °C) and 101.3 kPa (760 mm Hg), plot the equilibrium diagram and determine if Henry's Law applies.

Equilibrium data

c_{SO_2} (g of SO_2 per 100 g of H_2O)	$p^*_{SO_2}$ (partial pressure of SO_2)
0.5	6 kPa (42 mm Hg)
1.0	11.6 kPa (85 mm Hg)
1.5	18.3 kPa (129 mm Hg)
2.0	24.3 kPa (176 mm Hg)
2.5	30.0 kPa (224 mm Hg)
3.0	36.4 kPa (273 mm Hg)

Solution:

1. The data must first be converted to mole fraction units.

   The mole fraction in the gas phase, y*, is obtained by dividing the partial pressure of SO_2 by the total pressure of the system:

$$y^* = \frac{p^*_{SO_2}}{P_T} = \frac{6 \text{ kPa}}{101.3 \text{ kPa}} = 0.06$$

   The mole fraction in the liquid phase, x, is obtained by dividing the moles of SO_2 in solution by the total moles of liquid:

$$x = \frac{\text{moles } SO_2 \text{ in solution}}{\text{moles } SO_2 \text{ in solution} + \text{moles of } H_2O}$$

   moles of SO_2 in solution $= c_{SO_2}/64$ g SO_2 per mole

   moles of H_2O = 100 gm of H_2O/18 g H_2O per mole = 5.55

$$x = \frac{c_{SO_2}/64}{c_{SO_2}/64 + 5.55} = \frac{\frac{0.5}{64}}{\frac{0.5}{64} + 5.55} = 0.0014$$

2. Completing the following for Example 4-1:

$c_{SO_2} = \frac{\text{g of } SO_2}{100 \text{ g } H_2O}$	p* (kPa)	$y^* = p/101.3$	$x = \frac{c_{SO_2}/64}{c_{SO_2}/64 + 5.55}$
0.5	6.0	0.060	0.0014
1.0	11.6	0.115	0.0028
1.5	18.3	0.180	0.0042
2.0	24.3	0.239	0.0056
2.5	30.0	0.298	0.0070
3.0	36.4	0.359	0.0084

The above data is plotted in Figure 4-3. Henry's Law applies in the given concentration range with Henry's Law constant equal to 42.7 mole fraction SO_2 in air/mole fraction SO_2 in water.

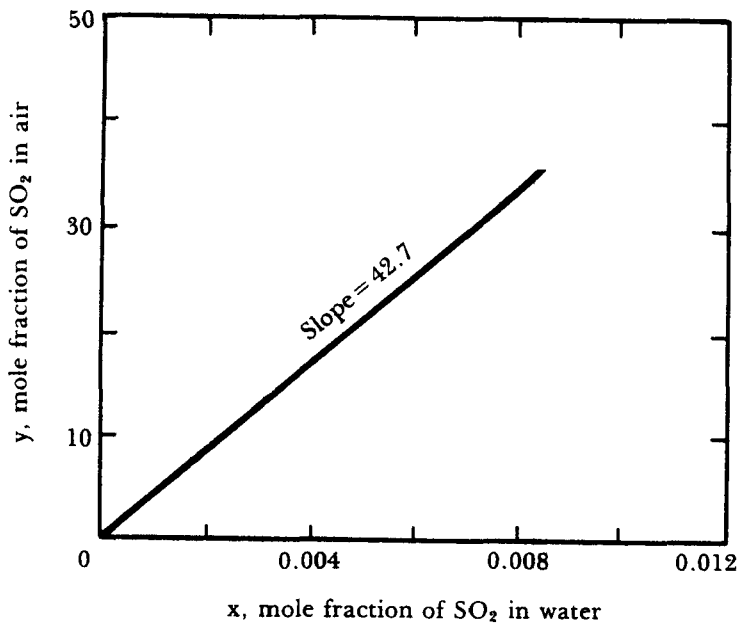


Figure 4-3. Equilibrium diagram for SO_2—H_2O system.

Absorption Design Theory

The first step in designing an air pollution control device is to develop a mathematical expression describing the observed phenomenon. A valid mathematical expression describing absorber performance makes it possible to: determine the proper absorber size for a given set of conditions, and predict how a change in operating conditions affects absorber performance. There are a number of theories or models which attempt to analytically describe the absorption

mechanism. However, in practice, none of these analytical expressions can solely be used for design calculations. Experimental or empirical data must also be used to obtain reliable results.

The basic model for describing the absorption process is the *two film* or *double resistance* theory which was first proposed by Whitman in 1923. The model starts with the three step mechanism of absorption, previously discussed and illustrated in Figure 4-1. From this mechanism, the rate of mass transfer was shown to be dependent on the rate of diffusion of a molecule in either the gas or liquid phase. The two film model starts by assuming that the gas and liquid phases are in turbulent contact with each other, separated by an interface area where they meet. This assumption may be correct; but no mathematical expressions adequately describe the diffusion (transport) of a molecule through both phases in turbulent motion. Therefore, the model proposes that a mass transfer zone exists to include a small portion (film) of the gas and liquid phases on either side of the interface. The mass transfer zone is comprised of two films, a gas film and a liquid film on their respective sides of the interface. These films are assumed to flow in a laminar or streamline motion, for which mathematical expressions that describe diffusion do exist. This concept of the two film theory is illustrated in Figure 4-4.

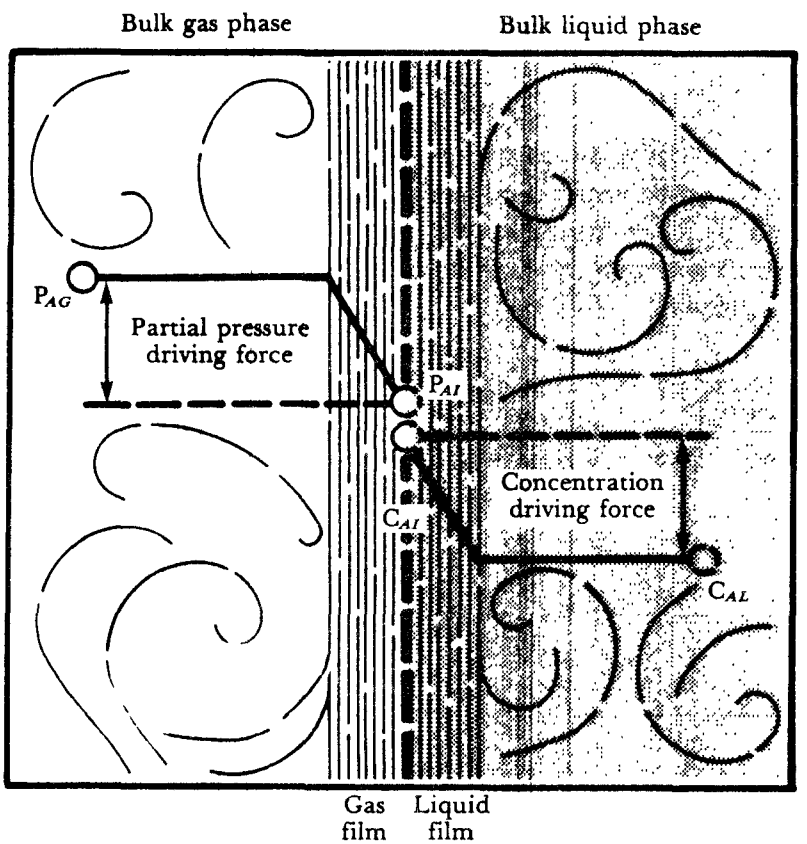


Figure 4-4. Visualization of two film theory.

According to the two film theory, for a molecule of substance A to be absorbed it must proceed through a series of five steps. The molecule must diffuse:

1. from the bulk gas phase to the gas film,
2. through the gas film
3. across the interface,
4. through the liquid film,
5. and finally mix into the bulk liquid.

The theory assumes that there is complete mixing in both the gas and liquid bulk phases and that the interface is at equilibrium with respect to molecules transferring in or out. This implies that all resistance to diffusion occurs when the molecule is diffusing through the gas and liquid films, hence the name *double resistance* theory. The concentration in the gas phase changes from p_{AG} in the bulk gas to p_{AI} at the interface.

A gas concentration is expressed by its partial pressure. Similarly, the concentration in the liquid changes from c_{AI} at the interface to c_{AL} in the bulk phase as mass transfer occurs. The rate of mass transfer is then equal to the amount of molecule A transferred times the resistance molecule A encounters in diffusing through the films:

(Eq. 4-3) $$N_A = k_g(p_{AG} - p_{AI})$$

(Eq. 4-4) $$N_A = k_\ell(c_{AI} - c_{AL})$$

Where: N_A = rate of transfer of component A, g mole/hr•m² (lb mole/hr•ft²)
k_g = mass transfer coefficient for gas film, g mole/hr•m²•Pa (lb mole/hr•ft³•atm)
k_ℓ = mass transfer coefficient for liquid film, g mole/hr•m²•Pa (lb mole/hr•ft²•atm)

The mass transfer coefficients, k_g and k_ℓ, represent the flow resistance the solute encounters in diffusing through each film respectively (Figure 4-5). An analogy is the resistance electricity encounters as it flows through a circuit.

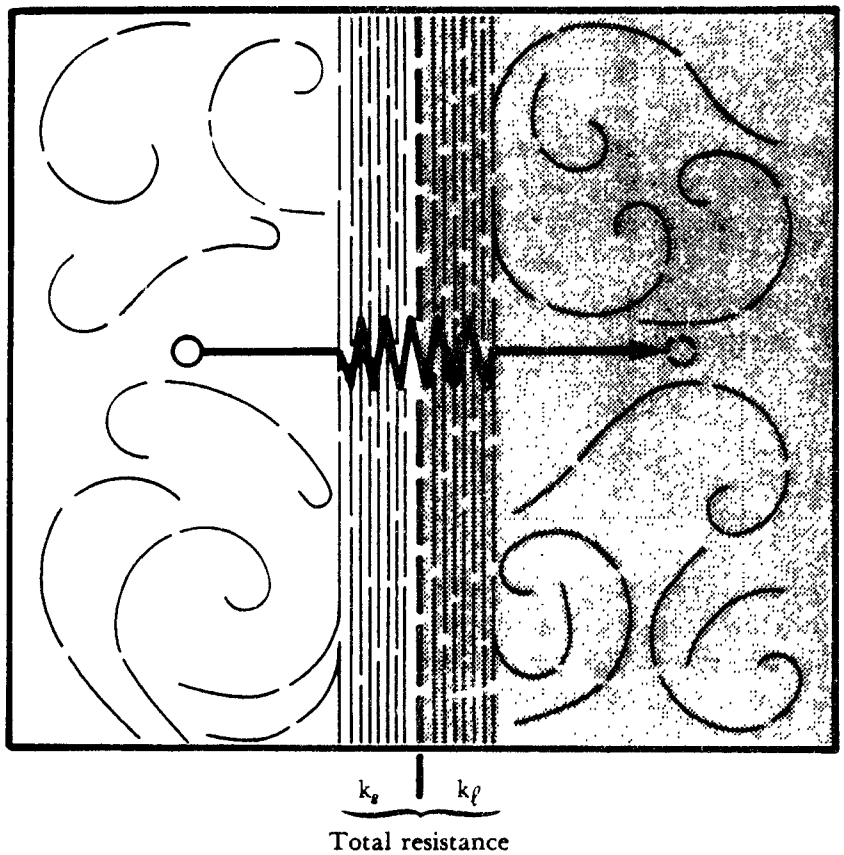


Figure 4-5. Resistance to motion encountered by a molecule being absorbed.

Equations 4-3 and 4-4 define the general case of absorption and are applicable to both curved and straight equilibrium lines. In practice, Equations 4-3 and 4-4 are difficult to use, since it is impossible to measure the interface concentrations p_{AI} and c_{AI}. The interface is a fictitious quantity used in the model to represent an observed phenomenon. The interface concentrations can be avoided by defining the mass transfer system at equilibrium conditions and combining the individual film resistances into an overall resistance. If the equilibrium line is straight, the rate of absorption is given by:

(Eq. 4-5) $$N_A = K_{OG}\,(p_{AG} - p_A^*)$$

(Eq. 4-6) $$N_A = K_{OL}\,(c_A^* - c_{AL})$$

Where: p_A^* = equilibrium partial pressure of solute A at operating conditions
c_A^* = equilibrium concentration of solute A at operating conditions
K_{OG} = overall mass transfer coefficient based on gas phase, g mole/hr•m²•Pa (lb mole/hr•ft²•atm)
K_{OL} = overall mass transfer coefficient based on liquid phase, g mole/hr•m²•Pa (lb mole/hr•ft²•atm)

An important fact to note concerning Equations 4-5 and 4-6 is that they impose an upper limit on the amount of solute that can be absorbed. The rate of mass transfer is dependent on the concentration departure from equilibrium in either the gas ($p_{AG} - p_A^*$) or liquid ($c_A^* - c_{AL}$) phase. The larger these concentration differences are, the greater the rate of mass transfer. If equilibrium is ever reached ($p_{AG} = p_A^*$ or $c_{AL} = c_A^*$) absorption stops and no net transfer occurs. Thus, the equilibrium concentrations determine the maximum amount of solute that is absorbed.

At equilibrium the overall mass transfer coefficients are related to the individual mass transfer coefficients by:

(Eq. 4-7)
$$\frac{1}{K_{OG}} = \frac{1}{k_g} + \frac{\mathscr{H}}{k_\ell}$$

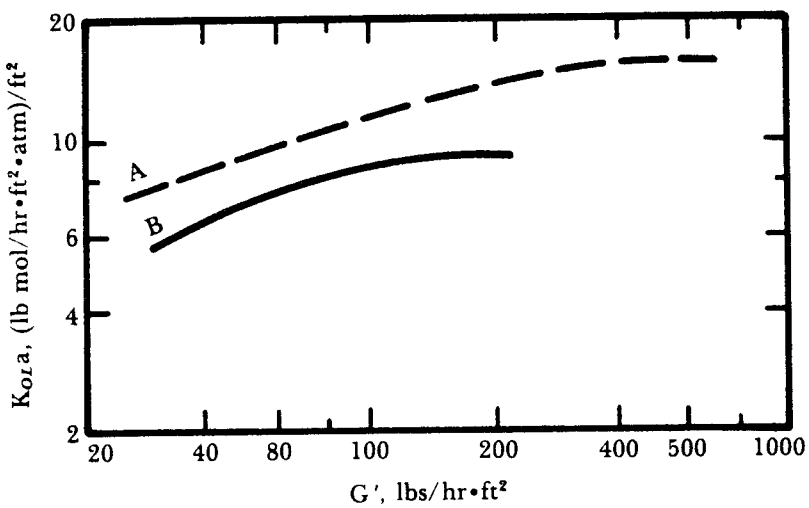
(Eq. 4-8)
$$\frac{1}{K_{OL}} = \frac{1}{k_\ell} + \frac{1}{\mathscr{H} k_g}$$

$\mathscr{H}$ is Henry's Law constant (the slope of the equilibrium). Equations 4-7 and 4-8 are useful in determining which phase controls the rate of absorption. From Equation 4-7, if $\mathscr{H}$ is very small (which means the gas is very soluble in the liquid) then $K_{OG} \approx k_g$, and absorption is said to be gas film controlled. The major resistance to mass transfer is in the gas phase. Conversely, if a gas has limited solubility $\mathscr{H}$ is large, and Equation 4-8 reduces to $K_{OL} \approx k_\ell$. The mass transfer rate is liquid film controlled and depends on the solute's dispersion rate in the liquid phase. Most systems in the air pollution control field are gas phase controlled since the liquid is chosen so that the solute will have a high degree of solubility.

The discussion so far has been based on the two film theory of absorption. There are other theories which offer different descriptions of gas molecule movement from the gas to the liquid phase. For these theories, the mass transfer rate equation does not differ from that of the two film model. The difference lies in the way they predict the mass transfer coefficient. It has been shown that the rate of mass transfer is dependent on a *concentration difference* multiplied by a *resistance factor*. Like most theories describing how something functions, absorption theories provide a basic understanding of the process, but due to the complexities of "real life" operations it is difficult to apply them directly. Concentrations can easily be determined from operating (c and p) and equilibrium (c_A^* and p_A^*) data of the system. Mass transfer coefficients are very difficult to determine from theory. Theoretically predicted values of the individual mass transfer coefficients (k_g and k_ℓ) based on the two film theory, do not correlate well with observed values. Overall mass transfer coefficients are more easily determined from experimental or operational data. However, the overall coefficients **apply only** when the equilibrium line is straight.

Mass transfer coefficients are often expressed by the symbols $K_{OG}a$, $k_\ell a$, etc. where "a" represents the surface area available for absorption per unit volume of the column. This allows for easy determination of the column area required to accomplish the desired separation. These mass transfer coefficients are developed from experimental data and usually reported in one of two ways: as an empirical relationship based on a function of the liquid flow or gas flow or slope of the

equilibrium line; or correlated in terms of a dimensionless number, usually either the Reynolds or Schmit Number. Figure 4-6 compares the effect on the mass transfer coefficient for SO_2 in water using two types of packing materials (Perry, 1973). Packing A is one-inch rings and packing B is three-inch spiral tiles. Similar figures are used extensively to compare different absorbers or similar absorbers with varying operating conditions.



Where:
Packing A = 1 in. rings
Packing B = 3 in. spiral tiles

Source: Perry 1973.

Figure 4-6. Comparison of overall absorption coefficient for SO_2 in water.

Although the science of absorption is considerably developed, much of the work in practical design situations is empirical in nature. The following section will apply the principles discussed, to the design of gas absorption equipment. Emphasis has been placed on presenting information that can be used to estimate absorber size and liquid flow rate.

Design Procedures

Effective gas absorption depends on intimate contact between the gas and liquid phases. This requires a high degree of turbulence between phases. However, high turbulent flow can result in extremely short contact time between the two phases, limiting absorption efficiency. The two most common absorbers are plate absorbers and packed tower absorbers. Plate absorbers are vertical towers with one or more plates mounted horizontally inside (see Figure 4-28). Gas enters the bottom of the tower and flows upward through openings in the plates. Liquid is distributed across the plates and mixing occurs as the gas bubbles up through the plates. Packed towers are either horizontal or vertical columns with liquid sprays and packing material (see Figure 4-23). Liquid introduced by sprays or weirs coats the inert

packing material that provides a large surface area for continuous gas-liquid contact. Both of these devices are used extensively to control gaseous pollutants. Absorber design calculations presented in this chapter will focus on these two devices.

Numerous procedures are used to design an absorption system. These procedures range in difficulty and cost from shortcut "rules of thumb" equations to in-depth design procedures based on pilot plant data. Procedures presented here will be based on the shortcut "rules of thumb". These procedures allow for a quick review of proposed system designs and an easy estimation of process change effects on absorber operation.

To design an absorption system, certain parameters are set by either operating conditions or regulations. The gas stream to be treated is usually the exhaust from a process in the plant. Therefore, the volume, temperature, and composition of the gas stream are given parameters. The outlet composition of the contaminant is set by the emission standard which must be met. The temperature and inlet composition of the absorbing liquid are also usually known. The main unknowns in designing the absorption system are: the *flow rate of liquid* required; the *diameter of the vessel* needed to accommodate the gas and liquid flow; and the *height of absorber* required to achieve the needed removal. Procedures for estimating these three unknowns will be discussed in the following sections.

Material Balance

In designing or reviewing the design of an absorption control system, the first task is to determine the flow rates and composition of each stream entering the system. From the law of conservation of mass, the material entering a process must either accumulate or exit. In other words, "what comes in must go out". A material balance is used to help determine flow rates and compositions of individual streams. Figure 4-7 illustrates a typical countercurrent flow absorber in which a material balance is drawn. The solute is used as the "material" in the material balance.

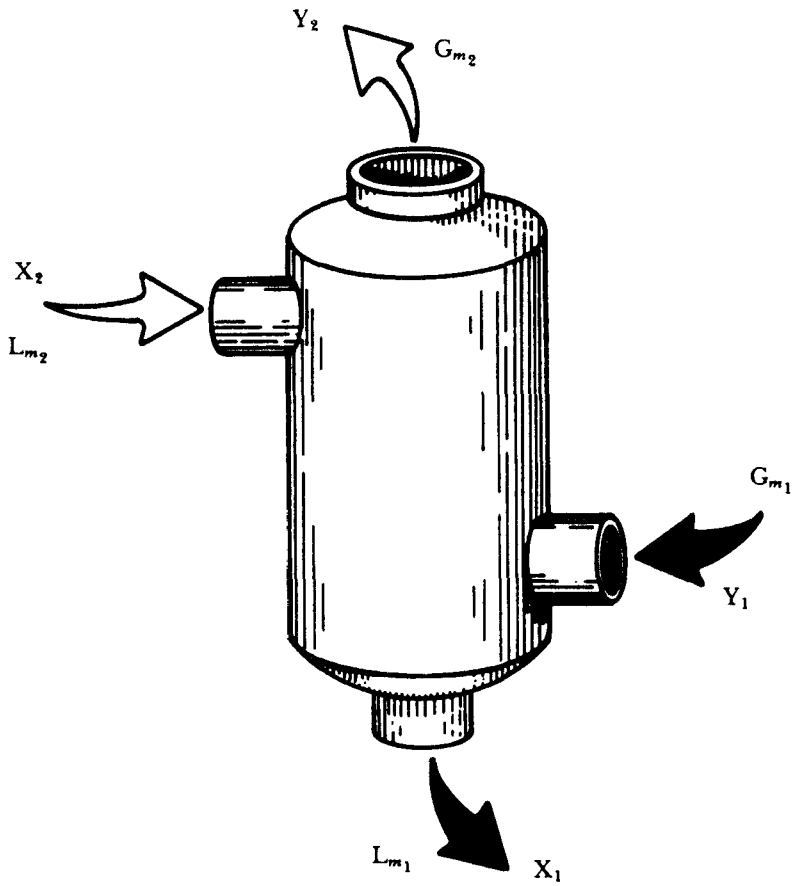


Figure 4-7. Material balance for countercurrent flow absorber.

The following procedure to set up a material balance and determine the liquid flow rate will focus on a countercurrent gas-liquid flow pattern. This is the most common flow pattern used to achieve high efficiency gas absorption. For cocurrent flow, only a slight modification (to inlet and outlet gas flows) of this procedure is required. Equations for crosscurrent flows are very complicated as they involve a gradient pattern that changes in two directions. They will not be presented here.

The terms in Figure 4-7 are defined as:

X	mole fraction of solute in pure liquid
Y	mole fraction of solute in inert gas
L_m	liquid flow rate, g mole/hr
G_m	gas flow rate, g mole/hr

Engineering design work is usually done on a solute free basis (X, Y), to make the material balance calculations easier. The solute free basis is defined as:

(Eq. 4-9) $$Y = \frac{y}{1-y}$$

(Eq. 4-10) $$X = \frac{x}{1-x}$$

In air pollution control systems the percent of pollutant transferred, y and x, is generally small. Therefore, from Equations 4-9 and 4-10 $Y \approx y$ and $X \approx x$. In this chapter, it is assumed that X and Y are always equal to x and y respectively. If y (inlet gas concentration) ever becomes larger than a few percent by volume this assumption is invalid and will cause errors in the material balance calculations.

An overall mass balance across the absorber in Figure 4-7 yields:

(Eq. 4-11)

$$\text{lb mole in} = \text{lb mole out}$$
$$G_m(\text{in}) + L_m(\text{in}) = G_m(\text{out}) + L_m(\text{out})$$

For convenience the top of the absorber is labeled as point 2 and the bottom as point 1. This changes Equation 4-11 to:

(Eq. 4-12)

$$G_{m1} + L_{m2} = G_{m2} + L_{m1}$$

In this same manner, a material balance for the contaminant to be removed is obtained:

(Eq. 4-13)

$$G_{m1}Y_1 + L_{m2}X_2 = G_{m2}Y_2 + L_{m1}X_1$$

Equation 4-13 can be simplified by assuming that as the gas and liquid streams flow through the absorber their total mass does not change appreciably (i.e. $G_{m1} = G_{m2}$ and $L_{m1} = L_{m2}$). This is justifiable for most air pollution control systems since the mass flow rate of contaminant is very small compared to the liquid and gas mass flow rates. For example, a 10,000 cfm exhaust stream containing 1000 ppm SO_2 would be only 0.1% SO_2 by volume or 1.0 cfm. If the scrubber were 100% efficient the gas mass flow rate would change from 10,000 cfm at G_{m1} to 9,999 cfm at G_{m2}. The transfer of a quantity this small is negligible in an overall material balance. Therefore, Equation 4-13 can be reduced to:

(Eq. 4-14)

$$G_m(Y_1 - Y_2) = L_m(X_2 - X_1)$$

and rearranging

(Eq. 4-15)

$$Y_1 - Y_2 = \frac{L_m}{G_m}(X_2 - X_1)$$

Equation 4-15 is the equation of a straight line. When this line is plotted on an equilibrium diagram it is referred to as an *operating line*. This line defines operating conditions within the absorber; what is going in and what is coming out. An equilibrium diagram with a typical operating line plotted on it is shown in Figure 4-8. The slope of the operating line is the liquid mass flow rate divided by the gas mass flow rate which is the *liquid-to-gas ratio* or L_m/G_m. The liquid-to-gas ratio is used extensively when describing or comparing absorption systems. Determination of the liquid-to-gas ratio is discussed in the next section.

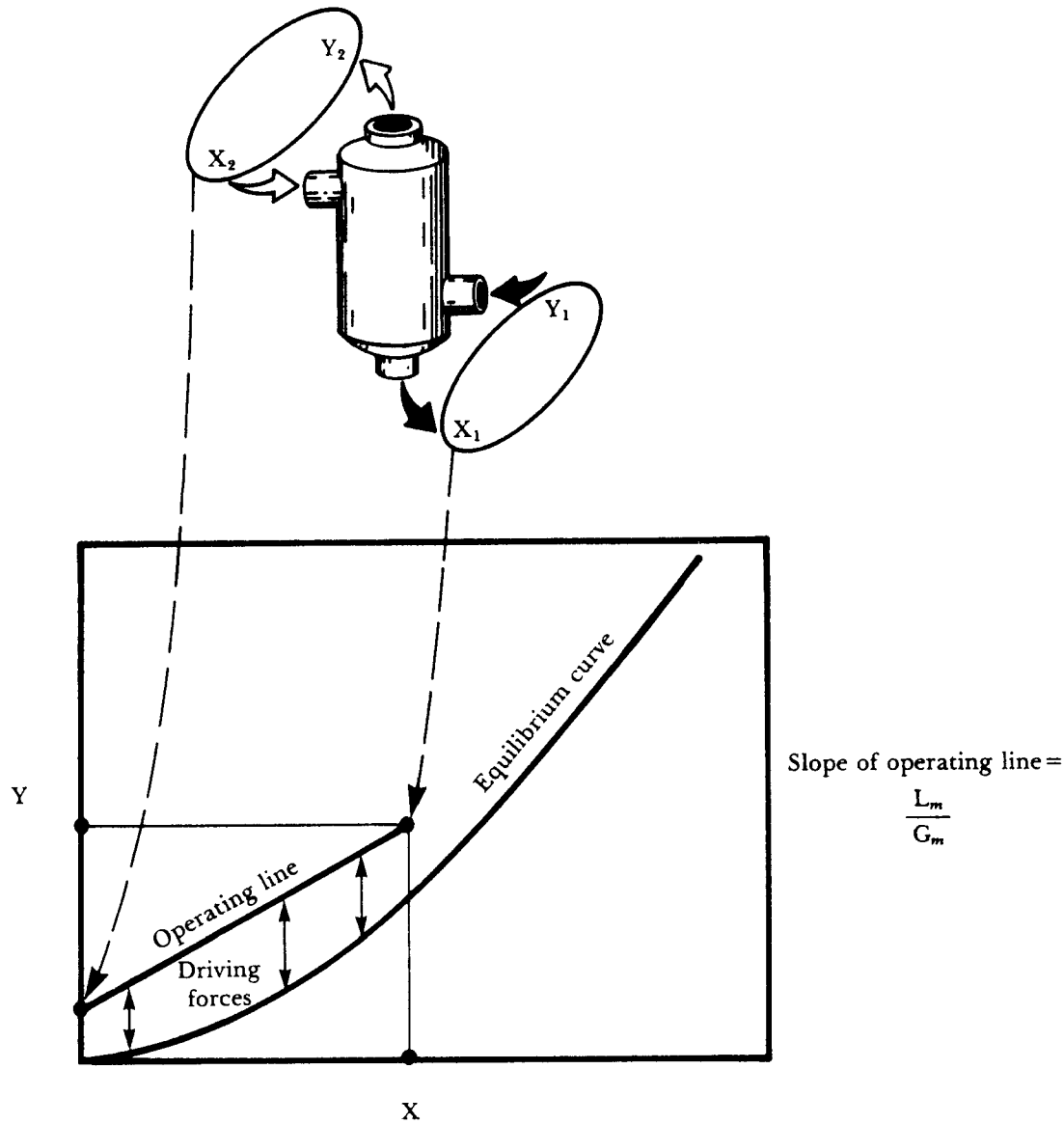


Figure 4-8. Typical operating line diagram.

Determining the Liquid Requirement

In the design of most absorption columns the quantity of gas to be treated (G_m) and the inlet solute concentration (Y_1) are set by process conditions. Minimum acceptable standards specify the outlet solute concentration (Y_2). The composition of the liquid into the absorber (X_2) is also generally known or can be assumed to be zero if there is no recycling of liquid. By plotting this data on an equilibrium diagram, the minimum amount of liquid required to achieve the required outlet concentration (Y_2) can be determined.

Figure 4-9a is a typical equilibrium diagram with operating points plotted for a countercurrent flow adsorber. At the minimum liquid rate, the inlet gas concentration of solute (Y_1) is in equilibrium with the outlet liquid concentration of solute (X_{max}). The liquid leaving the absorber is saturated with solute and can no longer dissolve any more solute unless additional liquid is added. This condition is represented by point B on the equilibrium curve.

The slope of the line drawn between point A and point B represents the operating conditions at minimum flow rate in Figure 4-9b. Note how the driving force decreases to zero at point B. The slope of line AB is (L_m/G_m)min, and may be determined graphically or from the equation for a straight line. By knowing the slope of the minimum operating line, the minimum liquid rate can easily be determined by substituting in the known gas flow rate. This procedure is illustrated in Example 4-2.

Determining the minimum liquid flow rate, (L_m/G_m)min, is important since absorber operation is usually specified as some factor of it. Generally, liquid flow rates are specified at 25 to 100% greater than the required minimum. Typical absorber operation would be at 50% greater than the minimum liquid flow rate (i.e. 1.5 times the minimum liquid-to-gas ratio). Setting the liquid rate in this way assumes that the gas flow rate set by the process does not change appreciably. Line AC in Figure 4-9c is drawn at a slope of 1.5 times the minimum L_m/G_m. Line AC is referred to as the operating line since it describes absorber operating conditions.

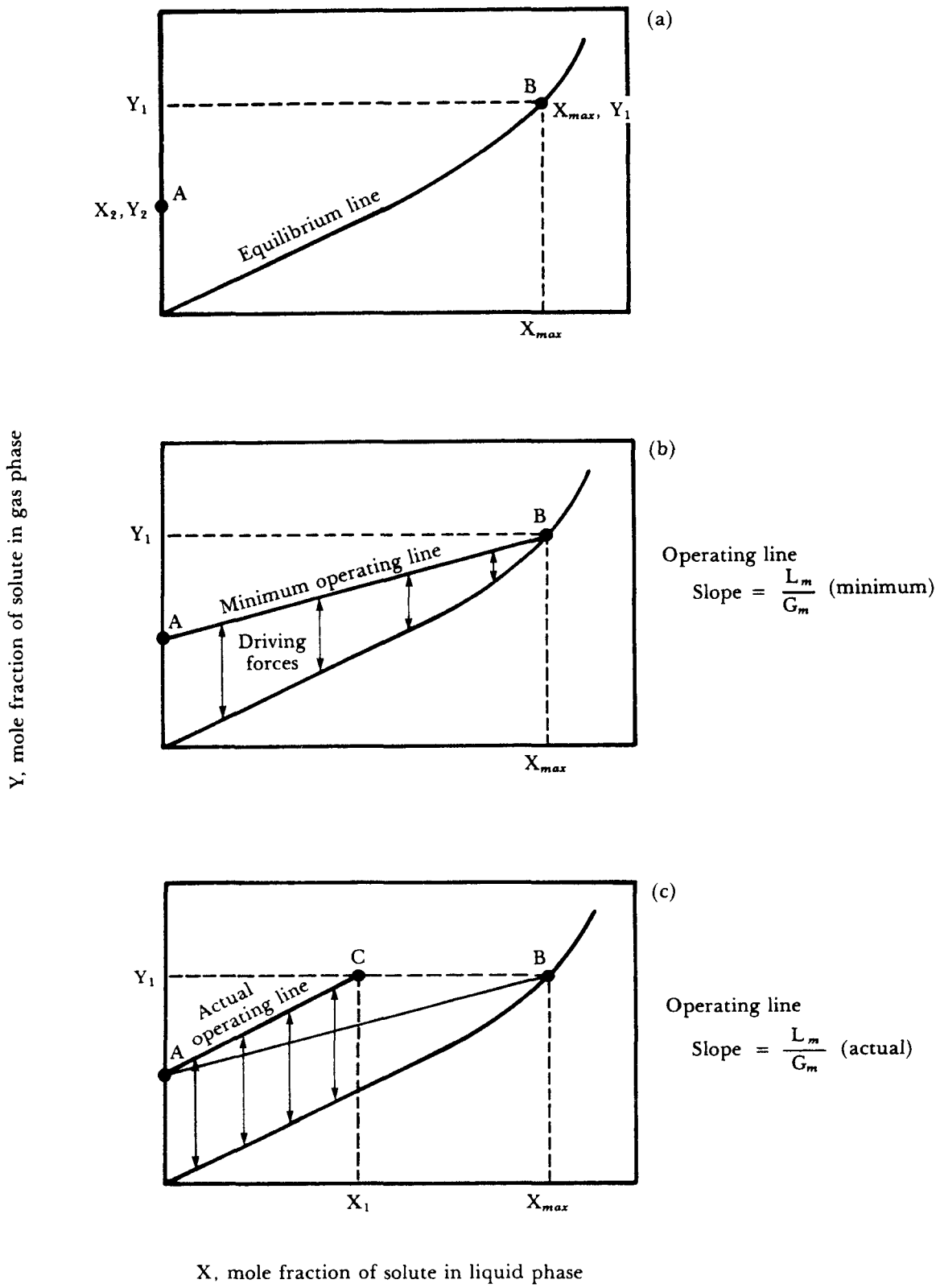


Figure 4-9. Graphic determination of liquid flow rate.

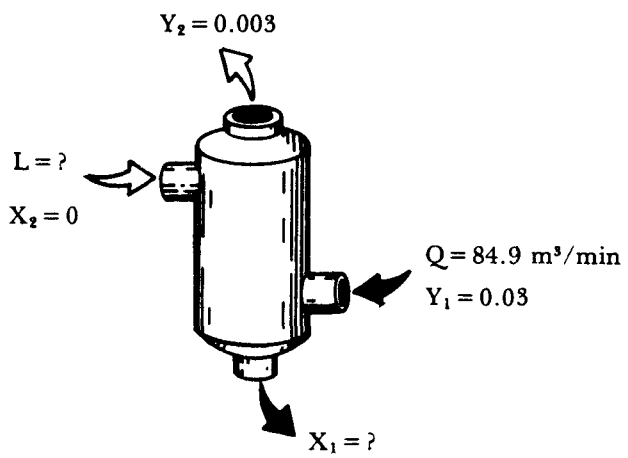
The following example problem illustrates how to compute the minimum liquid rate required to achieve a desired removal efficiency.

Example 4-2

Using the data and results from Example 4-1, compute the minimum liquid rate of pure water required to remove 90% of the SO_2 from a gas stream of 84.9 m^3/min (3000 acfm) containing 3% SO_2 by volume. The temperature is 293 K and the pressure is 101.3 kPa.

Solution:

1. First, sketch and label a drawing of the system.



$Y_1 = 3\%$ by volume $= 0.03$
$Y_2 = 90\%$ reduction from Y_1 or only 10% of Y_1; therefore
$Y_2 = (0.10)(0.03) = 0.003$

2. At the minimum liquid rate Y_1 and X_1 will be in equilibrium; the liquid will be saturated with SO_2.

   At equilibrium:

   $Y_1 = \mathscr{H}' X_1$

   $$\mathscr{H}' = 42.7 \frac{\text{mole fraction } SO_2 \text{ in air}}{\text{mole fraction } SO_2 \text{ in water}} \text{ (from Example 4-1)}$$

   $0.03 = 42.7\ X_1$
   $X_1 = 0.000703$ mole fraction

3. The minimum liquid-to-gas ratio from Equation 4-15 is:

   $$Y_1 - Y_2 = \frac{L_m}{G_m}(X_1 - X_2)$$

   $$\frac{L_m}{G_m} = \frac{Y_1 - Y_2}{X_1 - X_2}$$

$$\frac{L_m}{G_m} = \frac{0.03 - 0.003}{0.000703 - 0} = 38.4 \frac{\text{g mole water}}{\text{g mole of air}}$$

4. To compute the minimum required liquid flow rate:

   First, convert m^3 of air to g moles.

   At 0°C and 101.3 kPa there are 0.0224 m^3/g mole (359 ft^3/ mole) of an ideal gas.

   $$\text{At } 30°\text{C: } 0.0224 \frac{m^3}{\text{g mole}} \left(\frac{293 \text{ K}}{273 \text{ K}}\right) = 0.024 \text{ m}^3/\text{g mole}$$

   $$G_m = 84.9 \frac{m^3}{\text{min}} \left(\frac{\text{g mole air}}{0.024 \text{ m}^3}\right) = 3538 \frac{\text{g mole air}}{\text{min}}$$

   $$\frac{L_m}{G_m} = 38.4 \frac{\text{g mole water}}{\text{g mole air}} \text{ at minimum conditions}$$

   $L_m = (38.4)(3538) = 136.0$ kg mole water min.

   In mass units:

   $$L = \frac{136 \text{ kg mole}}{\text{min}} \left(\frac{18 \text{ kg}}{\text{kg mole}}\right) = 2448 \frac{\text{kg}}{\text{min}} \text{ (647 gal/min)}$$

5. Figure 4-10 illustrates the graphical solution to this problem.

   Multiply the slope of the minimum operating line times 1.5 to get the slope of the actual operating line (line AC).

   $38.4 \times 1.5 = 57.6$

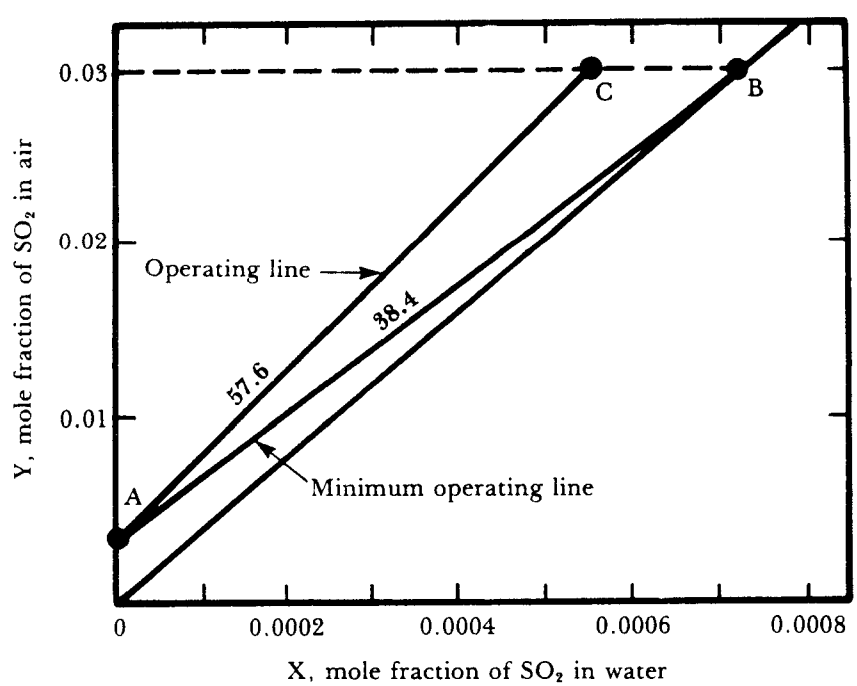


Figure 4-10. Solution to Example 4-2.

Sizing the Absorber

Packed Column Diameter

The main parameter which affects the size of a packed column is the gas velocity at which liquid droplets become entrained in the exiting gas stream. Consider a packed column operating at set gas and liquid flow rates. By decreasing the diameter of the column, the gas flow rate (m/s or ft/sec) through the column will increase. If the gas flow rate through the column is gradually increased (by using smaller and smaller diameter columns) a point will be reached where the liquid flowing down over the packing begins to be held in the void spaces between the packing. This gas to liquid flow ratio is termed the *loading* point. The pressure drop over the column begins to increase and the degree of mixing between the phases decreases. A further increase in gas velocity will cause the liquid to completely fill the void spaces in the packing. The liquid forms a layer over the top of the packing and no more liquid can flow down through the tower. The pressure drop increases substantially and mixing between the phases is minimal. This condition is referred to as *flooding* and the gas velocity at which it occurs is the *flooding velocity*. Using an extremely large diameter tower would eliminate this problem. However, as the diameter increases the cost of the tower increases.

Normal practice is to size a packed column diameter to operate at a certain percent of the flooding velocity. A typical operating range for the gas velocity through the columns is 50 to 75% of the flooding velocity. It is assumed that by operating in this range the gas velocity will also be below the loading point.

A common and relatively simple procedure to estimate the flooding velocity (thus setting a minimum column diameter) is to use a generalized flooding and pressure drop correlation. One version of the flooding and pressure drop relationship in a packed tower is the Sherwood Correlation shown in Figure 4-11 (EPA, 1972). This correlation is based on the physical properties of the gas and liquid streams and tower packing characteristics. The procedure to determine the tower diameter is:

1. Calculate the value of the abscissa.

(Eq. 4-16)

$$\frac{L}{G}\left(\frac{\varrho_g}{\varrho_\ell}\right)^{0.5}$$

Where: L and G = mass flow rates: any consistant set of units may be used as long as the abscissa is dimensionless
ϱ_g = density of the gas stream
ϱ_ℓ = density of the absorbing liquid

2. From the point calculated in Equation 4-16 proceed up the graph to the flooding line and read the ordinate ϵ.

3. Rearrange the equation of the ordinate and solve for G′.

(Eq. 4-17)

$$G' = \left[\frac{(\epsilon)(\varrho_g)(\varrho_\ell)(g_c)}{F\phi\mu_\ell^{0.2}}\right]^{0.5}$$

Where: F = packing factor given in Table 4-3 for different types of packing (Bhatia, 1977)

ϕ = ratio of specific gravity of the scrubbing liquid to that of water

μ_ℓ = viscosity of liquid (for water = 0.8 centipoise)

G′ = mass flow rate of gas per unit cross sectional area of column at flooding, g/s•m² (lb/s•ft²)

ϱ_ℓ = density of the absorbing liquid, kg/m³ (lb/ft³)

ϱ_g = density of the gas stream, kg/m³ (lb/ft³)

g_c = gravitational constant = 9.82 m/s² (32.2 ft/s²)

4. G′ at operating is a fraction of G′ at flooding.

(Eq. 4-18)
$$G'_{operating} = (f)(G'_{flooding})$$

5. The cross sectional area of column A is calculated from:

(Eq. 4-19)
$$A = \frac{G}{G'_{operating}}$$

6. The diameter of the column is obtained from:

(Eq. 4-20)
$$d_t = \left(\frac{4A}{\pi}\right)^{0.5} = 1.13\ A^{0.5}$$

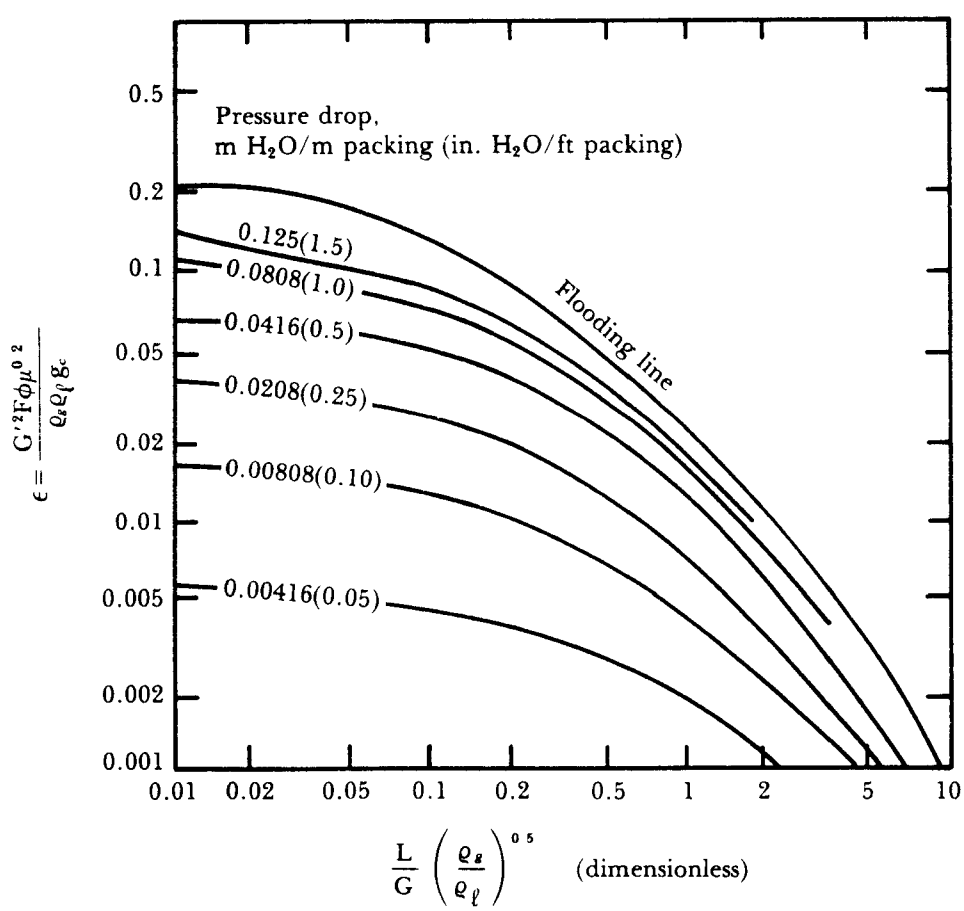


Figure 4-11. Generalized flooding and pressure drop correlation.

Table 4-3. Packing data.*

Packing	Size (in.)	Weight (lb/ft³)	Surface area, a (ft²/ft³ packing volume)	Void fraction (%)	Packing factor, F (ft²/ft³)	Price ($/ft³)
Raschig rings (ceramic and porcelain)	½	52	114	65	580	14.00[a], 19.00[b]
	1	44	58	70	155	7.00[a], 9.30[b]
	1½	42	36	72	95	6.30[a], 8.40[b]
	2	38	28	75	65	6.20[a], 8.30[b]
	3	34	19	77	37	6.50[a], 9.50[b]
Raschig rings (steel)	½ × 1/32	77	128	84	300	105.50
	1 × 1/32	40	63	92	115	75.10
	2 × 1/16	38	31	92	57	39.10
Berl saddles (ceramic and porcelain)	¼	55	274	63	900	—
	½	54	155	64	240	24.00[a], 32.00[b]
	1	48	79	68	110	9.50[a], 12.00[b]
	2	38	32	75	45	7.00[a], 9.30[b]
Intalox saddles (ceramic)	¼	54	300	75	725	—
	½	45	190	78	200	28.35
	1	44	78	77	98	10.85
	2	42	36	79	40	—
Intalox saddles (plastic)	1	6.00	63	91	30	—
	2	3.75	33	93	20	7.65
	3	3.25	27	94	15	—
Pall rings (plastic)	5/8	7.0	104	87	97	—
	1	5.5	63	90	52	10.75
	2	4.5	31	92	25	—
Pall rings (metal)	5/8 × 0.018 thick	38	104	93	73	81.25[c]
	1½ × 0.03 thick	24	39	95	28	33.23[c]
Tellerettes	1	7.5	55	87	40	12.50
	2	3.9	38	93	20	7.30
	3	5.0	30	92	15	5.25

[a]Prices for ceramic packing
[b]Prices for porcelain packing
[c]Prices for 304 S.S. packing

*Note: data for guide purposes only.

Source: Bhatia, 1977.

The following example illustrates the use of Figure 4-11 for computing the minimum allowable diameter for a packed tower.

Example 4-3

For the scrubber in Example 4-2, determine the column diameter if the operating liquid rate is 1.5 times the minimum. The gas velocity should be no greater than 75% of the flooding velocity and the packing material is two-inch ceramic Intalox saddles.

Solution:

From Example 4-2:

$G_m = 3538$ g mole/min

$L_{(minimum)} = 2448$ kg/min

Convert gas molar flow to a mass flow, assuming molecular weight of the gas to be 29 kg/mole.

$$G = 3.538 \frac{\text{kg mole}}{\text{min}} \times \frac{29 \text{ kg}}{\text{mole}} = 102.6 \text{ kg/min}$$

Adjusting the liquid flow to 1.5 times the minimum:

$L = 1.5\ (2448) = 3672$ kg/min

The densities of air and water at 30°C are:

$\varrho_\ell = 1000$ kg/m³

$\varrho_g = 1.17$ kg/m³

1. Now using the relationship in Equation 4-16, compute the abscissa.

$$\left(\frac{L}{G}\right)\left(\frac{\varrho_g}{\varrho_\ell}\right)^{0.5}$$

$$\left(\frac{3672}{102.6}\right)\left(\frac{1.17}{1000}\right)^{0.5} = 1.22$$

2. From Figure 4-11, proceed up to the flooding line from 1.22. The ordinate ϵ is 0.019.

3. Use Equation 4-17 to solve for the superficial flooding velocity, G′.
   A superficial velocity is a flow rate per unit cross sectional area.

$$G' = \left[\frac{(\epsilon)(\varrho_g)(\varrho_\ell)(g_c)}{F\phi\mu_\ell^{0.2}}\right]^{0.5}$$

For water, $\phi = 1.0$ and $\mu_\ell = 0.0008$ Pa•s

From Table 4-3 for two-inch Intalox saddles, $F = 40$ ft²/ft³ or 131 m²/m³.

$g_c = 9.82$ m/s²

$$G' = \left[\frac{(0.019)(1.17)(1000)(9.82)}{(1)(131)(0.0008)^{0.2}}\right]^{0.5}$$

$= 2.63$ kg/m²•s at flooding

4. The superficial gas velocity at operating is obtained from Equation 4-18.

$$G'_{operating} = fG'_{flooding} = (0.75)(2.63) = 1.97 \text{ kg/m}^2\bullet\text{s}$$

5. From Equation 4-19, the cross sectional area of the tower is:

$$A = \frac{G}{G'_{operating}}$$

$$= \frac{(102.6 \text{ kg/min})(\text{min}/60\text{s})}{1.97 \text{ kg/m}^2\bullet\text{s}} = 0.87 \text{ m}^2$$

6. The diameter of the tower, from Equation 4-20 is:

$$d_t = \left(\frac{4A}{\pi}\right)^{0.5}$$

$$= \left[\frac{(4)(0.87 \text{ m}^2)}{\pi}\right]^{0.5}$$

$$= 1.05 \text{ m or at least } 1.1 \text{ m } (3.5 \text{ ft})$$

Figure 4-11 may also be used to estimate the pressure drop of the tower once $G'_{operating}$ is set.

This is done by plugging G' back into the equation to compute ϵ.

$$\epsilon = \frac{G'^2 \phi F \mu_\ell^{0.2}}{\varrho_g \varrho_\ell g_c}$$

$$= \frac{(1.97)^2(1)(131)(0.0008)^{0.2}}{(1.17)(1000)(9.82)}$$

$$= 0.0106$$

The abscissa remains unchanged and equals 1.22.

The pressure drop through the column is the point at which these two lines cross. From Figure 4-11:

$$\Delta p \approx \frac{0.0416 \text{ meter of } H_2O}{\text{meter of packing}} \left(\frac{0.5 \text{ inches of } H_2O}{\text{feet of packing}}\right)$$

Packed Column Height

The height of a packed column refers to the depth of packing material needed to accomplish the required removal efficiency. The more difficult the separation, the larger the packing height required. For example, a much larger packing height would be required to remove SO_2 than to remove Cl from an exhaust stream using water as the absorbent. This is because Cl is more soluble in water than SO_2. Determining the proper height of packing is important since it affects both the rate and efficiency of absorption.

A number of theoretical equations are used to predict the required packing height. These equations are based on diffusion principles. Depending on which phase is controlling the absorption process, either Equation 4-5 or 4-6 is used as the starting point to derive an equation to predict column height. A material balance is then set up over a small differential section of the column. Derivations of the equations used can be found in a number of chemical engineering texts or other books dealing with mass transfer, and will not be covered here.

The general form of the design equation for a *gas phase controlled resistance* is given by:

(Eq. 4-21)

$$Z = \frac{G'}{K_g aP} \int_{Y_2}^{Y_1} \frac{dY}{(1-Y)(Y-Y^*)}$$

Where: Z = height of packing, m
a = interfacial contact area, m^2
P = pressure of the system, kPa

In analyzing Equation 4-21, the term $G'/K_g aP$ has the dimension of meters and is defined as the *height of a transfer unit*. The term inside the integral is dimensionless and represents the *number of transfer units* needed to make up the total packing height. Using the concept of transfer units, Equation 4-21 can be simplified to:

(Eq. 4-22)

$$Z = HTU \times NTU$$

Where: HTU = height of a transfer unit, m
NTU = number of transfer units

The concept of a transfer unit comes from the operation of plate columns. Discrete stages (plates) of separation occur in plate columns. These stages can be visualized as a transfer unit with the number and height of each giving the total tower height. Although packed columns operate as one continuous separation process, in design terminology it is treated as if it were broken into discrete sections (height of a transfer unit). The number and the height of a transfer unit are based on either the gas or liquid phase. Equation 4-22 now becomes:

(Eq. 4-23)

$$Z = N_{OG}H_{OG} = N_{OL}H_{OL}$$

Where: N_{OG} = number of transfer units based on overall gas film coefficient
N_{OL} = number of transfer units based on overall liquid film coefficient
H_{OG} = height of a transfer unit based on overall gas film coefficient, m
H_{OL} = height of a transfer unit based on an overall liquid film coefficient, m

Values for the height of a transfer unit used in designing absorption systems are usually obtained from experimental data. To ensure greatest accuracy, vendors of absorption equipment normally perform pilot plant studies to determine the height of a transfer unit. For common absorption systems, such as NH_3-water, manufacturers have developed graphs which can be used to estimate the height of a transfer unit. These graphs do not provide the accuracy of pilot plant data, but they are less expensive and easier to use. Figure 4-12 gives a typical example of these graphs for an ammonia-water system. In this figure, the superficial gas flow rate is plotted versus the H_{OG} with the superficial liquid rate as a parameter. It is also common to plot liquid rate versus the H_{OG} and have the gas rate as a parameter. Additional information on other gas-liquid systems can be found in *Chemical Engineers' Handbook* (Perry, 1973). In applying these data it is important that the process conditions be similar to the conditions at which the HTU was measured.

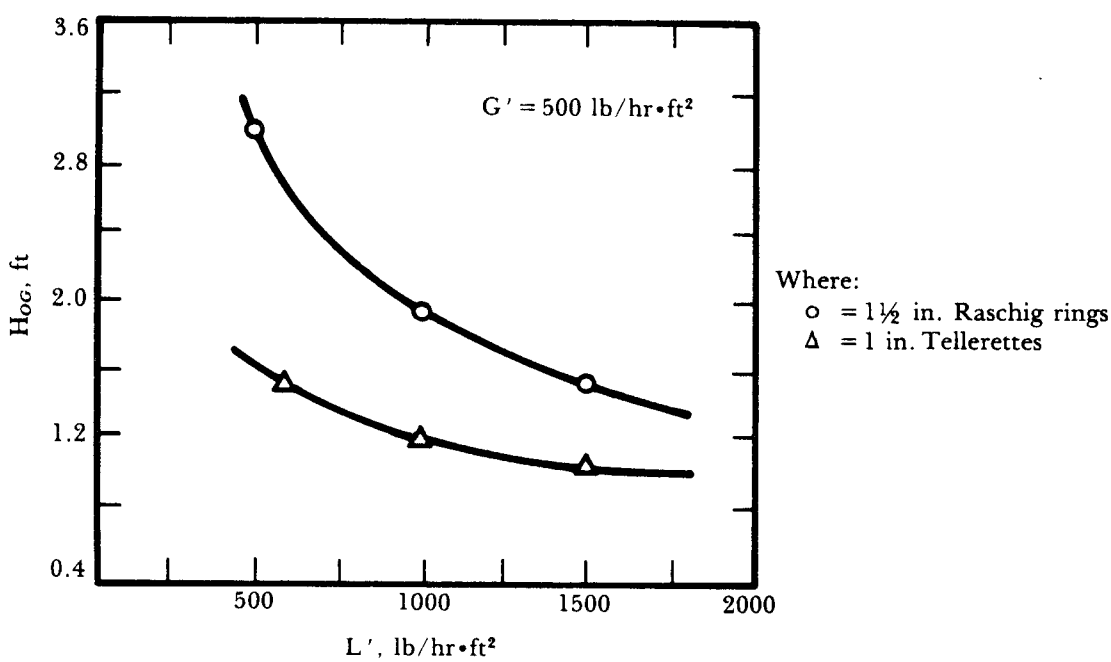


Figure 4-12. Column packing comparison for ammonia and water system.

When no experimental data are available or if only a preliminary estimate of absorber efficiency is needed, there are generalized correlations available to predict the height of a transfer unit. The correlations for predicting the H_{OG} or the H_{OL} are empirical in nature and are a function of: *type of packing; liquid* and *gas flow rates; concentration* and *solubility* of the contaminant; *liquid properties;* and *system temperature*. These correlations can be found in engineering texts such as *Chemical Engineers' Handbook* (Perry, 1973), EPA (1972), or Treybal (1968). For most applications, the height of a transfer unit ranges between 0.3 to 1.2 m (1 to 4 ft) (Calvert, 1977). As a rough estimate, 0.6 m (2.0 ft) can be used.

The number of transfer units, NTU, can be obtained experimentally or calculated from a variety of methods. For the case where the solute concentration is very low and the equilibrium line is straight, Equation 4-24 can be used to determine the number of transfer units (N_{OG}) based on the gas phase resistance. Equation 4-24 can be derived from the integral portion of Equation 4-21.

(Eq. 4-24)

$$N_{OG} = \frac{\ln\left[\frac{Y_1 - mX_2}{Y_2 - mX_2}\left(1 + \frac{mG_m}{L_m}\right) + \frac{mG_m}{L_m}\right]}{1 - \frac{mG_m}{L_m}}$$

Where: m = slope of equilibrium line
G_m = molar flow rate of gas, kg mole/hr
L_m = molar flow rate of liquid, kg mole/hr
X_2 = mole fraction of solute entering the column
Y_1 = mole fraction of solute in entering gas
Y_2 = mole fraction of solute in exiting gas

Equation 4-24 may be solved directly or graphically by using the Colburn Diagram which is presented in Figure 4-13. The Colburn Diagram is a plot of the N_{OG} versus $\ln[Y_1 - mX_2/Y_2 - mX_2]$ at various values of (mG_m/L_m). The term (mG_m/L_m) is referred to as the *absorption factor*. Figure 4-13 is used by first computing the value of $\ln[Y_1 - mX_2/Y_2 - mX_2]$, reading up the graph to the line corresponding to (mG_m/L_m), and then reading across to obtain the N_{OG}.

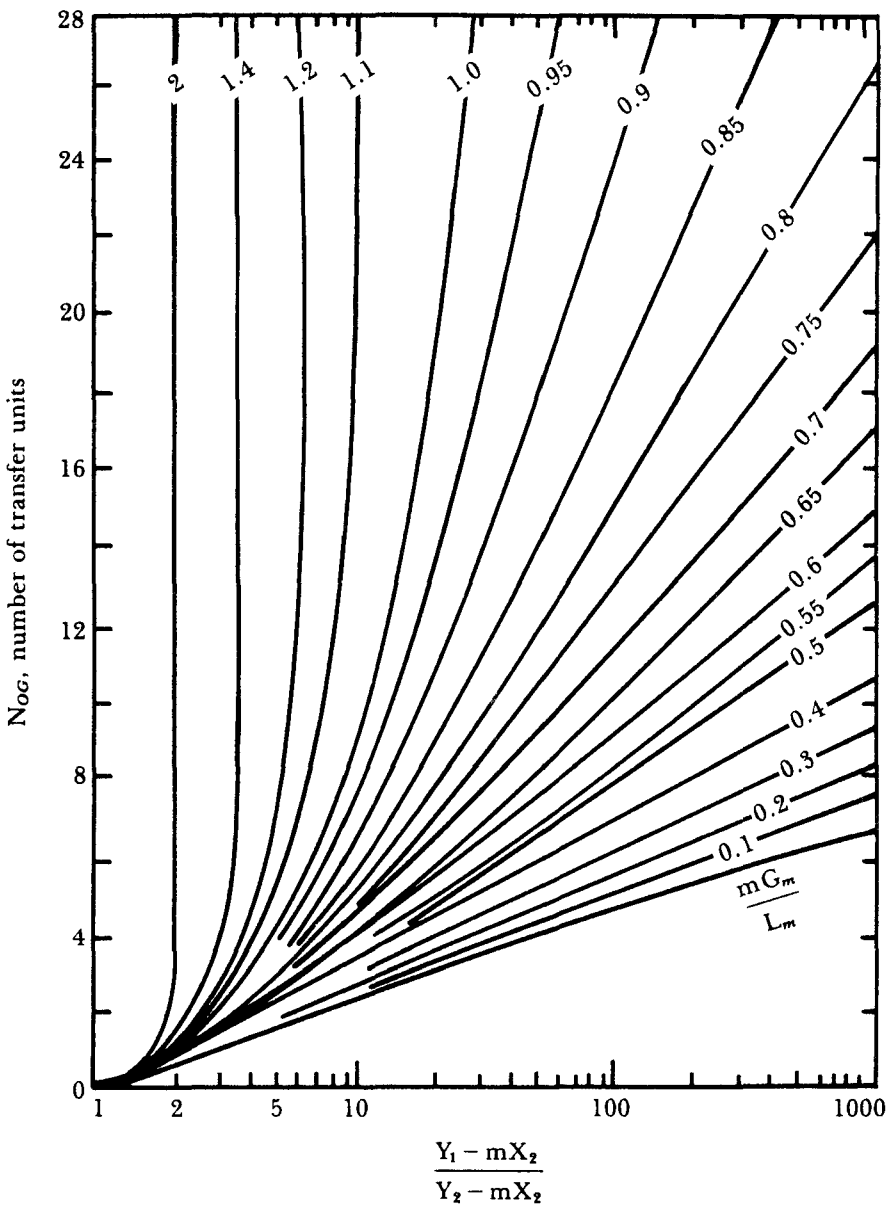


Figure 4-13. Colburn Diagram.

Equation 4-24 can be further simplified for situations where a **chemical reaction** occurs or if the solute is **extremely** soluble. In these cases the solute exhibits almost no partial pressure and therefore the slope of the equilibrium line approaches zero ($m \rightarrow 0$). For either of these cases Equation 4-24 reduces to:

(Eq. 4-25)

$$N_{OG} = \ln \frac{Y_1}{Y_2}$$

The number of transfer units only depends on the inlet and outlet concentration of the solute (contaminant). For example, if the conditions of Equation 4-25 are met, to achieve 90% removal of any pollutant requires 2.3 transfer units. Equation 4-25 applies **only** when the **equilibrium line is straight** and approaches zero (for very soluble or reactive gases). The following example illustrates the use of Equation 4-24.

Example 4-4

From pilot plant studies of the absorption system in Example 4-2, it was determined that the H_{OG} for the SO_2-water system is 0.829 m (2.72 ft). Calculate the total height of packing required to achieve the 90% removal. The following data were taken from the previous examples:

$$m = 42.7 \frac{\text{kg mole } H_2O}{\text{kg mole air}}$$

$$G_m = 3.5 \frac{\text{kg mole}}{\text{min}}$$

$$L_m = 3672 \frac{\text{kg}}{\text{min}} \times \frac{\text{kg mole}}{18 \text{ kg}} = 204 \frac{\text{kg mole}}{\text{min}}$$

$X_2 = 0$ (no recycle liquid)

$Y_1 = 0.03$

$Y_2 = 0.003$

Solution:

Compute the N_{OG} from Equation 4-24:

$$N_{OG} = \frac{\ln\left[\frac{Y_1 - mX_2}{Y_2 - mX_2}\left(1 - \frac{mG_m}{L_m}\right) + \frac{mG_m}{L_m}\right]}{1 - \frac{mG_m}{L_m}}$$

$$= \frac{\ln\left[\left(\frac{0.03}{0.003}\right)\left(1 - \frac{(42.7)(3.5)}{204}\right) + \frac{(42.7)(3.5)}{204}\right]}{1 - \frac{(42.7)(3.5)}{204}}$$

$$= 4.58$$

The total packing height is thus:

$$Z = H_{OG} \times N_{OG}$$

$$Z = (0.829)(4.58) = 3.79 \text{ m of packing height}$$

Plate Tower

In a plate tower the scrubbing liquid enters at the top of the tower, passes over the top plate, and then down over each lower plate until the liquid reaches the bottom. Absorption occurs as the gas, which enters at the bottom, passes up through the plate and contacts the liquid. In a plate tower, absorption occurs in a stepwise or stage process. The operation of plate towers is discussed in greater detail in the equipment section of this manual.

There are various accepted procedures available to size a plate tower. Detailed summaries can be found in *Chemical Engineers' Handbook* (Perry, 1973), McCabe and Smith (1956), and Theodore and Buonicore (1975). The following discussion presents a simplified method for sizing or reviewing the design of a plate tower.

Plate Tower Diameter

The minimum diameter of a single-pass plate tower is determined by using the gas velocity through the tower. If the gas velocity is too fast, liquid droplets are entrained, causing a condition known as *priming*. Priming occurs when the gas velocity through the tower is so fast that it causes liquid on one tray to foam and then rise to the tray above. Priming reduces absorber efficiency by inhibiting gas and liquid contact. For the purpose of determining tower diameter, priming in a plate tower is analogous to the flooding point in a packed tower. It determines the **minimum acceptable** diameter. The **actual** diameter should be larger.

The smallest allowable diameter for a plate tower is expressed by:

(Eq. 4-26)

$$d_t = \psi (Q \sqrt{\varrho_g})^{0.5}$$

Where: Q = volumetric gas flow, m^3/hr
ψ = empirical correlation, $m^{0.25}\ hr^{0.25}/kg^{0.25}$
ϱ_g = gas density, kg/m^3

The term ψ is an empirical correlation and is a function of both the tray spacing and the densities of the gas and liquid streams. Values for ψ in Table 4-4 are for a tray spacing of 61 cm (24 in.) and a liquid specific gravity of 1.05 (EPA, 1972). If the specific gravity of a liquid varies significantly from 1.05, the values for ψ in Table 4-4 **cannot** be used.

Table 4-4. Empirical constants for Equation 4-26.

Tray	Metric ψ^a	Engineering ψ^b
Bubble cap	0.0162	0.1386
Sieve	0.0140	0.1198
Valve	0.0125	0.1069

[a] Metric ψ is expressed in $m^{0.25}hr^{0.5}/kg^{0.25}$, for use with Q expressed in m^3/hr, and ϱ_g expressed in kg/m^3.
[b] English ψ is expressed in $ft^{0.25}min^{0.5}/lb^{0.25}$, for use with Q in cfm, and ϱ_g expressed in lb/ft^3.

Source: EPA, 1972.

Depending on operating conditions, trays are spaced at a minimum distance between plates to allow the gas and liquid phases to separate before reaching the plate above. Trays should be spaced to allow for easy maintenance and cleaning. Trays are normally spaced 45 to 70 cm (18 to 28 inches) apart. In using Table 4-4 for a tray spacing different than 61 cm, a correction factor must be used. Figure 4-14 is used to determine the correction factor which is multiplied times the estimated diameter. Example 4-5 illustrates how to estimate the minimum diameter of a plate tower.

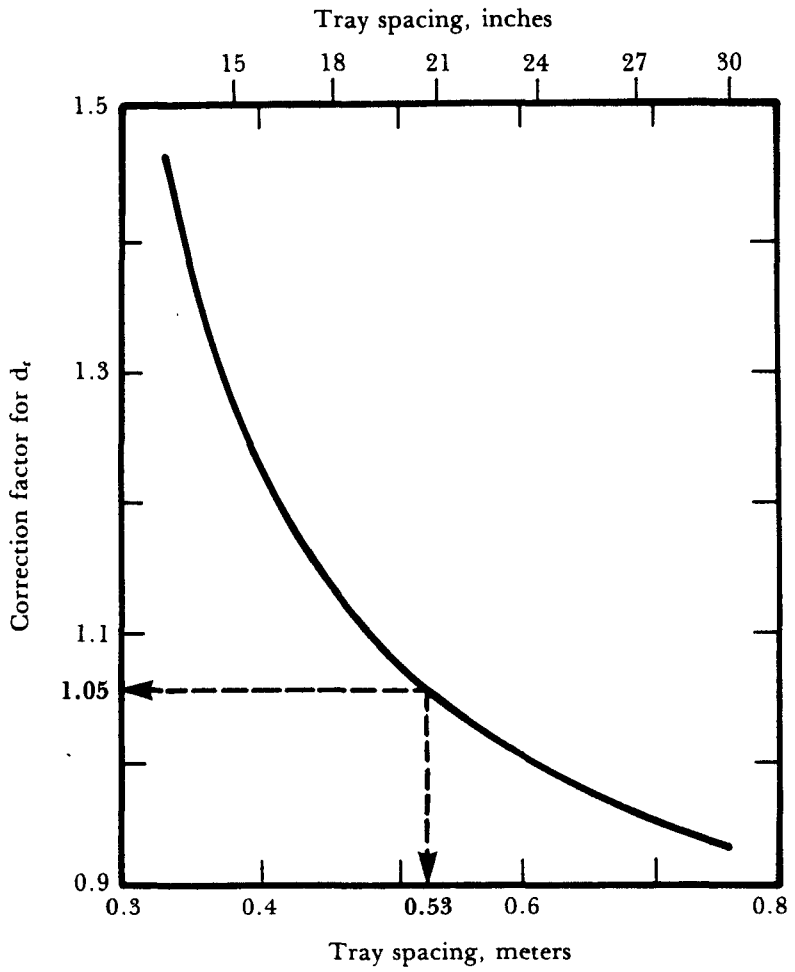


Figure 4-14. Tray spacing correction factor.

Example 4-5

For the condition described in Example 4-2, determine the minimum acceptable diameter if the scrubber is a bubble cap tray tower. The trays are spaced 0.53 m (21 inches) apart.

Solution:

From Example 4-2, the following information is obtained:

Gas flow rate = Q = 84.9 m^3/min

Density = ϱ_g = 1.17 kg/m^3

From Table 4-4 for a bubble cap tray:

$\psi = 0.0162\ m^{0.25}\ hr^{0.25}/kg^{0.25}$

Before Equation 4-26 can be used, Q must be converted to m^3/hr.

$$Q = 84.9\frac{m^3}{min} \times \frac{60\ min}{hr} = 5094\ m^3/hr$$

Substituting these values into Equation 4-26 for a minimum d_t:

$$d_t = \psi(Q\sqrt{\varrho_g})^{0.5}$$

$$= (0.0162)[5094(\sqrt{1.17})]^{0.5} = 1.2\ m$$

Correct this diameter for a tray spacing of 0.53 m.

From Figure 4-14 read the correction factor as 1.05.

Therefore, the minimum diameter is:

d_t = 1.2 (1.05) = 1.26 m (4.13 ft)

Note: that this estimated diameter is a **minimum** acceptable diameter based on primary conditions. In practice, a larger diameter based on economic conditions is usually chosen.

Number of Theoretical Plates or Trays

There are several methods used to determine the number of ideal plates or trays which are required for a given removal efficiency. These methods, however, can become quite complicated. One method used is a graphical technique.

The number of ideal plates is obtained by drawing "steps" on an operating diagram. This procedure is illustrated in Figure 4-15. This method can be rather time consuming and inaccuracies can result at both ends of the graph.

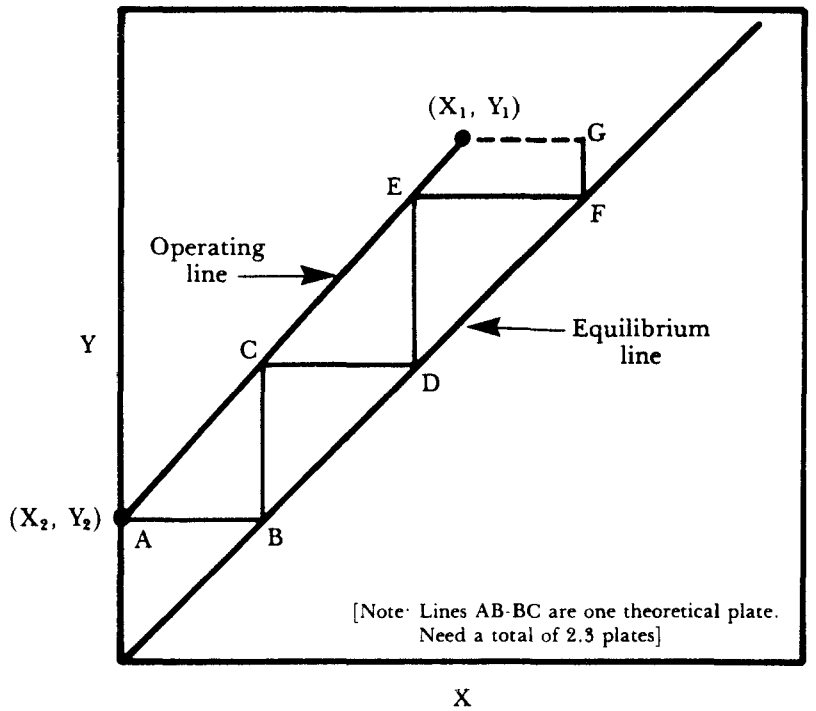


Figure 4-15. Graphic determination of the number of theoretical plates.

Equation 4-27 is a simplified method to estimate the number of plates. Equation 4-27 can only be used if both the equilibrium and operating lines for the system are straight. This is a valid assumption for most air pollution control systems. Equation 4-27, taken from Sherwood and Pigford (1952), is very similar to Equation 4-24 for computing the N_{OG} of a packed tower.

(Eq. 4-27)

$$N_p = \frac{\ln\left[\left(\frac{Y_1 - mX_2}{Y_2 - mX_2}\right)\left(1 - \frac{mG_m}{L_m}\right) + \frac{mG_m}{L_m}\right]}{\ln\left(\frac{L_m}{mG_m}\right)}$$

Equation 4-27 is used to predict the *number of theoretical plates* required to achieve a given removal efficiency. The operating conditions for a theoretical plate assume that the gas and liquid stream leaving the plate are in equilibrium with each other. This **ideal** condition is never achieved in practice. A larger number of **actual trays** are required to compensate for this decreased tray efficiency.

Three types of efficiencies are used to describe absorption efficiency for a plate tower: an *overall efficiency*, which is concerned with the entire column; *Murphree efficiency*, which is applicable with a single plate; and *local efficiency*, which pertains to a specific location on a plate. A number of methods are available to predict these plate efficiencies. These methods are complex and values predicted by two different methods for a given system can vary by as much as 80% (Zenz, 1972).

The simplest of the tray efficiency concepts, the overall efficiency, is the ratio of the number of theoretical plates to the number of actual plates. Since overall tray efficiency is an oversimplification of the process, reliable values are difficult to obtain. For a rough estimate, overall tray efficiencies for absorbers operating with low viscosity liquid normally fall in a 65 to 80% range (Zenz, 1972).

Example 4-6

Calculate the number of theoretical plates required for the scrubber in Example 4-5, using the same conditions as in Example 4-4. Estimate the total height of the column if the trays are spaced at 0.53 m intervals and assume an overall tray efficiency of 70%.

Solution:

From Example 4-5 and the previous examples the following data is obtained:

Slope of equilibrium line, $m = 42.7$

$Y_1 = 0.03$

$Y_2 = 0.003$

$X_2 = 0.0$

$L_m = 204$ kg mole/min

$G_m = 3.5$ kg mole/min

The number of theoretical plates from Equation 4-26 is:

$$N_p = \frac{\ln\left[\frac{Y_1 - mX_2}{Y_2 - mX_2}\left(1 - \frac{mG_m}{L_m}\right) + \frac{mG_m}{L_m}\right]}{\ln\left(\frac{L_m}{mG_m}\right)}$$

$$= \frac{\ln\left[\frac{0.03 - 0}{0.003 - 0}\left(1 - \frac{42.7(3.5)}{204}\right) + \frac{42.7(3.5)}{204}\right]}{\ln\left[\frac{204}{42.7(3.5)}\right]}$$

$$= 3.94 \text{ theoretical plates}$$

Assuming that the overall plate efficiency is 70%, the actual number of plates is:

$$\text{Actual plates} = \frac{3.94}{0.70} = 5.6 \text{ or } 6 \text{ plates}$$

The height of the tower is given by:

$Z = N_p \times$ tray spacing + top height

The top height is the distance over the top plate which allows the gas-vapor mixture to separate. This distance is usually the same as the tray spacing.

$$Z = 6 \text{ plates } (0.53 \text{ m}) + 0.53 \text{ m}$$
$$= 3.71 \text{ m}$$

Note: that this height is approximately the same as that predicted for the packed tower in Example 4-4. This seems logical since both the packed and plate towers are efficient gas absorption devices. However, due to the many assumptions, no concrete generalization can be made.

Absorption Equipment

Introduction

The primary function of an absorber is to remove gaseous contaminants from an exhaust air stream. To accomplish this, absorption equipment is designed to maximize the mass transfer rate. In absorption, the rate of mass transferred depends largely on the surface area of the air and liquid stream exposed to each other. Absorption proceeds at a finite rate. Increasing the time the two streams are in contact will increase the potential for absorption to occur. Absorbers are designed to provide the necessary surface area and sufficient contact time between the gas and liquid streams.

Although the primary objective of an absorber is to remove gaseous contaminants, it can also perform other functions. Removal of particulate matter is a prime example. When used for gaseous removal the control device is referred to as an "absorber". When used for particulate matter removal, the device is termed a "wet scrubber". These terms are used interchangeably as they describe the same piece of equipment. The difference lies in the manner in which the equipment is operated. Absorption is enhanced by slowing the relative velocity between the gas and liquid streams (increase contact time). Whereas for particulate matter removal, the exact opposite is true. Collection efficiency increases by increasing the relative velocity of the two streams (decrease contact time). The same piece of equipment can be used for either gas or particulate matter removal. But it is extremely difficult to achieve a high removal efficiency for both pollutants unless the gaseous contaminant is extremely soluble.

The physical and chemical characteristics of the exhaust gas stream play an important role in both the selection and proper operation of an absorption system. The solubility of the gaseous contaminant is the first characteristic to evaluate. If the gaseous contaminant is very soluble, then high removal efficiencies can be achieved by almost any absorption device. For a relatively insoluble contaminant only certain systems may be able to achieve the required removal efficiency. In some cases a chemical reagent may have to be added to the absorbing liquid to increase the solubility of the contaminant. These reagents may increase the physical solubility of the contaminant (i.e. sodium citrate added to absorb SO_2) or can

chemically react with the contaminant (i.e. lime scrubbing of SO_2). If a precipitate is formed by a chemical reaction when a reagent is used, plugging or corrosion problems may arise.

The temperature of the exhaust stream is another important characteristic which affects absorption. The solubility of a gas decreases with an increase in operating temperature. As the temperature increases, so does the kinetic energy of the gas molecules in solution. At these higher states of energy the gas molecules will come out of solution. This loss of solubility at higher temperatures necessitates that some gas streams be cooled before effective absorption occurs. For example, in certain FGD systems the exhaust gases from the boiler must be cooled from 150°C (300°F) to approximately 50°C (125°F) to achieve the desired SO_2 removal efficiency. This is accomplished by adding inlet sprays or adjusting the liquid flow rate to the absorber. The temperature of the exhaust stream also affects the size of the absorption system. Decreasing the temperature decreases the volume of gases which must be handled. This decreases the size of the absorption system.

Selection of the proper absorbing liquid is based on the efficiency required and the liquid cost. Water is the usual choice because many gaseous contaminants are soluble in it, it is readily available, and relatively inexpensive. The following properties must also be kept in mind when selecting a liquid:

Gas solubility: high solubility increases the absorption rate and minimizes the quantity of liquid needed.

Volatility: low volatility of the liquid will reduce the amount of vapor that is lost in the existing gas stream.

Viscosity: low viscosity promotes rapid absorption rates, improves flooding characteristics, and lowers the pressure drop.

Chemical stability: the absorbent should not degrade but remain effective throughout its useful lifetime.

Flammability: if at all possible, the liquid should be **nonflammable, noncorrosive, nontoxic,** and **inexpensive.**

An important factor in the successful operation of any absorber is to initially select the proper construction materials. Quenching hot gases to their saturation temperatures forms corrosive acids. Depending on the substances present in the exhaust stream, sulfuric, hydrochloric, or hydrofluoric acids may be formed. The presence of these acids can cause severe corrosion problems unless special materials of construction are used. The construction materials typically used include rubber and PVC lined absorber vessels. Currently there is a trend toward using more absorbers made of reinforced plastics when the gaseous contaminant and/or the scrubbing solution is corrosive (McDonald, 1977).

Spray Towers or Chambers

Spray towers, the simplest devices used for gas absorption, consist of an empty tower and a set of nozzles to spray liquid. A spray tower is similar in operation to spraying water in an open barrel. Typically, the contaminant gas stream enters the bottom of the tower and passes up through the device while liquid is being sprayed

at one or more levels by nozzles. The flow of liquid and gas streams in opposite directions is referred to as *countercurrent flow*. Figure 4-16 illustrates the operation of a typical countercurrent flow spray tower.

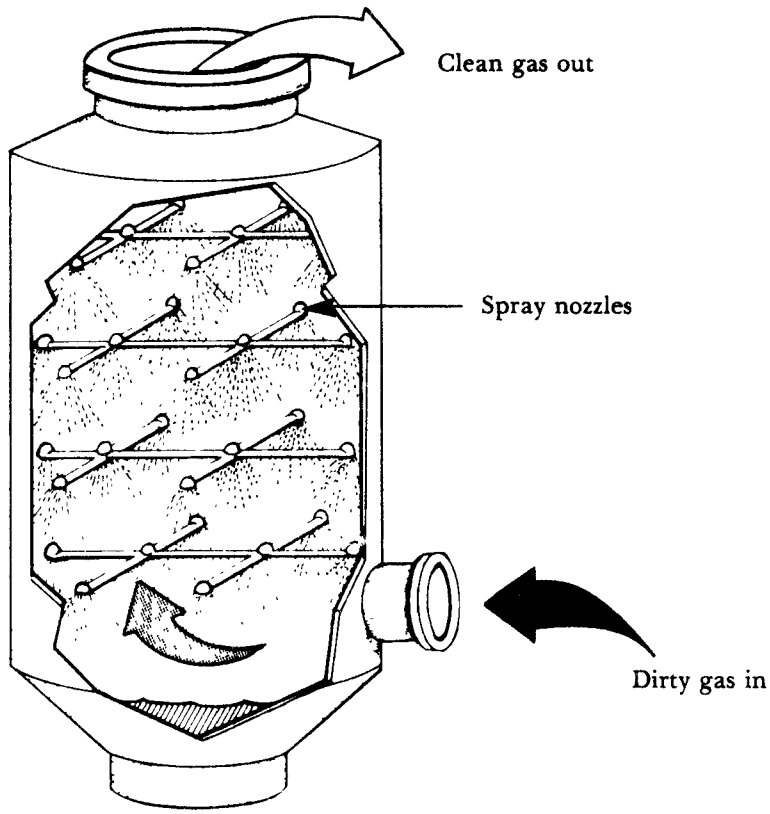


Figure 4-16. Spray tower.

To provide a large liquid surface for contacting the gas, nozzles are arranged to wet the entire cross section of the tower with fine liquid droplets. Theoretically, the height between levels of nozzles and/or the bottom of the tower determines the residence time. In practice, physical laws limit the removal efficiency of spray towers. After falling short distances the liquid droplets tend to agglomerate or hit the sides of the tower. Both of these effects reduce the total liquid surface in contact with the gas stream and the residence time. Therefore, spray chambers are limited to applications where the gases are extremely soluble or a high removal efficiency is not required.

The main advantage of spray chambers is that they are completely open; they have no internals except for the spray nozzles. Therefore, they have a very low pressure drop, about 1.25 to 4 cm (0.5 to 1.5 in.) of water over the tower. If an entrainment separator is used it will add another 2.5 to 5 cm (1 to 2 in.) of water to the total pressure drop. An entrainment separator is used to prevent mist from exhausting out the stack. Spray towers range in size from 42 to 170,000 m^3/hr (25 to 100,000 cfm) with 34,500 m^3/hr (20,000 cfm) as a typical size (EPA, 1972).

Spray towers are primarily used for gas conditioning—cooling, humidifying, or for the first stage particulate matter and gas removal. But they can be used effectively for gas absorption if the contaminant to be removed is highly soluble. For example, spray towers are used to remove HCl gas from the tail gas exhaust in the manufacturing of hydrochloric acid. In the production of superphosphate used in manufacturing fertilizer, SiF_4 and HF gases are vented from various points in the processes. Spray towers have been used to remove these highly soluble compounds. Spray towers are also used for odor removal in bone meal and tallow manufacturing industries by scrubbing the exhaust gases with a solution of $KMnO_4$. Due to their ability to handle large volumes of exhaust gases in corrosive atmospheres, spray towers are also used in a number of flue gas desulfurization systems as first or second stage removal devices.

No universally acceptable mathematical equation or correlation is presently available for sizing spray towers. Although they are simple devices, variables such as liquid drop size, settling velocity and residence time vary considerably with height or location in the tower. Tower designs are based on experimental or operational data from similar systems. However, certain limiting factors must be balanced against each other. First, the gas velocity through the tower must be kept low, around 0.3 to 1.0 m/sec (1 to 3 ft/sec) or excessive liquid entrainment occurs. This implies using a large diameter tower to keep the gas velocity low. But if the diameter is large, compared to the distance between sprays, back mixing occurs. Back mixing is a condition where the gas changes direction or swirls within the tower and removal efficiency is decreased.

The smaller the liquid droplet size, the greater the rate of absorption, since this increases the surface area available for absorption. High pressure spray nozzles are used to produce fine liquid droplets. These spray nozzles consume more power in pumping liquid than a packed tower or venturi scrubber for the same liquid rate. The fine openings of the spray nozzles are subject to erosion and plugging problems, especially if scrubbing liquid is recirculated.

Another parameter which characterizes spray tower operation is the liquid to gas ratio (L/G). By increasing the L/G ratio, absorption efficiency can be directly increased. Figure 4-17 illustrates how the percent SO_2 removal is affected by the L/G ratio. Figure 4-17 was developed by Ontario Hydro Electric from data for SO_2 removal in a spray tower using limestone on a 4000 cfm, 30 MW test unit. The disadvantage of increasing the liquid flow rate is that is adds to the operating costs both in liquid usage and power consumed.

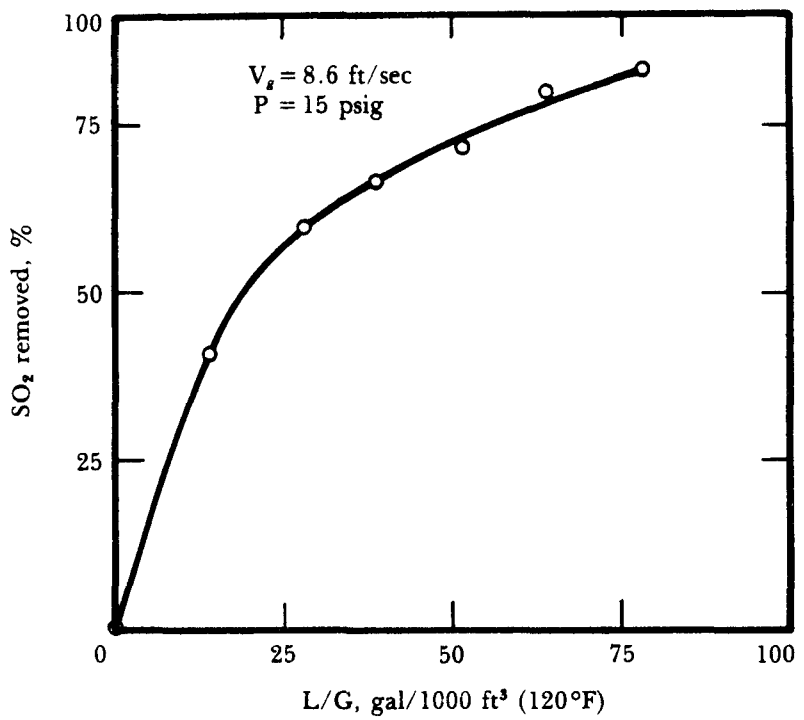


Figure 4-17. Spray tower efficiency vs. L/G.

Where there is limited space in a plant, spray chambers can operate by *crosscurrent* or *cocurrent flow*. In crosscurrent absorbers the gas flow is perpendicular to the liquid flow. In cocurrent absorbers the gas and liquid flow in the same direction. Because the gas stream does not "push" against the liquid stream as in countercurrent flow, much higher gas stream velocities can be used. Higher gas stream velocities mean that the size of the unit can be reduced. However, crosscurrent or cocurrent spray towers are not as efficient as the countercurrent type. Countercurrent flow exposes the most dilute portion of the contaminant gas stream to the freshest liquid throughout the entire tower length.

Another spray device sometimes used for gas absorption is the cyclonic scrubber pictured in Figure 4-18. The gas stream enters the device tangentially which causes it to swirl up through the tower in a corkscrew motion. Liquid is sprayed from the center outward into the swirling gas stream (vortex). This configuration allows for the use of higher gas velocities without excessive liquid entrainment. The liquid droplets are forced to the walls of the device by centrifugal force before they can be carried out the top of the device. Due to the short residence time of the liquid droplets in the gas stream, removal efficiency is very limited.

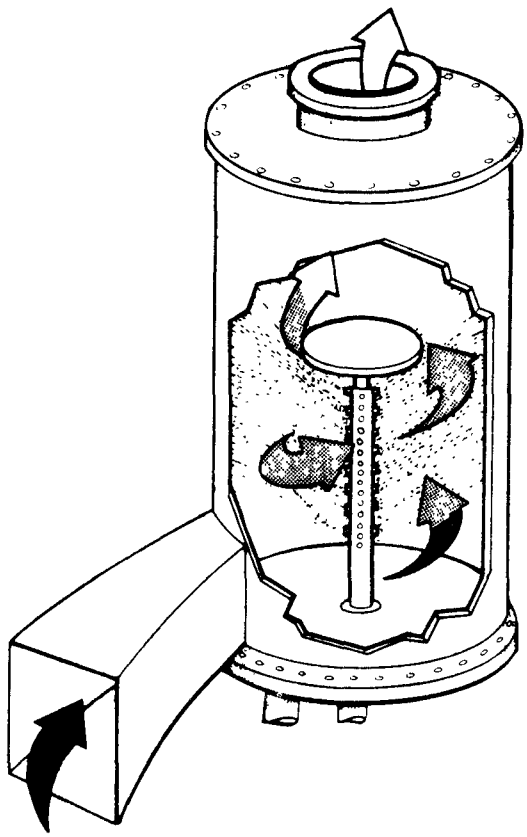


Figure 4-18. Cyclonic spray scrubber.

High Energy Absorbers

High energy scrubbers (venturis and ejectors) are primarily used for high efficiency particulate matter collection in the submicrometer range. These devices can also be used for absorption although they have a limited gas collection efficiency due to the short residence time of gas-liquid contact. For the collection of particulate matter with high energy scrubbers, increased energy input increases collection efficiency. For gas absorption, almost the direct opposite holds true. An increase in pressure drop decreases residence time, thus reducing the amount of mass transfer.

The most common high energy scrubber is the venturi. A typical venturi configuration is shown in Figure 4-19. A venturi consists of a converging section through which the gas enters and is accelerated to a high velocity. The gas shears the liquid from the walls of the throat section producing an enormous number of fine droplets. The throat section is a constant cross sectional area where the majority of absorption occurs. The gas then exits through an expander section where it is slowed down, reducing energy consumption. Simple venturis, as shown in Figure 4-19, can be of two types depending on where the liquid is injected. If the liquid is injected at the throat section it is referred to as a *nonwetted approach venturi*. If the liquid is introduced through pipes at the entrance of the converging

section it is called a *wetted approach venturi*. Wetted approach venturis are used to saturate the incoming gas stream. This eliminates the possibility of having a **wet-dry interface** area at the entrance to the throat. At these wet-dry interface areas, scale build-up can eventually cause plugging problems. The wetted approach is usually more expensive than nonwetted approach units, so if the inlet gas is already saturated a nonwetted approach may be preferred (Brady, 1977).

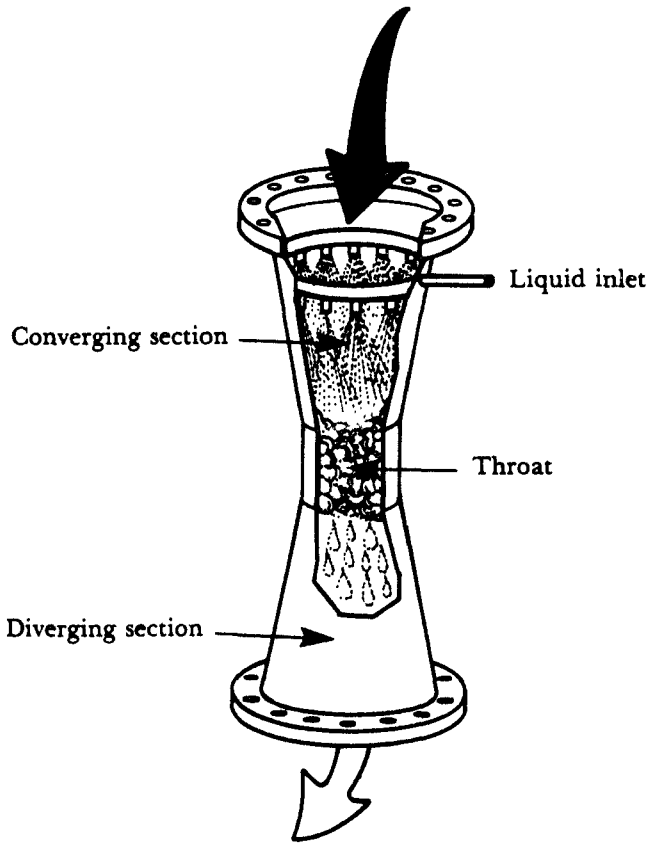


Figure 4-19. Typical venturi configuration.

The simple venturi shown in Figure 4-19 is adequate for flows up to 88,000 m^3/hr (40,000 acfm). At flows greater than this it is difficult to achieve uniform liquid distribution unless additional weirs or baffles are added. To handle the larger gas flows, manufacturers have gone to a rectangular configuration with a long narrow throat as shown in Figure 4-20 (Brady, 1977). Other types of venturis can have variable throat openings. An adjustable plumb bob in the throat section is used to accommodate variable flows (Figure 4-21). Water is sprayed on the plumb bob in the throat section, spreads out and is sheared off by the gas. The plumb bob moves up or down depending on the gas flow rate and the desired pressure drop.

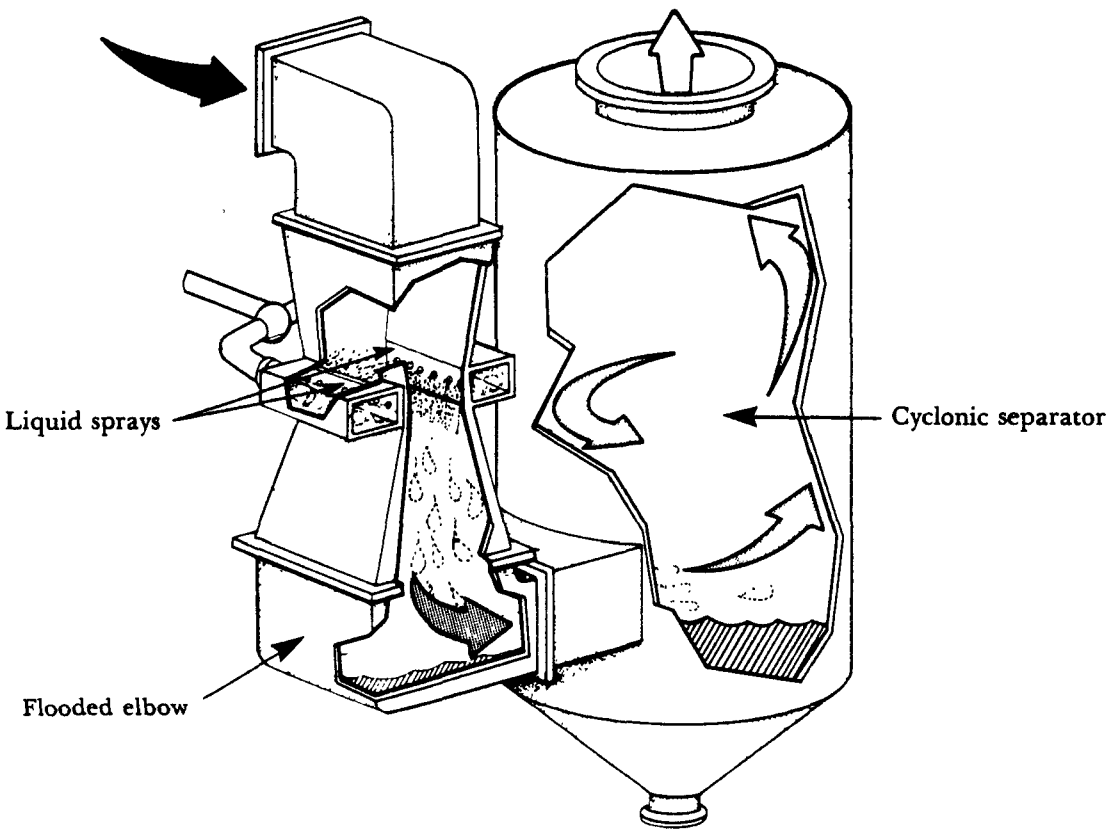


Figure 4-20. Spray venturi.

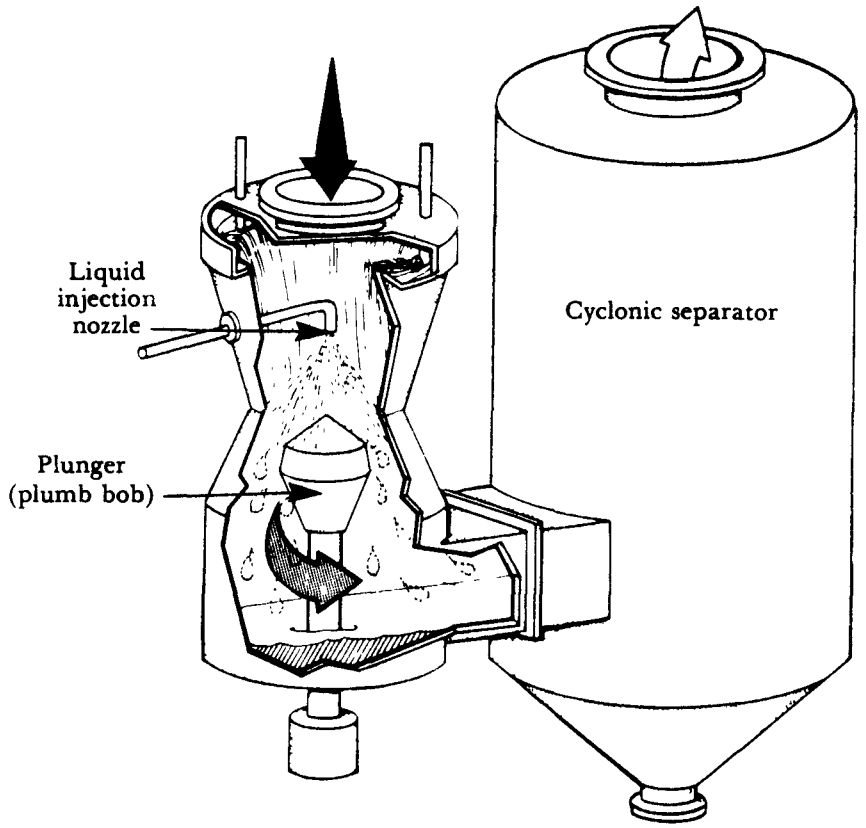


Figure 4-21. Adjustable throat venturi.

Venturi scrubbers normally operate with high gas side pressure drop, ranging from 50 to 250 cm (20 to 100 in.) of water. These high pressure drops result both in high energy costs and high maintenance costs for the fans needed to move the gas stream. The inlet gas velocity is greatest through the throat section of the venturi. In the throat, gas velocities can vary between 30 to 125 m^3/s (100 to 400 ft/sec). These high velocities are good for particulate matter removal, but result in insufficient contact time for most absorption applications. High gas velocities also make it mandatory that some type of entrainment separator be used (Figure 4-20). The entrainment separator will eliminate any liquid carryout in the exhaust from the scrubbing system.

To overcome the short residence time, most venturi systems operate at high liquid-to-gas (L/G) ratios when used for gaseous emission control. Typical liquid-to-gas ratios for venturis used as absorbers range between 47 and 105 liters per min/100 m^3 per hr (20 to 40 gpm/1000 cfm) and in some instances up to 195 liters per min/1000 m^3 per hr (80 gpm/1000 cfm). For particulate matter removal, venturis normally operate between 4.7 and 38 liters per min/1000 m^3 per hr (2 to 15 gpm/1000 cfm).

Another type of venturi is the jet or ejector venturi (Figure 4-22). This device uses high pressure spray nozzles (up to 100 psig) aimed at the throat section of a venturi. Using high pressure sprays in this manner serves two purposes. First the movement of the liquid creates a suction (vacuum) which pulls the gas stream through the venturi. This eliminates the need of a fan or blower to move the gas stream. Secondly, the high pressure sprays, along with the venturi effect, form numerous fine liquid droplets which provide a high degree of turbulence between gas and liquid phases. This limits contact time, therefore, absorption efficiency is usually low.

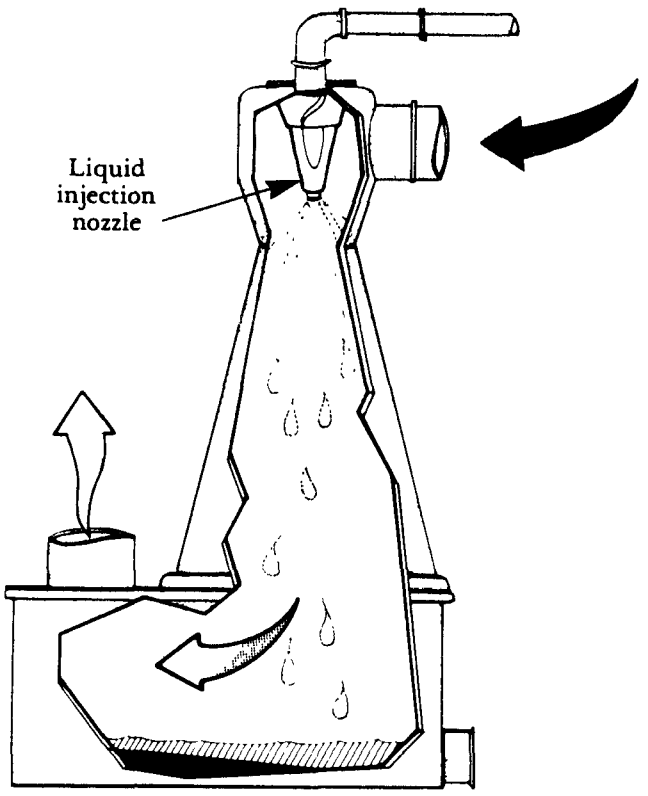


Figure 4-22. Jet or ejector venturi.

Ejector venturis operate at high liquid-to-gas ratios (around 100 gpm/1000 cfm). The gas side pressure drop usually ranges between 10 to 20 cm (4 to 8 in.) of water. The overall power consumption, however, is much higher, due to liquid pumping requirements. Ejector venturis can be designed for smaller gas flow rates (under 1000 cfm) than most other absorbers are capable of handling.

Venturi scrubbers are the most widely used wet collector for particulate matter removal. Unless a gaseous contaminant is extremely soluble, venturis are seldom used strictly as an absorber. A variety of venturis are used by power plants to remove both flyash and sulfur dioxide. To achieve the required SO_2 removal efficiencies venturis are usually used in combination with another absorber. Venturis have been used with spray towers, packed towers, and other venturis to obtain the

desired removal efficiency. Venturi scrubbers are also used to control emissions from the smelting of lead, zinc and copper. Smelting operations typically produce emissions of very fine acid mists, high concentrations of sulfur dioxide, and a variety of particulate matter.

Packed Column

The packed column (tower) is the most common scrubber used for gas absorption. Packed columns disperse the scrubbing liquid over packing material which provides a large surface area for continuous gas-liquid contact. Packed towers are classified according to the relative direction of gas-to-liquid flow.

The most common packed tower is the countercurrent (gas-to-liquid) flow tower, shown in Figure 4-23. The gas stream being treated enters the bottom of the tower and flows upward over the packing material. Liquid is introduced at the top of the packing by sprays or weirs and flows downward over the packing material. This flow arrangement results in the highest theoretical achievable efficiency. The most dilute gas is contacted with the purest absorbing liquor, providing a constant, maximized concentration difference (driving force) for the entire length of the column.

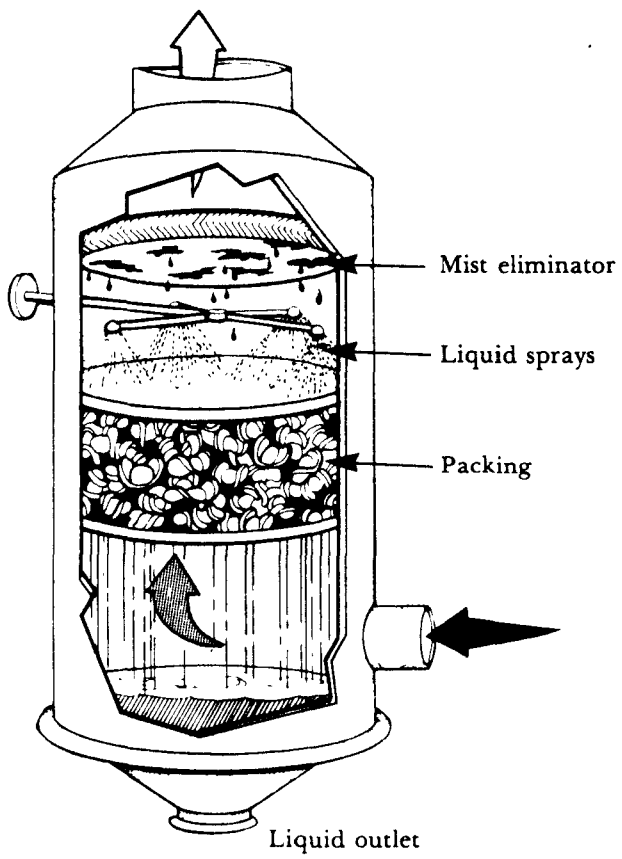


Figure 4-23. Countercurrent packed tower.

Pressure drop in the gas phase ranges from 0.63 to 2.54 cm (0.25 to 1 inch per ft) of water per meter of column packing at 40 to 70% of the flooding velocity. This absorber cannot handle large variations in liquid or gas flow rates or loading and flooding will occur. Loading and flooding conditions were described in the preceding design section on packed tower diameter.

In another flow arrangement used with packed towers, cocurrent flow, both the gas and liquid phases enter at the top of the absorber and move downward over the packing material. This allows the absorber to be operated at higher liquid and gas flow rates since flooding is not a problem. The pressure drop is lower than with countercurrent flow since both streams move with gravity. The major disadvantage is that removal efficiency is very limited due to the decreasing driving force (concentration differential) as the streams travel down through the column. This limits the areas of application for cocurrent absorbers. They are used almost exclusively in situations where limited equipment space is available since the tower diameter is smaller than a countercurrent or plate tower for equivalent flow rates.

In a cross flow absorber, the gas stream flows horizontally through the packed bed which is irrigated by the scrubbing liquid flowing down through the packing material perpendicular to the gas flow. A typical cross flow absorber is shown in Figure 4-24. Inlet sprays aimed at the face of the bed may also be included. These sprays scrub both the entering gas and the face of the packed bed. The leading face of the packed bed is slanted in the direction of the on-coming gas stream as shown in Figure 4-24. This insures complete wetting of the packing by allowing the liquid at the front face of the packing time to drop to the bottom before being pushed back by the entering gas.

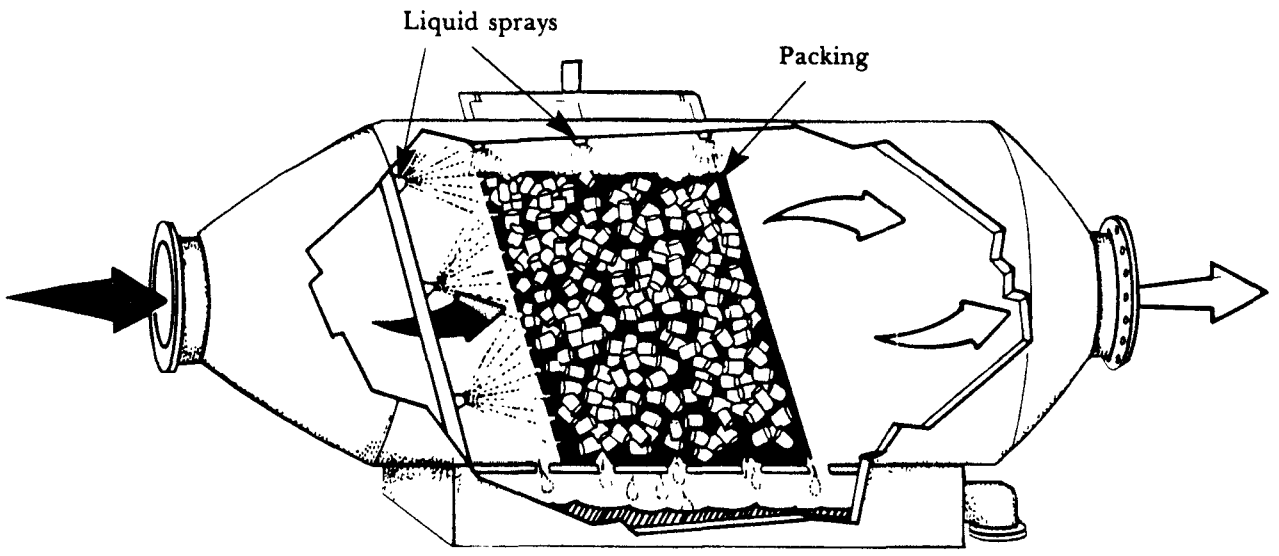


Figure 4-24. Cross flow packed tower.

Cross flow absorbers are smaller and have a lower pressure drop than any other packed or plate tower for the same application (removal efficiency and flow rates). In addition, they are better suited to handle exhaust streams with heavy particulate matter concentration. By adjusting the liquid flow rate, incoming particulate matter can be removed and washed away in the front half of the bed. This also results in a liquid savings by enabling the absorber to use less liquid in the rear sprays. This practice is carried one step further by actually constructing the absorber into sections as shown in Figure 4-25. The front section can be equipped with water sprays and used for particulate matter removal. In the second section, sprays may contain a reagent in the scrubbing liquor for gas removal. The last section is left dry to act as an entrainment separator. Cross flow absorbers do require complex design procedures since concentration gradients exist in two directions in the liquid; from top to bottom and from front to rear.

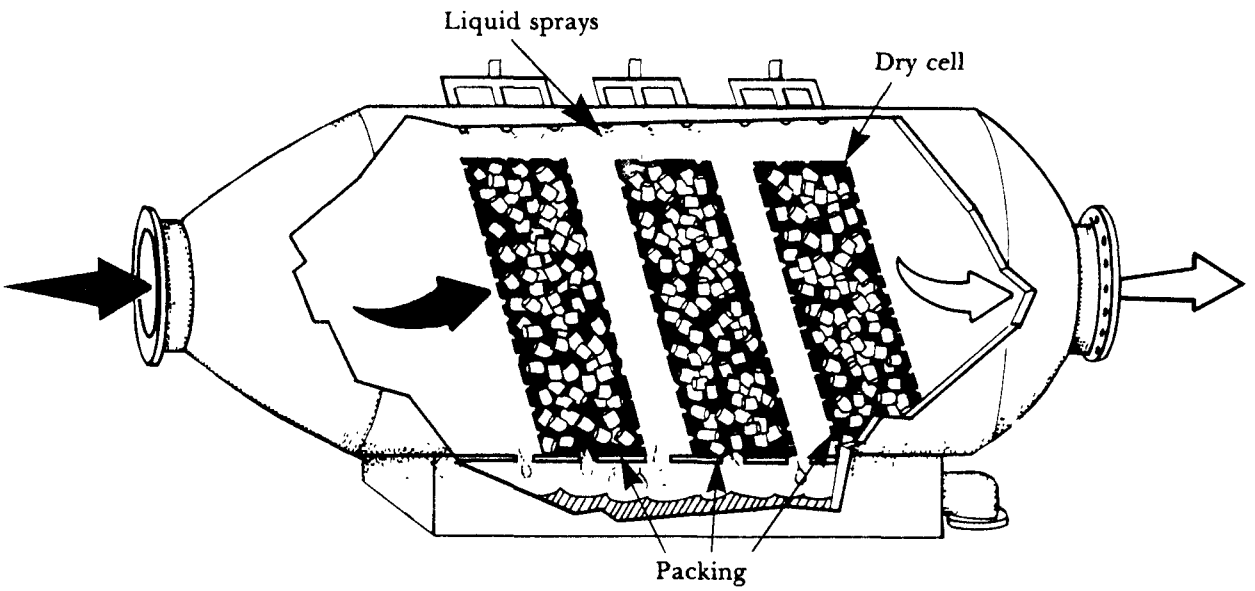


Figure 4-25. Three bed cross flow packed tower.

Packed towers are most suited to applications where a high gas removal efficiency is required and the exhaust stream is relatively free from particulate matter. In the production of both sulfuric and hydrochloric acids, packed towers are used to control tail and exhaust emissions (SO_2 and HCl respectively). The scrubbing liquor for these processes can be a weak acid solution, with the spent liquor from the packed tower sent back to the process. Packed towers are also used to control HCl and H_2SO_4 fume emissions from pickling operations in the primary metals industry. They are also used to control odors in rendering plants, petroleum refineries, and in sewage treatment plants. For odor control, the scrubbing liquor usually contains an oxidizing reagent such as potassium permanganate or sodium hypochlorite. In these applications, an acid backwash must be added if a precipitate is formed or plugging can be a problem. The gas flow rate through packed towers can vary from 17 to 51,000 m^3/hr (10 to 30,000 cfm) with 8,600 m^3/hr (5100 cfm) as an average size (EPA, 1972).

Packing Material

Packing material is the heart of the absorber. It provides the surface over which the scrubbing liquid flows, presenting a large area for mass transfer to occur. Packing material represents the largest material cost of the packed tower. Pictured in Figure 4-26 are some of the more commonly used packings which are made in numerous geometric shapes and sizes. These materials were originally made of stoneware, porcelain, or metal; but presently a large majority are being made of high-density thermoplastics (polythylene and polypropylene). A specific packing is described by its trade name and overall size. For example, a column can be packed with 5 cm (2 inch) Raschig rings or 2.5 cm (1 inch) Tellerettes. The overall dimension of packing materials normally range from 0.6 to 10 cm (¼ to 4 inches).

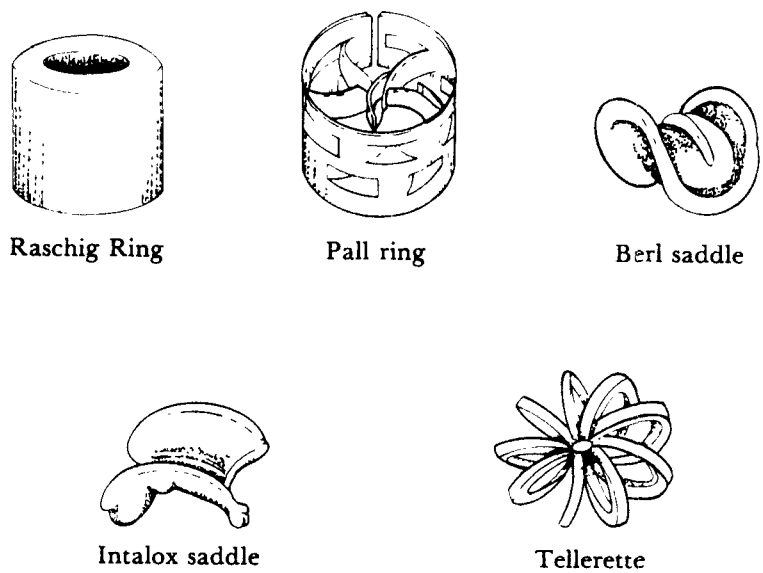


Figure 4-26. Common packing materials.

The specific packing that is selected depends on the nature of the contaminants, geometric mode of contact, size of the absorber, and scrubbing objectives. The following factors provide a general guide for selecting packing materials (McDonald, 1977). (For some specific packing materials, these factors are listed in Table 4-3).

Cost: Generally plastic packings are cheaper than metal with ceramic being the most expensive. Cost of packings are given in dollars per cubic meter ($\$/m^3$).

Low pressure drop: Pressure drop is a function of the volume of void space in a tower when filled with packing. Generally, the larger the packing size, the smaller the pressure drop.

Corrosion resistant: Ceramic or porcelain packing is commonly used in a very corrosive atmosphere.

Large specific area: A large surface area per cubic foot of packing, m^2/m^3 (ft^2/ft^3), is desirable for mass transfer.

Structural strength: Packing must be strong enough to withstand normal loads during installation, service, physical handling, and thermal fluctuations. Ceramic packing is subject to cracking under sudden temperature changes.

Weight: Heavier packing may require additional support materials or heavier tower construction. Plastics have a big advantage in this area since they are much lighter than either ceramic or metal packings.

Design flexibility: The efficiency of a scrubber changes as the liquid and gas flow rates are varied. Packing material must be able to handle the process changes without substantially affecting the removal efficiency.

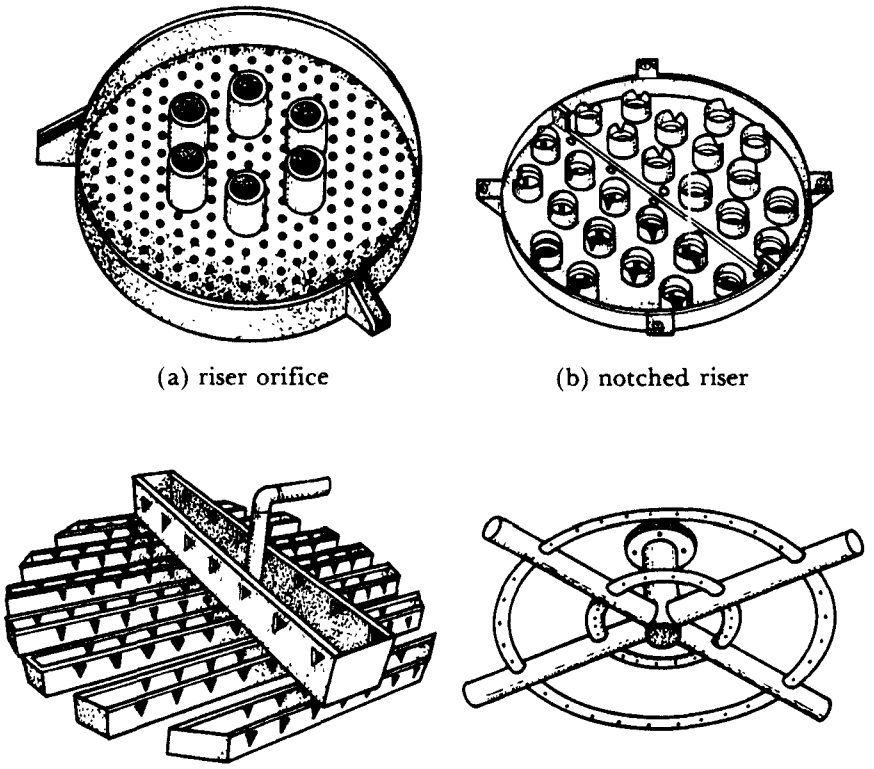
Packing material may be arranged in an absorber in either of two ways. The packing may be dumped into the column randomly or in certain cases systematically stacked, as bricks are laid atop each other. Randomly packed towers provide a higher surface area, m^2/m^3 (ft^2/ft^3), but also cause a higher pressure drop than stacked packing. In addition to the lower pressure drop, the stacked packing provides better liquid distribution over the entire surface of the packing. The large installation costs required to stack the packing material usually make it impractical unless high flow rates are required.

Liquid Distribution

As stated previously, one of the keys to effective absorber operation is to intimately contact the gas stream with the liquid stream. This contact must be maintained throughout the **entire** column length. No packing material will adequately distribute liquid which is poured onto it at only one point. Liquid introduced into the absorber in this manner tends to flow down over a relatively small cross section of the tower diameter. Known as channeling, the liquid flows in little streams down through the tower without wetting the entire packing area. Liquid should be distributed over the entire cross sectional top of the packing. This is commonly achieved by weirs or feed tube arrangements as shown in Figure 4-27. Although liquid sprays are also used, weirs or tube arrangements require less power, are more flexible to changing liquid loads, and handle recycled liquid better than sprays.

Once the liquid is distributed over the packing, it flows down, by the force of gravity, through the packing, following the path of least resistance. The liquid tends to flow towards the tower wall where the void spaces are greater than in the center. Once the liquid hits the wall, it flows straight down the tower from that point (channels). A way must be provided to redirect the liquid from the tower wall back to the center of the column. This is usually done by using liquid redistributors which funnel the liquid back over the entire surface of packing. It is recommended that redistributors be placed at intervals of no more than 3 meters (10 feet) or 5 tower diameters, whichever is smaller (Zenz, 1972).

Uniform distribution of the feed gas is also very important. This is accomplished by properly designing the support trays which hold up the packing material. If the tower is broken into sections, each support grid would act as a distribution baffle.



(a) riser orifice (b) notched riser

(c) trough and weir type (d) perforated pipe

Figure 4-27. Liquid feed distributors.

Plate Tower

A plate tower is a vertical column with one or more plates (or trays) mounted horizontally inside. The gas stream enters at the bottom and flows upward passing through openings in the plates. Liquid enters at the top of the tower and travels across each plate, then through a downcomer to the plate below until it reaches the bottom of the tower. Mixing occurs as the gas bubbles up through the layer of liquid covering each plate. Figure 4-28 illustrates the flow pattern of a typical plate tower.

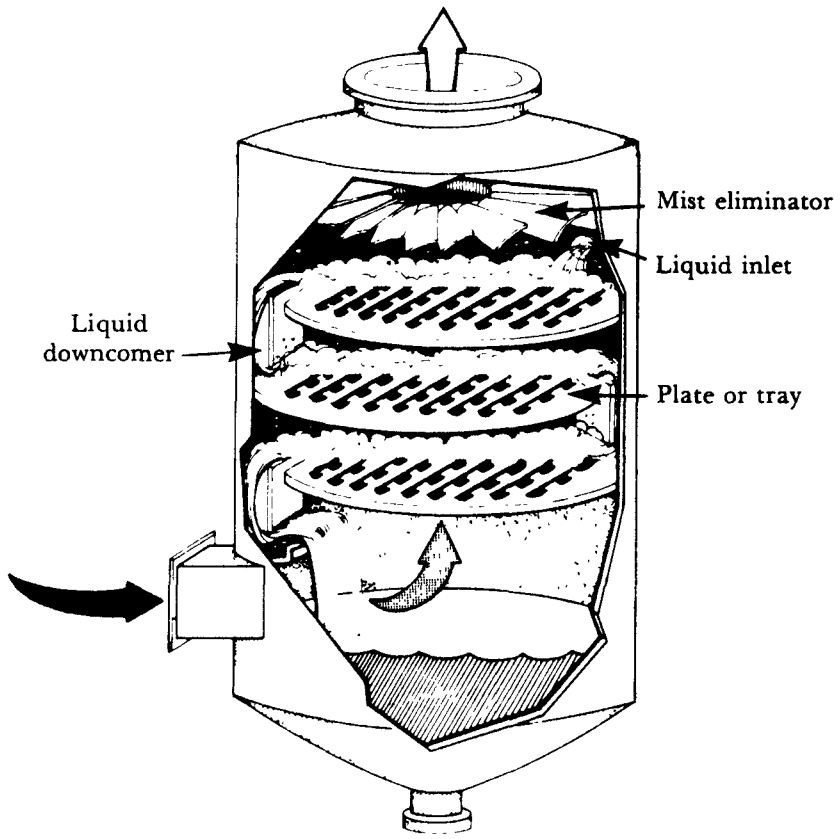


Figure 4-28. Plate tower.

Mass transfer occurs as the gas stream bubbles up through the openings and into the liquid layer on the plate. The smaller the bubbles and the more numerous they are, the more effective the mass transfer. The formation of smaller and more numerous bubbles promotes better mixing and exposes a larger gas-to-liquid surface area. As the gas disengages from the liquid, a froth is formed atop the liquid. Ideally, the gas and liquid leaving each plate are in equilibrium with each other at the conditions of that particular plate. Each plate acts as a separate absorption stage. The amount of absorption that can occur at one plate or stage is referred to as a transfer unit.

The function of the plates or trays is to disperse the gas into numerous bubbles, exposing a large surface for mass transfer. As with packing material a variety of different plates are in use, most commonly:

Bubble cap: Vapor enters the cap through *risers* and out of the slots as bubbles into the surrounding liquid on the plate (Figure 4-29a). The main advantage is that this design can handle a wide range of gas and liquid rates without affecting efficiency. Plugging and corrosion can be a problem.

Sieve trays: Vapor rises through small holes 0.60 to 2.50 cm in diameter (¼ to 1 inch) in the plate and bubbles through the liquid on the plate (Figure 4-29b). Trays can contain 600 to 3000 holes per square foot of surface. The installed costs are lower than for any other type of plate and the fouling tendency (with large holes) is lower.

Valve trays: Vapor flows up through small holes and lifts up metal valves or caps that cover the openings in the plate. The valves are restrained by legs which limit vertical movement (Figure 4-29c). The liftable caps act as variable orifices; adjusting the opening proportional to vapor flow. Caps are arranged in different weights to provide flexibility. Valve trays are more expensive than sieve trays but cheaper than bubble cap trays. They can handle high liquid-to-gas ratios. Wear or corrosion of retaining legs can be a problem.

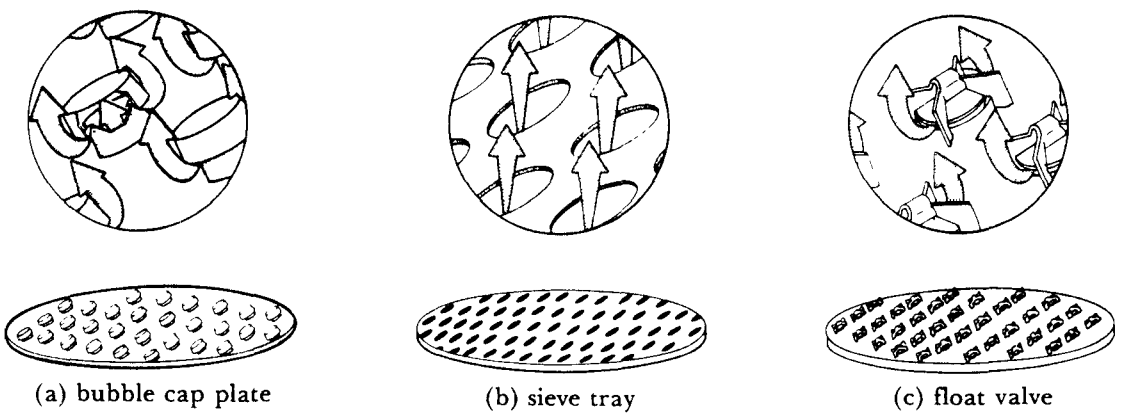


Figure 4-29. Scrubbing plates or trays.

Plate towers, like packed towers, can achieve a high removal efficiency even if the gaseous contaminant is relatively insoluble. Therefore, plate towers are used to control emissions from many of the same processes that use packed towers, such as from acid plants, fertilizer production plants, chemical industry and petroleum refineries.

Several advantages and disadvantages must be considered when choosing a plate tower instead of a packed tower for a given control operation. The following list gives some factors used in comparing plate towers to packed towers.

1. Plate towers are able to handle particulate matter and other solids better than packed towers. Maintenance openings can be installed so that the plates may be easily cleaned.
2. Plate towers are chosen for operations that require a large number of transfer units or that must handle large gas volumes. Packed towers can experience channeling problems if the diameter or height of the tower is too large. Redistribution trays must be installed in large diameter and tall packed towers to avoid channeling.
3. The total weight of a plate tower is less than that of a comparable packed tower.
4. Packed towers are much cheaper to construct if corrosive substances are to be handled. Packed towers can be constructed with a fiberglass reinforced polyester shell which is generally about half the cost of a carbon steel plate tower.
5. Plate towers can handle volume and temperature fluctuations better than packed towers. Expansion or contraction due to temperature changes can crush or melt packing material.

Mobile Packed Beds

Mobile packed beds are very similar to packed towers. However, instead of being stationary, they provide a zone of moving packing material. The gas and the liquid streams are mixed in this moving zone where mass transfer occurs. These scrubbers are primarily used when both particulate and gaseous contaminant removal is necessary. They provide the mass transfer efficiency of a packed or plate tower without the plugging problems. They do not, however, have the efficiency of a packed tower per energy unit consumed. Mobile packed beds are classified as either a flooded bed or a fluidized bed, depending on the degree of packing movement. The gas and liquid streams can flow either cocurrently or countercurrently.

The flooded bed absorber pictured in Figure 4-30 is made up of one section of mobile spheres 10 to 20 cm (4 to 8 inches) deep. The spheres are usually made of plastic, however, glass or marble spheres are also used. The gas stream enters from the bottom while liquid is sprayed from the top and/or bottom over the packing. The gas velocity is such that it causes the packing to rotate in a gentle rubbing motion, but does not fluidize the packing. The rotating movement acts as a self-cleaning mechanism plus enhances gas and liquid mixing. Bubbles formed in the bed create a layer of froth over each section of bed which is about twice as high as the bed itself.

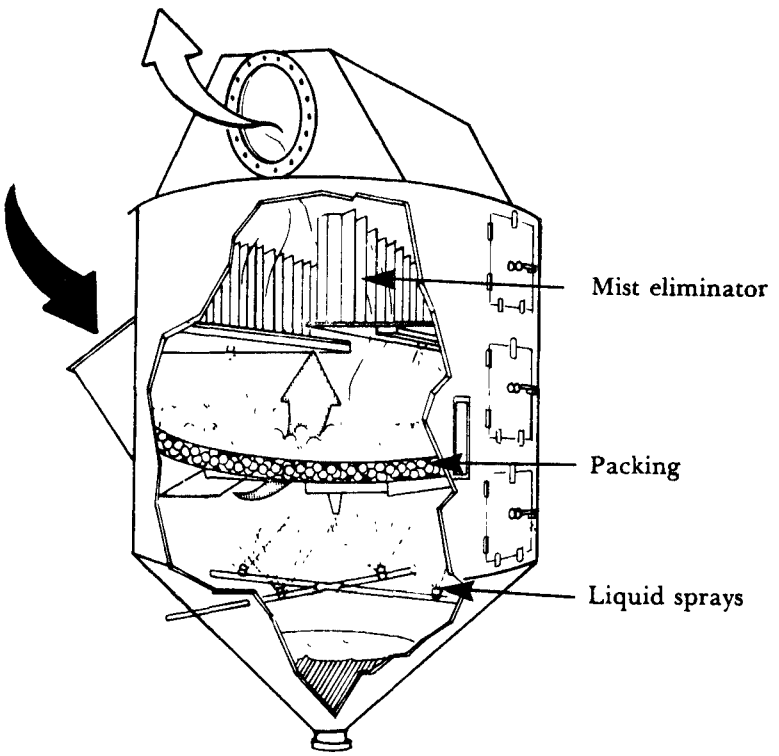


Figure 4-30. Flooded bed absorber.

A fluidized bed absorber is very similar to a flooded bed absorber. The difference is in the degree of movement of the packing. In a fluidized bed absorber, the gas velocity is such (1.8 to 2.4 m/sec) (6 to 8 ft/sec) that it keeps the packing in constant motion between a lower and upper retaining grid. This is shown in Figure 4-31. The packing is made of either polypropylene or polyethylene plastic balls that are hollow, resembling a ping pong ball. The packed sections are usually 0.3 to 0.6 m (1 to 2 ft) thick with a froth zone about 0.6 meters (2 ft) thick above the packing.

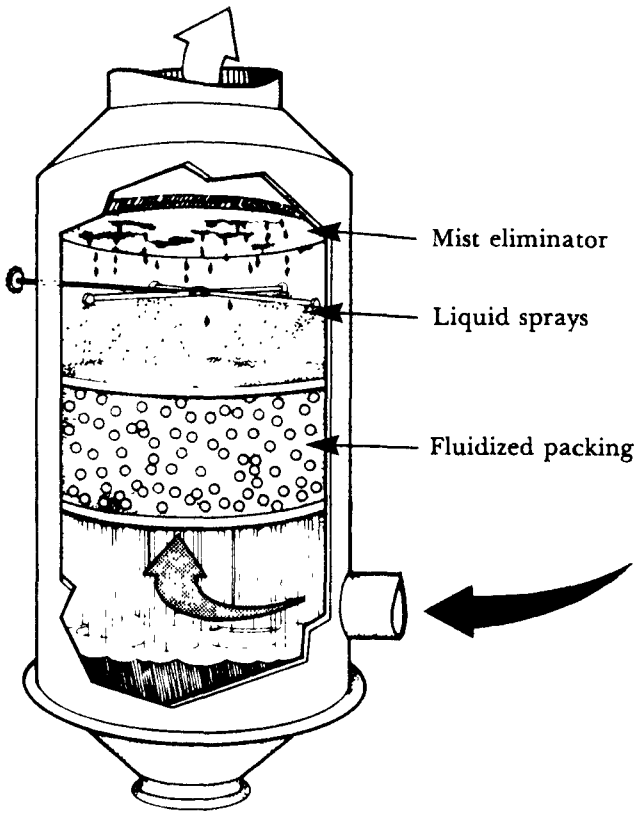


Figure 4-31. Fluidized bed absorber.

The pressure drop in these mobile packed beds ranges from 5 to 15 cm (2 to 6 inches) of water per stage of packing. The capacity of these units can be 5 to 6 times that of a packed or plate tower of the same diameter due to the high gas velocities through the units (Bethea, 1978). Mobile packed beds are used to control emissions from exhaust streams that require a high degree of gaseous contaminant removal and particulate matter removal. Mobile packed beds have been used to control emissions from Kraft pulp mills, cupola furnaces, and aluminum foundries. These units have also been used for SO_2 and flyash control from large power plants.

Entrainment Separators

Gas moving at high velocities that mixes with a liquid will entrain drops of that liquid. The liquid droplets must be removed from the gas stream before being exhausted to the atmosphere. Entrainment separators are used to remove the liquid droplets. Although the major function of an entrainment separator is to prevent liquid carryover, it also performs additional scrubbing and recovers the scrubbing liquor which saves on operating costs. Entrainment separators are therefore usually an integral part of any wet scrubbing system.

Entrained liquid droplets vary in size depending on how the droplets were formed. Drops that are torn from the body of a liquid are large (10 to 100 micrometers in diameter); whereas drops that are formed by a chemical reaction or by condensation are on the order of 5 micrometers or less in diameter. Numerous types of entrainment separators are capable of removing these droplets. Those most commonly used for air pollution control purposes are the *cyclone, wire mesh pads,* and *blade types.*

The cyclone (centrifugal) separator is a cylindrical tank with a tangential inlet or turning vanes. The inlet or vanes impart a swirling motion to the droplet laden gas stream. The droplets are thrown outward by centrifugal force to the walls of the cylinder. Here they coalesce and drop down the walls to a central location and are recycled to the absorber. These units are simple in construction, having no moving parts. Therefore, they have few plugging problems. Good separation down to the 10 micrometer range can be expected. The pressure drop across the cyclone is 10 to 15 cm (4 to 6 inches) of water for a 98% removal efficiency of droplets down to the 20 to 25 micrometer range. Cyclonic separators are commonly used with venturi scrubbers as seen in Figures 4-20 and 4-21.

Wire mesh or plastic pad separators consist of woven material about 10 to 15 cm (4 to 6 inches) thick that fit across the entire diameter of the scrubber (Figure 4-32). The mesh allows droplets to impact on the material surface, agglomerate with other droplets and drain off by gravity. The pad is usually slanted (no more than a few degrees) to permit the liquid to drain off. Essentially 100% collection of droplets larger than 3 micrometers is obtained with pressure drops of approximately 10 to 15 cm (4 to 6 inches) of water (the pressure drop is dependent on depth and compaction of fibers). The disadvantage with mesh pads is that their small passages are subject to plugging.

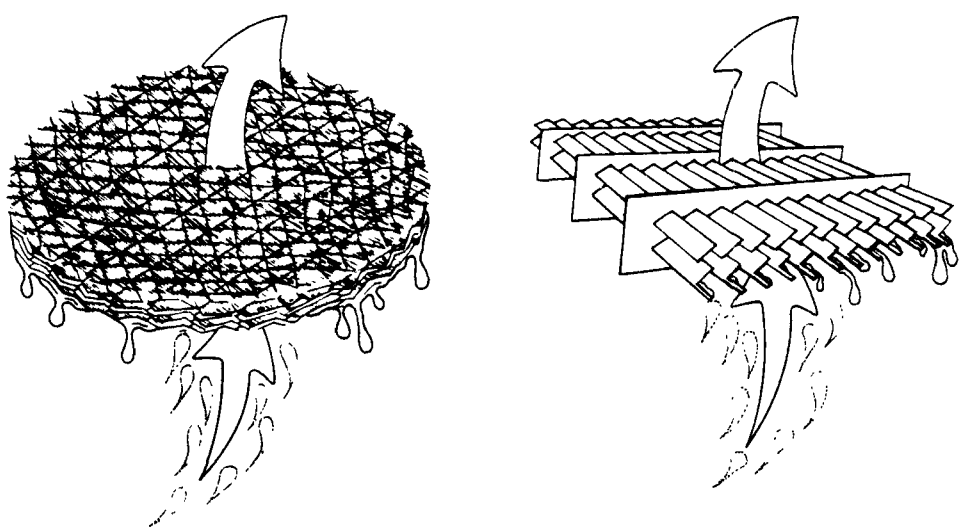


Figure 4-32. Mist eliminators.

Impingement blade separators employ blades bent into "S" curves or a *chevron* configuration (Figure 4-32). As gas passes between the blades, it is forced to travel in a zigzag pattern. The liquid droplets cannot follow the flow lines and impinge on the blade surface, coalesce and fall back into the scrubber chamber (illustrated in Figure 4-30). The pressure drop is approximately 6.4 cm (2.5 inches) of water for the capture of droplets as small as 5 micrometers in diameter.

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Chapter 5
Adsorption

Introduction

During adsorption, one or more gaseous components are removed from an effluent gas stream by adhering to the surface of a solid. The gas molecules being removed are referred to as the *adsorbate*, while the solid doing the adsorbing is called the *adsorbent*. Adsorbents are highly porous particles. Adsorption occurs on the internal surfaces of the particles as illustrated in Figure 5-1.

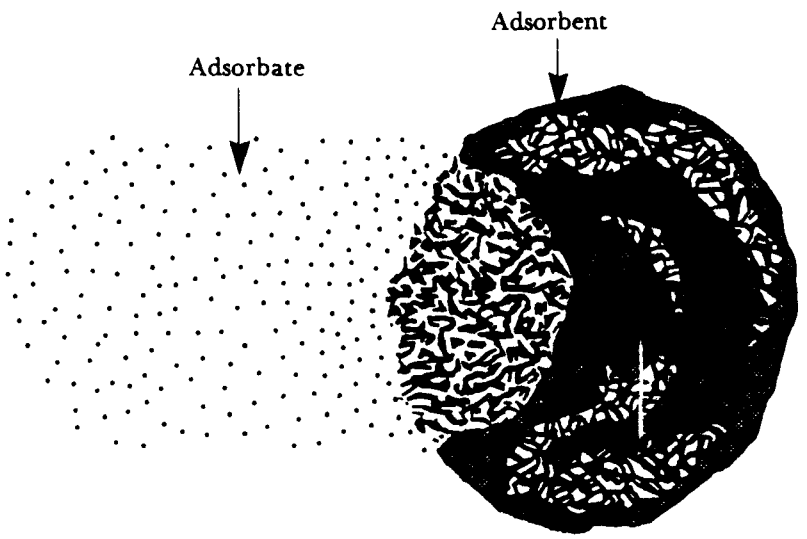


Figure 5-1. Vapor adsorbed into pores of activated carbon.

The attractive forces which hold the gas to the surface of the solid are the same that cause vapors to condense (van der Waals' forces). All gas-solid interfaces exhibit this attraction, some more than others. Adsorption systems use materials which are highly attracted to each other to separate these gases from the non-adsorbing components of an air stream. For air pollution control purposes, adsorption is not a final control process. The contaminant gas is merely **stored** on the surface of the adsorbent. After it becomes saturated with adsorbate, the adsorbent must either be disposed of and replaced, or the vapors must be **desorbed**. Desorbed vapors are highly concentrated and may be recovered more easily and more economically than before the adsorption step.

Traditionally, adsorption has been used for **air purification** and **solvent recovery**. Air purification processes are those in which the contaminant is present in trace quantities (less than 1.0 ppm) but is highly odorous or toxic. Systems used

for air purification are small *thin bed* adsorbers. When the bed becomes saturated with contaminant, it is taken out and replaced. Solvent recovery processes require much larger systems and are usually designed to control organic emissions whose concentrations are greater than 1000 ppm. This has been the point where the recovery value of the solvent could justify the expense of the large adsorption-desorption system. Currently, adsorption is used as a method of recovering valuable organic vapors from flue gases at all concentration levels. This is due to present regulations limiting volatile organic emissions and the higher costs of solvents.

Theory of Adsorption

Mechanism of Adsorption

Adsorption occurs by a series of three steps. In the first step, the contaminant diffuses from the major body of the air stream to the external surface of the adsorbent particle. In the second step, the contaminant molecule migrates from the relatively small area of the external surface (a few m^2/g) to the pores within each adsorbent particle. The bulk of adsorption occurs in these pores because the majority of available surface area is there (hundreds of m^2/g). In the third step, the contaminant molecule adheres to the surface in the pore. Figure 5-2 illustrates this overall diffusion and adsorption process.

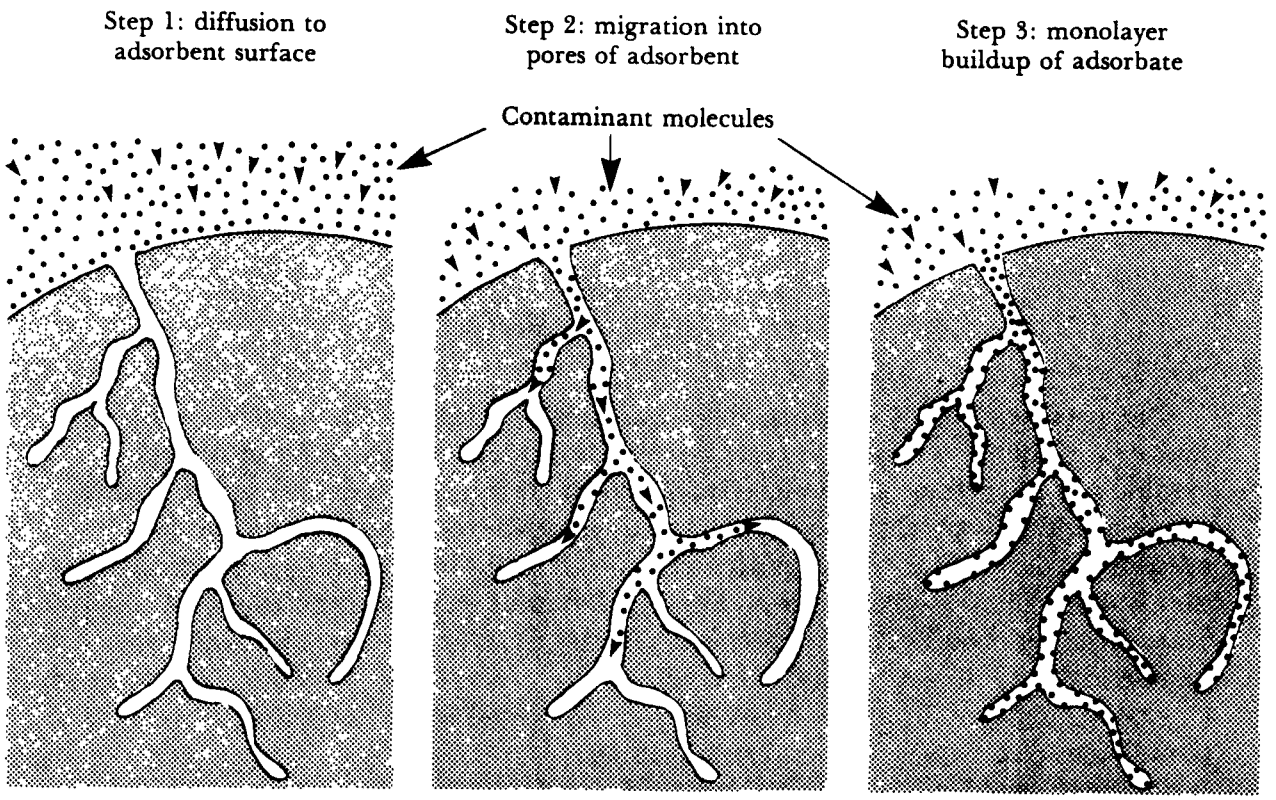


Figure 5-2. Mechanism of adsorption.

The purpose of analyzing the mechanism of adsorption is to determine which step controls the overall process. By describing this step mathematically, adsorber performance can be predicted from physical data. The actual adsorption of a molecule, step 3, proceeds relatively quickly compared to steps 1 or 2. Therefore, step 3 can be ignored when developing design equations. Steps 1 and 2 are both diffusional processes. They involve the transport of the adsorbate through a carrier gas phase to an adsorption site. In the first step, diffusion occurs because of a concentration difference. The rate of mass transferred by this type of diffusion can be predicted from Equation 5-1.

Equation 5-1 is based on a film resistance theory of bulk diffusion as presented in Chapter 4 on absorption. Bulk diffusion assumes that the only resistance a gas molecule encounters in movement through the carrier gas stream occurs during collisions with other gas molecules.

(Eq. 5-1)

$$N_A = \frac{k_g A \beta (p - p_i)}{\varrho_B}$$

Where: N_A = rate of mass transfer, kg mol/s
k_g = local mass transfer coefficient, kg mol/s•m²•Pa
β = void area between adsorbent granules, m³/m³
A = surface area of adsorbent, m²/kg
p = partial pressure of adsorbate in gas phase, Pa
p_i = partial pressure of adsorbate at the gas-solid interface, Pa
ϱ_B = bulk density of adsorbent, kg/m³

The mass transfer coefficient (k_g) is a function of the velocity, viscosity, and density of the carrier gas stream; the diffusivity of the gas molecule that is being adsorbed; and the diameter of the adsorbent. Equations to estimate the transfer coefficient based on these parameters can be found in Perry (1973).

Once the gas molecule has reached the external surface of the adsorbent, it must diffuse (move) into the pores. Diffusion in the pores of the adsorbent can occur by a number of different diffusion mechanisms depending on the size of the pore. When the pores are large, bulk diffusion predominates. As the pores begin to narrow, collisions with the wall of the adsorbate become more likely than intermolecular collisions. The gas molecules strike the wall, remain for a short time, then return to the gas phase. This is termed Knudson (or molecular) diffusion and occurs much more slowly than does bulk diffusion for a given pore length. Finally, in the smallest pores, surface diffusion is the predominant mechanism of gas transport. Gas molecules can either migrate along the surface of the solid or jump between adsorption sites.

Due to these varied mechanisms by which diffusion occurs, mass transfer rates in the pores are extremely difficult to predict. Unless extensive data are available concerning the specific adsorption application, determining the rate-controlling step (step 1 or step 2) is impossible.

One approach to determining the mass transfer rate is to rewrite Equation 5-1 in terms of an overall mass transfer coefficient.

(Eq. 5-2) $$N_A = K_{OG}a(p - p^*)$$

Where: a = external adsorbent area, m^2/m^3
p^* = partial pressure in equilibrium with the surface concentration of adsorbate, Pa
p = partial pressure of adsorbate in the gas phase, Pa
K_{OG} = overall mass transfer coefficient, kg mol/h•m²•Pa

The overall mass transfer coefficient represents the resistance to molecular motion encountered both outside and inside the pore.

(Eq. 5-3) $$\frac{1}{K_{OG}} = \frac{1}{k_g} + \frac{1}{k_s}$$

Where: k_s = local mass transfer coefficient for combined surface migration and pore diffusion, kg mol/h•m²•Pa

The local mass transfer coefficient cannot be satisfactorily predicted from basic theory at the present time. However, it (and therefore K_{OG}) can be determined with some certainty from experimental data. Therefore, Equation 5-2 still does not give a simple and accurate means of predicting adsorber performance from physical data. What Equation 5-2 does show is that the equilibrium partial pressure of the adsorbate (p^*) determines the theoretical minimum adsorber bed size. Empirical design procedures based on adsorption equilibrium conditions are the easiest and most common methods used to predict adsorber size and performance. These methods will be discussed later.

Adsorption Forces—Physical and Chemical

The adsorption process is classified as either physical or chemical. The basic difference between physical and chemical adsorption is the manner in which the gas molecule is bonded to the adsorbent. In physical adsorption the gas molecule is bonded to the solid surface by weak forces of intermolecular cohesion. The chemical nature of the adsorbed gas remains unchanged; therefore, physical adsorption is a readily reversible process. In chemical adsorption a much stronger bond is formed between the gas molecule and adsorbent. A sharing or exchange of electrons takes place—as happens in a chemical bond. Chemical adsorption is not easily reversible.

The forces active in physical adsorption are electrostatic in nature. These forces are present in all states of matter: gas, liquid, and solid. They are the same forces of attraction which cause gases to condense and real gases to deviate from ideal behavior. This electrostatic force can be measured by the constant "a" in van der Waals' equation describing nonideal gas behavior. Physical adsorption is sometimes also referred to as van der Waals' adsorption. The electrostatic effect which produces the van der Waals' forces depends on the polarity of both the gas and solid molecules. Molecules in any state are either polar or nonpolar depending on their

chemical structure. **Polar** substances are those which exhibit a separation of positive and negative charges within the compound. This separation of positive and negative charges is referred to as a permanent dipole. Water is a prime example of a polar substance. **Nonpolar** substances have both their positive and negative charges in one center so they have no permanent dipole. Most organic compounds, because of their symmetry, are nonpolar.

Physical, or van der Waals' adsorption can occur from three different effects: an orientation effect, dispersion effects, or induction effects (Figure 5-3). For polar molecules, attraction to each other occurs because of the orientation effect. The orientation effect describes the attraction which occurs between the dipoles of two polar molecules. The negative area of one is attracted to the positive area of the other. An example of this type of adsorption would be the removal of water vapor (polar) from an exhaust stream by using silica gel (polar).

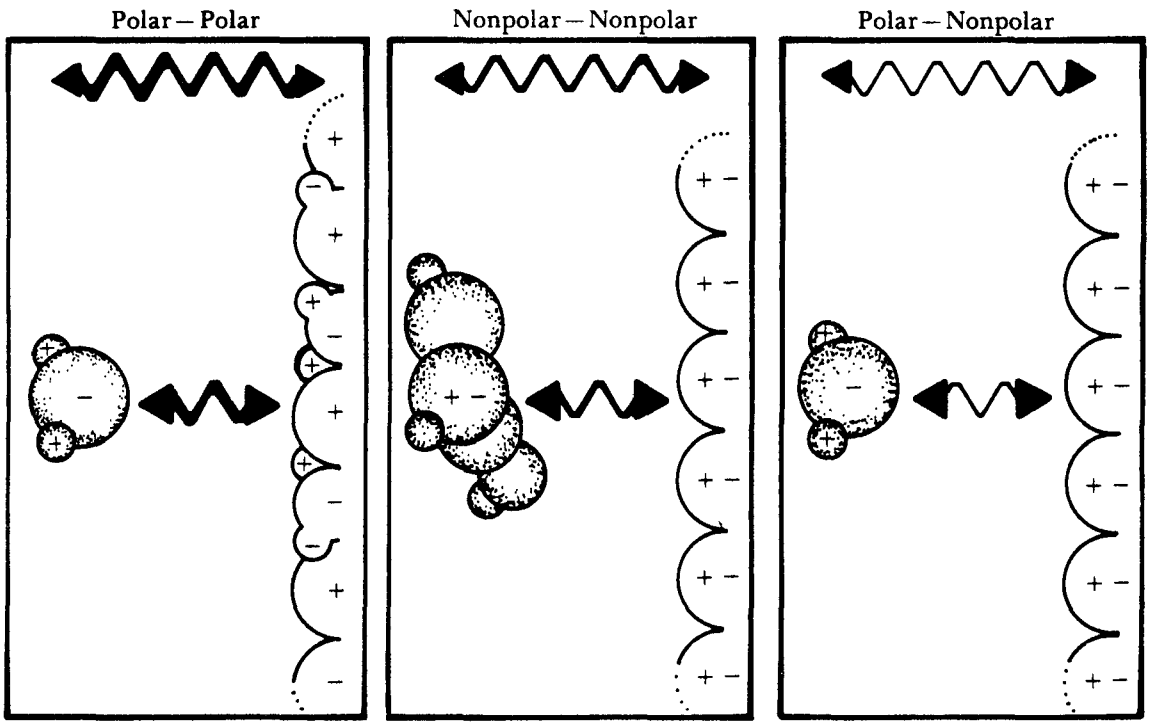


Figure 5-3. Physical forces causing adsorption.

The adsorption of a nonpolar gas molecule onto a nonpolar surface is accounted for by the dispersion effect. The dispersion effect is based on the fact that although nonpolar substances do not possess a permanent dipole, they do have a fluctuating or oscillating dipole. Fluctuating dipoles are a result of momentary changes in electron distribution around the atomic nuclei. In a nonpolar substance, when two fluctuating dipoles come close to one another, their total energy decreases, and they fluctuate in phase with each other. Oscillating dipoles disperse light. Consequently, this is where the name dispersion effect comes from. The adsorption of organic vapors onto activated carbon is an example of nonpolar molecular attraction.

The attraction between a molecule with a permanent dipole (polar molecule) and a nonpolar molecule is caused by the induction effect. A molecule with a permanent dipole can induce or polarize a nonpolar molecule when they come in close contact. The energy of this effect is determined by the polarizability of the nonpolar molecules. The induction effect is, however, very small when compared to the orientation or dispersion effects. Therefore, adsorption systems use polar adsorbents to remove polar contaminants and nonpolar adsorbents to remove nonpolar contaminants. This ensures that the intermolecular forces of attraction between the gas and solid will be greater than those between similar molecules in the gas phase.

Chemical adsorption or chemisorption results from the chemical interaction between the gas and the solid. The gas is held to the surface of the adsorbate by the formation of a chemical bond. Adsorbents used in chemisorption can be either pure substances or chemicals deposited on an inert carrier material. One example is using pure iron oxide chips to adsorb H_2S gases. Another example is using activated carbon which has been impregnated with sulfur to remove mercury vapors.

All adsorption processes are *exothermic*, whether adsorption occurs from chemical or physical forces. In adsorption, molecules are transferred from the gas to the surface of a solid. The fast-moving gas molecules lose their kinetic energy of motion to the adsorbent in the form of heat.

In chemisorption, the heat of adsorption is comparable to the heat evolved from a chemical reaction, usually over 10 kcal/mol. The heat given off by physical adsorption is much lower, approximately 100 cal/mol, which is comparable to the heat of condensation. Some additional general differences between physical adsorption and chemisorption which make physical adsorption more desirable for air pollution control are:

1. Molecules that are adsorbed by chemisorption are very difficult (and in some cases, impossible) to remove from the adsorbent bed. Physically adsorbed molecules can usually be removed by either increasing the operating temperature or reducing the pressure of the adsorbent bed.
2. Chemisorption is a highly selective process. A gas molecule must be capable of forming a chemical bond with the adsorbent surface for chemisorption to occur. Physical adsorption occurs under suitable conditions in most gas-solid systems. For industrial purposes specific solids are chosen which enhance the rate of adsorption.
3. Chemisorption stops when all the reactive sites on the surface of the adsorbent have reacted. Chemisorption forms only a monolayer of adsorbate molecules on the surface. Because of van der Waals' forces, physical adsorption can form multilayers of adsorbate molecules—one atop another.
4. The chemisorption rate increases with increasing temperature. For physical adsorption the exact opposite is true: the rate decreases with increasing temperature.

For these and other reasons, chemisorption is not used extensively in air pollution control systems. A summary of the characteristics of physical versus chemical adsorption is presented in Table 5-1.

Table 5-1. Summary of characteristics of chemisorption and physical adsorption.

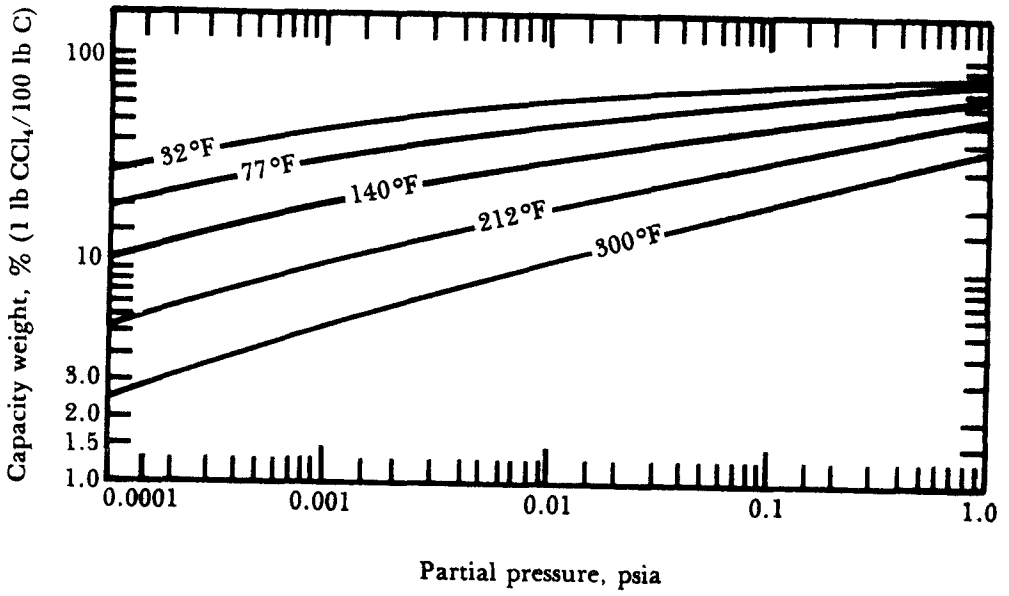
Chemisorption	Physical adsorption
Releases high heat 10,000 cal/mol	Releases low energy 100 cal/mol
Forms a chemical compound	Dipolar interaction
Desorption is difficult	Desorption is easy
Impossible adsorbate recovery	Easy adsorbate recovery

Adsorption Equilibrium Relationships

Most available data on adsorption systems is determined at equilibrium conditions. Adsorption equilibrium is the set of conditions at which the number of molecules arriving on the surface of the adsorbent equals the number of molecules that are leaving. The adsorbent bed is said to be "saturated with vapors" and can remove no more vapors from the exhaust stream. Equilibrium determines the maximum amount of vapor that may be adsorbed at a given set of operating conditions. Although a number of variables affect adsorption, the two most important in determining equilibrium for a given system are temperature and pressure. Three types of equilibrium graphs are used to describe adsorption systems: isotherm at constant temperature; isobar at constant pressure; and isostere at constant amount of vapors adsorbed.

Isotherm

The most common and useful adsorption equilibrium data is the adsorption *isotherm*. The isotherm is a plot of the adsorbent capacity versus the partial pressure of the adsorbate at a constant temperature. Adsorbent capacity is usually given in weight percent expressed as gram of adsorbate per 100 grams of adsorbent. Figure 5-4 is a typical example of an adsorption isotherm for carbon tetrachloride on activated carbon. Graphs of this type are used to estimate the size of adsorption systems as illustrated in Example 5-1, following. Attempts have been made to develop generalized equations which can predict adsorption equilibrium from physical data. This is very difficult because adsorption isotherms take many shapes depending on the forces involved. Isotherms may be concave upwards, concave downwards or "S" shaped. To date, most of the theories agree with data only for specific adsorbates-adsorbent systems and are valid over limited concentration ranges.



Source: Adapted from Technical Bulletin Calgon Corp.

Figure 5-4. Adsorption isotherm for carbon tetrachloride on activated carbon.

Example 5-1*

A dry cleaning process exhausts a 15,000 scfm air stream containing 680 ppm carbon tetrachloride. Given Figure 5-4 and assuming the exhaust stream is at approximately 140°F and 14.7 psia, determine the saturation capacity of the carbon.

Solution:

In the gas phase, the mole fraction (Y) is equal to the percent by volume.

$$Y = \% \text{ volume} = 680 \text{ ppm}$$

$$= \frac{680}{10^6}$$

$$= 0.00068$$

Obtaining the partial pressure:

$$p = YP$$

$$= (0.00068)(14.7 \text{ psia})$$

$$= 0.01 \text{ psia}$$

*This example problem is in English Units since Figure 5-4 needed to solve the problem is in English Units.

From Figure 5-4, at a partial pressure of 0.01 psia and a temperature of 140°F, the carbon capacity is read as 30%.

This means that at **saturation**, 30 lbs of vapor are removed per 100 lbs of carbon in the adsorber (30 kg/100 kg).

It must be noted that in practical applications adsorbers use more carbon than is predicted at saturation to ensure that uncaptured vapors are not being exhausted to the atmosphere. Example 5-2 will illustrate this point (page 5-18).

Polanyi Potential Theory

The most useful theory from an engineering design viewpoint, in trying to predict adsorption isotherms, is the Polanyi potential theory (Cerny, 1970). The Polanyi theory states that the adsorption potential is a function of the reversible isothermal work done by the system. This is expressed in Equation 5-4 relating the free energy of adsorption, ΔF, to the relative vapor pressure $p°/p$ (Polanyi, 1914).

(Eq. 5-4)

$$\Delta F = -RT\ln(p°/p)$$

Where: ΔF = free energy of adsorption, cal/mol
T = absolute temperature, K
$p°$ = vapor pressure of adsorbate, Pa
p = partial pressure of adsorbate, Pa
R = ideal gas constant

Polanyi observed that for any fixed amount of adsorbed material, ΔF is constant. This relationship makes it possible to calculate the relative vapor pressures of a gas in equilibrium with the adsorbent at a variety of temperatures given one **measured** isotherm. If $p°/p$ and T are known at a fixed amount of material adsorbed (therefore ΔF) then from Equation 5-4, $p°/p$ can be estimated for the same amount adsorbed but at different temperatures. This relationship still requires that experimental data be available and is only valid over a narrow temperature range: from 60 to 130°F (EPA, 1972).

Langmuir Theory

Another theory used to describe the isothermal adsorption process is the Langmuir theory. Langmuir assumed that adsorption occurs in a unimolecular layer so that a molecule can be adsorbed only if it strikes an uncovered area of the adsorbent (Langmuir, 1918). The derivation of the Langmuir equation also assumes that the forces acting in adsorption are the same as those for condensation of vapors. The equation has no direct application to pollution control system design, but is the basis for other useful equations. Brunauer, Emmett, and Teller (1938) developed equations for multimolecular adsorption based on Langmuir's ideas. A common form of the Brunauer, Emmett, and Teller (BET) equation is given as:

(Eq. 5-5)
$$\frac{p}{V(p^\circ - p)} = \frac{1}{V_m C} + \left(\frac{C-1}{V_m C}\right)(p/p^\circ)$$

Where: V = volume of vapor adsorbed per unit weight of adsorbent
V_m = volume of vapor adsorbed per unit weight of adsorbent with a layer one molecule thick
C = constant dependent on temperature

Equation 5-5 is used by manufacturers of adsorbents to report the surface area of their products. A graph of $[p/V(p^\circ - p)]$ versus p/p° gives a straight line with a slope of $C - 1/V_m C$ and an intercept of $1/V_m C$. From these relationships both C and V_m can be computed. Just as with the Polanyi equation, the BET equation has limitations to its use. Data from small-pore sized adsorbents deviate from the straight line predicted by the BET equation. To date, there are no universally accepted equations based on theory that are used to size adsorbers.

Isostere and Isobar

Two additional adsorption equilibrium relationships are the isostere and the isobar. The isostere is a plot of the ln p versus 1/T at a constant amount of vapor adsorbed. Adsorption isostere lines are usually straight for most adsorbate-adsorbent systems. Figure 5-5 is an adsorption isostere graph for the adsorption of H_2S gas onto molecular sieves. The isostere is important in that the slope of the isostere corresponds to the heat of adsorption. This can also be seen from Equation 5-4. The last equilibrium relationship is the isobar. The isobar is a plot of the amount of vapors adsorbed versus temperature at a constant pressure. Figure 5-6 shows an isobar line for the adsorption of benzene vapors on activated carbon. Note that the amount adsorbed decreases with increasing temperature, which is always the case for physical adsorption. Since these three relationships were developed at equilibrium conditions, they depend on each other. By determining one, such as the isotherm, the other two relationships can be determined for a given system. In the design of a pollution control system, the adsorption isotherm is by far the most commonly used equilibrium relationship.

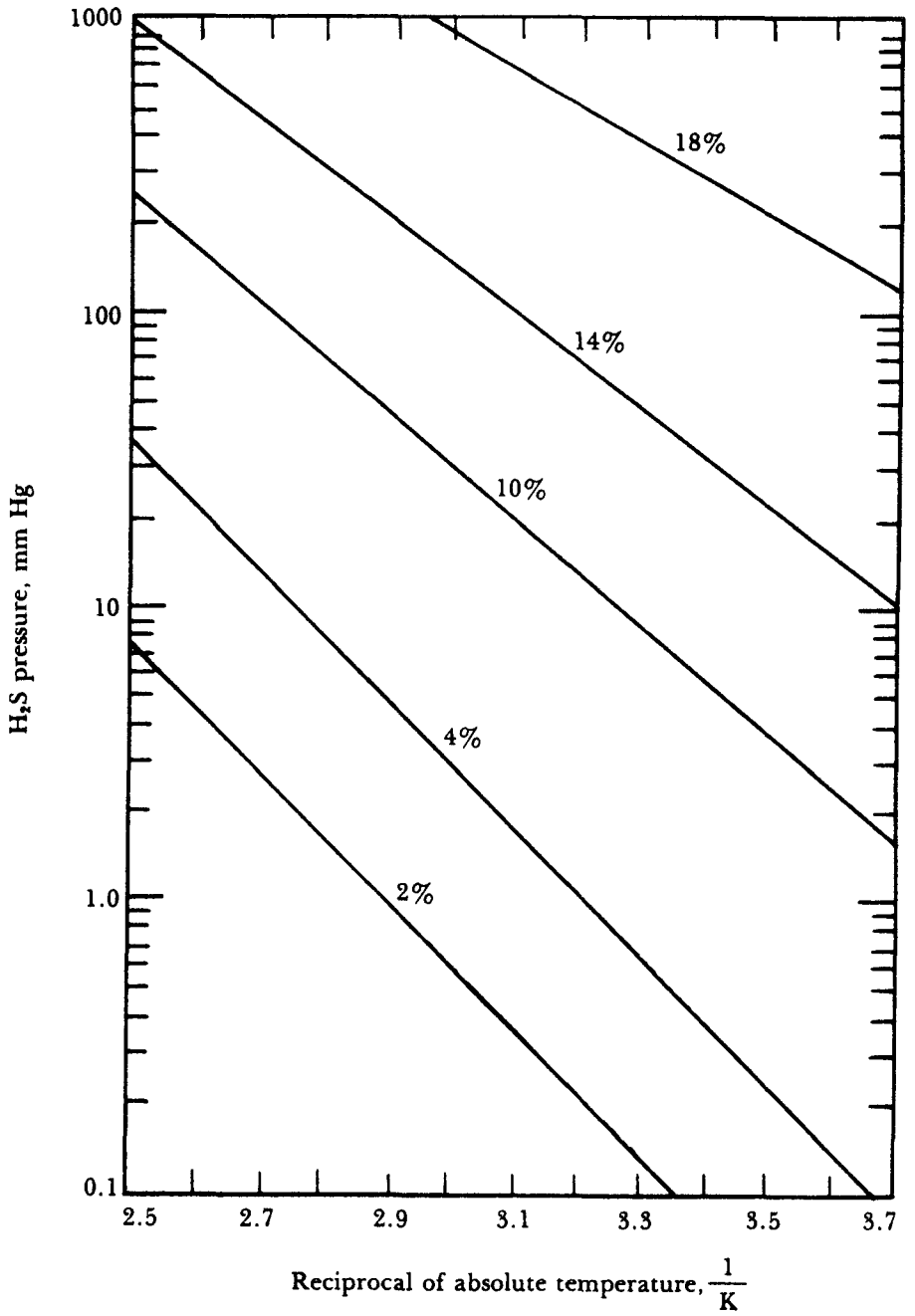


Figure 5-5. Adsorption isosteres of H_2S on 13X molecular sieve loading in % weight.

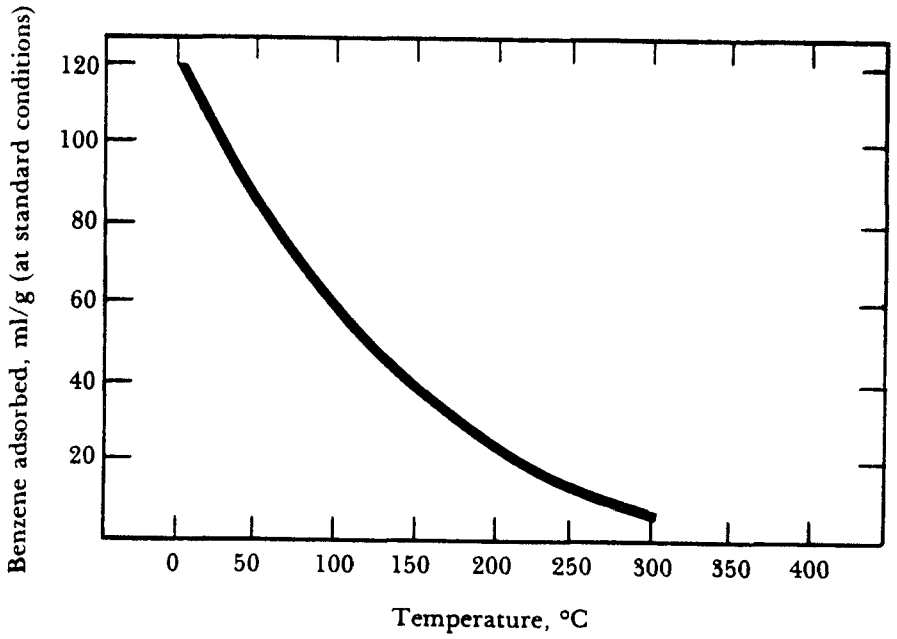


Figure 5-6. Adsorption isobar for benzene on carbon (benzene at 10.0 mm Hg).

Adsorbent Materials

Several materials are used effectively as adsorbing agents. The most common adsorbents used industrially are *activated carbon, silica gel, activated alumina* (alumina oxide), and *zeolites* (molecular sieves). Adsorbents are characterized by their chemical nature, extent of their surface area, pore size distribution, and particle size. In physical adsorption the most important characteristic in distinguishing between adsorbents is their surface polarity. As discussed previously, the surface polarity determines the type of vapors a particular adsorbent will have the greatest affinity for. Of the above adsorbents, activated carbon is the primary nonpolar adsorbent. It is possible to manufacture other adsorbing material having nonpolar surfaces. Since their surface area is much less than that of activated carbon, they are not used commercially. Polar adsorbents will preferentially adsorb any water vapor that may be present in a gas stream. Since moisture is present in most pollutant air streams, the use of polar adsorbents is severely limited for an air pollution system. Therefore, in further discussion, the emphasis is placed on the use of activated carbon, although some of the information is applicable to polar adsorption systems.

Activated Carbon

Activated carbon can be produced from a variety of feedstocks such as wood, coal, coconut, nutshells, and petroleum-based products. The activation process takes place in two steps. First, the feedstock is carbonized. Carbonization involves heating

(usually in the absence of air) the material to a temperature high enough (600°C) to drive off all volatile material. Thus, carbon is essentially all that is left. To increase the surface area, the carbon is then "activated" by using steam, air, or carbon dioxide at higher temperatures. These gases attack the carbon and increase the pore structure. The temperatures involved, the amount of oxygen present, and the type of feedstock, all greatly affect the adsorption qualities of the carbon. Manufacturers vary these parameters to produce activated carbons suitable for specific purposes. In sales literature, the activity and retentivity of carbons are based on their ability to adsorb a standard solvent, such as carbon tetrachloride (CCl_4).

Because of its nonpolar surface, activated carbon is used to control emission of organic solvents, odors, toxic gases, and gasoline vapors. Carbons used in gas phase adsorption systems are manufactured in granular form, usually between 4×6 to 4×20 mesh in size. Bulk density of the packed bed can range from 0.08 to 0.5 g/cm^3 (5 to 30 lb/ft^3) depending on the internal porosity of the carbon. Surface area of the carbon can range from 600 to 1600 m^2/g (2.9×10^6 to 7.8×10^6 ft^2/lb). This is equivalent to having the surface area of 2 to 5 football fields in one gram of carbon.

Silica Gel

Silica gels are made from sodium silicate. Sodium silicate is mixed with sulfuric acid, resulting in a jellylike precipitant from where the "gel" name comes. The precipitant is then dried and roasted. Depending on the processes used in manufacturing the gel, different grades varying in activity can be produced. Silica gels have surface areas of approximately 750 m^2/g (3.7×10^6 ft^2/lb). Silica gels are used primarily to remove moisture from exhaust streams, but are ineffective at temperatures above 260°C (500°F).

Molecular Sieves

Unlike the other adsorbents, which are amorphous (not crystalline) in nature, molecular sieves have a crystalline structure. The pores are, therefore, uniform in diameter. Molecular sieves can be used to capture or separate gases on the basis of molecular size and shape. An example of this are refining processes which sometimes use molecular sieves to separate straight chained paraffins from branched and cyclic compounds. However, the main use of molecular sieves is in the removal of moisture from exhaust streams. The surface area of molecular sieves range from 600 to 700 m^2/g (2.9×10^6 to 3.4×10^6 ft^2/lb).

Aluminum Oxide (Activated Alumina)

Aluminum oxides are manufactured by thermally activated alumina or bauxite. This is accomplished by heating the alumina in an inert atmosphere to produce a porous aluminum oxide pellet. Aluminum oxides are not commonly used in air pollution applications. They are primarily used for the drying of gases, especially

under high pressures, and as support material in catalytic reactions. A prime example is the impregnating of the alumina with platinum or palladium for use in catalytic incineration. Activated alumina's surface areas can range from 200 to 300 m^2/g (0.98×10^6 to 1.5×10^6 ft^2/lb).

Pore Size Distribution

The physical properties of the adsorbent affect the adsorption capacity, rate, and pressure drop across the adsorber bed. Table 5-2 summarizes these properties for the above adsorbents. Since adsorption occurs at the gas-solid interface, the surface area available to the vapor molecules determines the effectiveness of the adsorbent. Generally, the larger the surface area, the higher the adsorbent's capacity. However, the surface area must be available in certain pore sizes if it is to be effective as a vapor adsorber. Dubinin (1936) classified the pores in activated carbon as micropores, transitional pores, or macropores. Micropores are openings whose radii are 200 nanometers (20 Å) or less. Pores larger than 2000 nanometers (200 Å) are macropores. Transitional pores are those with radii between 200 and 2000 nanometers.

Table 5-2. Physical properties of major types of adsorbents.

Adsorbent	Internal porosity (%)	Surface area (m^2/g)	Pore volume (cm^3/g)	Bulk dry density (g/cm^3)	Mean pore diameter (nm)
Activated carbon	55-75	600-1600	0.80-1.20	0.35-0.50	150-200
Activated alumina	30-40	200-300	0.29-0.37	0.90-1.00	180-200*
Molecular sieves	40-55	600-700	0.27-0.38	0.80	30-90
Silica gel	70	750	0.40	0.70	220

*The 150-200 nanometer average is for the micropores only; since 95% of the surface area is associated with them.

Most gaseous air pollutant molecules are in the 40 to 90 nanometer size range. If a large portion of an adsorbent's surface area is in pores smaller than 40 nanometers, many contaminant molecules will be unable to reach these active sites. Figure 5-7 illustrates molecule movement in the pores. In addition, the larger pore sizes (macro and transitional) contribute little to molecule capture. The vapor pressure of the contaminant in these larger areas is too low to be effectively removed. These larger pores serve mainly as passageways to the smaller micropore area where the adsorption forces are strongest. Adsorption forces are strongest in pores that are not more than approximately twice the size of the contaminant molecule. These strong adsorption forces result from the overlapping attraction of the closely spaced walls.

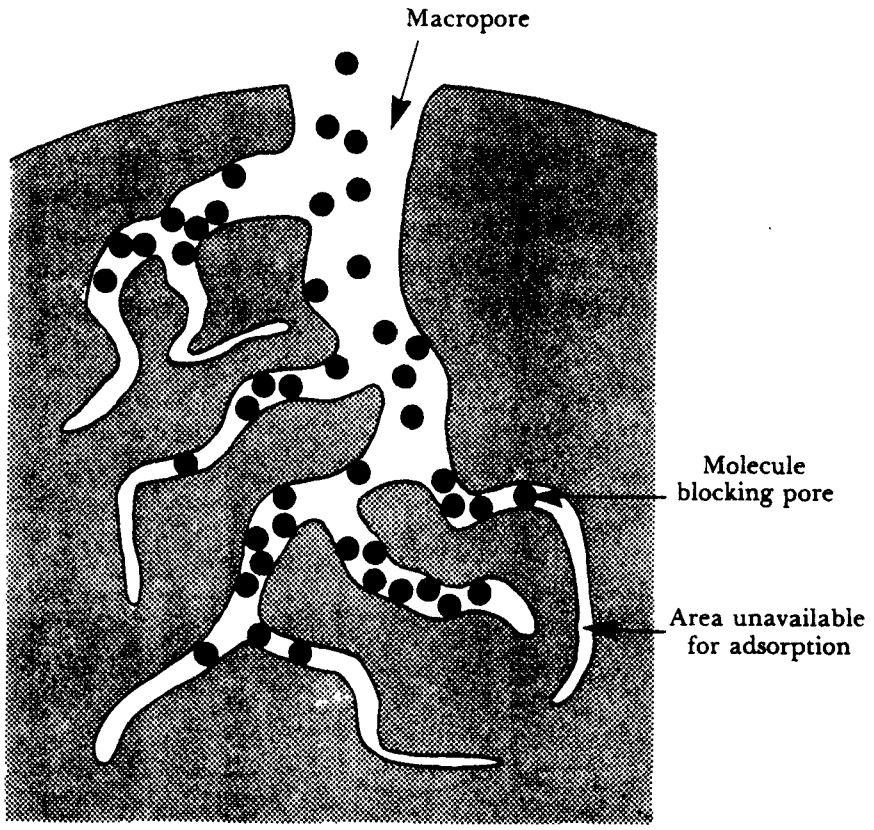


Figure 5-7. Molecular screening in pores of activated carbon.

Another phenomenon, *capillary condensation* usually only occurs in the micropores. Capillary condensation occurs when multilayers of adsorbed contaminant molecules build up from both sides of the pore wall, totally packing the pore and condensing in the pore. The amount of contaminant that is removed increases since additional molecules condense on the surface of the liquid which has formed. Contaminant molecules can also be removed at lower vapor pressures (more dilute concentrations) since capture forces are now acting from three sides instead of just one. However, desorption is not as complete if capillary condensation occurs, since the forces that hold a liquid together are much stronger than the physical adsorption forces.

Air pollution control involves removing contaminant vapors at low partial pressures. Therefore, the micropore structure of an adsorbent plays an important role in determining the overall efficiency. Another reason for the wide use of activated carbon is that 90 to 95% of its surface area is in the micropore size range.

Dynamic Adsorption Process

The movement of vapors through an adsorbent bed is often referred to as a dynamic process. The term dynamic refers to motion both in the movement of air through the adsorbent bed and change in vapor concentration as it moves through the bed. There are a variety of configurations in which the contaminant air stream and adsorbent are brought into contact. The most common configuration is to pass the air stream down through a fixed volume or bed of adsorbent. Figure 5-8 illustrates how adsorption (mass transfer) occurs as vapors pass down through the bed.

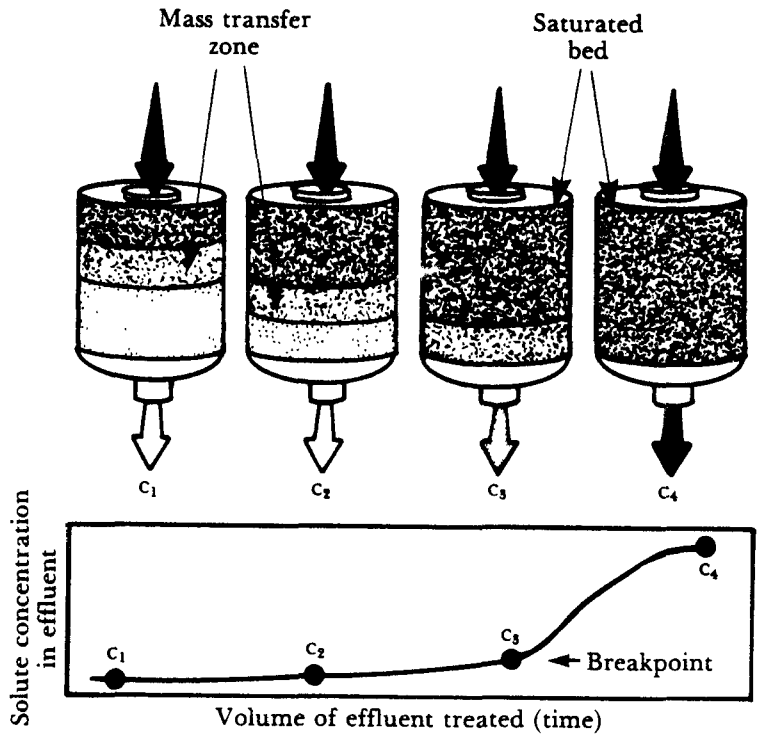


Figure 5-8. Breakthrough curve.

The gas stream containing the pollutant, at an initial concentration, c_i, is passed down through a deep bed of adsorbent material which is free of any contaminant. Most of the contaminant is readily adsorbed by the top portion of the bed. The small amount of contaminant that is left is easily adsorbed in the remaining section of the bed. The effluent from the bottom of the bed is essentially pollutant free, denoted at c_1.

After a period of time the top layer of the adsorbent bed becomes saturated with contaminant. The majority of adsorption (approximately 95%) now occurs in a narrow portion of the bed directly below this saturated section. The narrow zone of adsorption is referred to as the *mass transfer zone* (MTZ). As additional contaminant vapors pass through the bed, the saturated section of the bed becomes larger and the MTZ moves further down the length of the adsorber. The actual length of

the MTZ remains fairly constant as it travels through the adsorbent bed. Additional adsorption occurs as the vapors pass through the "unused" portion of the bed. The effluent concentration at c_2 is essentially still zero since there is still an unsaturated section of the bed.

Finally, when the lower portion of the MTZ reaches the bottom of the bed, the concentration of contaminant in the effluent suddenly begins to rise. This is referred to as the *breakthrough point*—where untreated vapors are being exhausted from the adsorber. If the contaminated air stream is not switched to a fresh bed, the concentration of contaminant in the outlet will quickly rise until it equals the initial concentration, illustrated at point c_4.

In most air pollution control systems even trace amounts of contaminants in the effluent stream are undesirable. To achieve continuous operation, adsorbers must be either replaced or cycled from adsorption to desorption before breakthrough occurs. In *desorption* or *regeneration*, the contaminant vapors are removed from the used bed in preparation for the next cycle. Most commercial adsorption systems are the regenerable type.

In regard to regenerable adsorption systems, three important terms are used to describe the *capacity* of the adsorbent bed. All the capacities are measured in lb of vapor per lb of adsorbent. First, the *breakthrough capacity* is defined as the capacity of the bed at which unreacted vapors begin to be exhausted. The *saturation capacity* is the maximum amount of vapors that can be adsorbed per unit weight of carbon. (This is the capacity read from the adsorption isotherm). The *working capacity* is the actual amount of adsorbent used in an adsorber. The working capacity is a certain fraction of the saturation capacity. Working capacities can range from 0.1 to 0.5 of the saturation capacity. (Note: a smaller capacity **increases** the amount of carbon required.) This fraction is set by the designer for individual systems by balancing the cost of carbon and adsorber operation versus preventing breakthrough allowing for an adequate cycle time.

Another factor in determining the working capacity is that it is uneconomical to desorb all the vapors from the adsorber bed. The small amount of residual vapors left on the bed is referred to as the *heel*. This heel accounts for a large portion of the difference between the saturation and the working capacity. In some cases the working capacity can be estimated by assuming it is equal to the saturation capacity minus the heel (Turk, 1977). The following example illustrates one method of estimating the working capacity. In all the examples in this manual and the accompanying workbook, a design factor of 0.5 of the saturation capacity is used. This is the same as assuming the working amount of carbon is twice the amount required at saturation.

Example 5-2

Assume the same conditions as stated in Example 5-1. Estimate the amount of carbon that would be required if the adsorber were to operate on a 4-hour cycle. The molecular weight of CCl_4 is 154 lb/lb mole.

Solution:

From Example 5-1 we know that the carbon used will remove 30 lbs of vapor for every 100 lbs of carbon at saturation conditions.

First compute the flow rate of CCl_4:

$$Q_{CCl_4} = 15{,}000 \text{ scfm} \times 0.00068$$
$$= 10.2 \text{ scfm } CCl_4$$

Converting to pounds per hour:

$$10.2 \frac{\text{ft}^3}{\text{min}} \times \frac{\text{lb mole}}{359 \text{ ft}^3} \times \frac{154 \text{ lb}}{\text{lb mole}} \times \frac{60 \text{ min}}{\text{hr}} = 262.5 \frac{\text{lb } CCl_4}{\text{hr}}$$

For a 4-hour cycle there are: $4 \times 262.5 = 1050$ lb CCl_4.

The amount of carbon (at saturation) required:

$$1050 \text{ lb } CCl_4 \times \frac{100 \text{ lb carbon}}{30 \text{ lb } CCl_4} = 3500 \text{ lb carbon}$$

The actual amount of carbon required can be estimated by doubling the amount needed at saturation.

$2 \times 3500 = 7000$ lb (3182 kg) carbon per 4-hour cycle per adsorber.

Note: this gives only a rough estimate of the amount of carbon needed.

Factors Affecting Adsorption

A number of factors or system variables influence the performance of an adsorption system. These variables and their effects on the adsorption process are discussed in the following section.

Temperature

For physical adsorption processes, the capacity of an adsorbent decreases as the temperature of the system increases. Figure 5-9 illustrates this concept. As the temperature increases the vapor pressure of the adsorbate increases, raising the energy level of the adsorbed molecules. Adsorbed molecules now have sufficient energy to overcome the van der Waals' attraction and migrate back to the gas phase. Molecules already in the gas phase tend to stay there due to their high

vapor pressure. As a general rule, adsorber temperatures are kept below 55 °C (130 °F) to ensure adequate bed capacities. Temperatures above this limit can be avoided by cooling the exhaust stream that is to be treated.

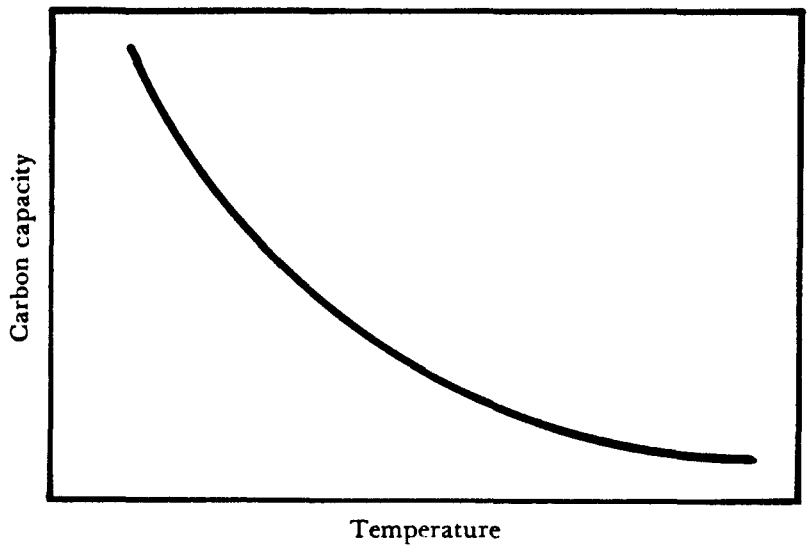


Figure 5-9. Carbon capacity vs. temperature.

Adsorption is an exothermic process with the heat released for physical adsorption approximately equal to the heat of condensation. At low concentrations (below 100 ppm) the heat release is minimal and is quickly dissipated by the air flow through the bed. At higher concentrations (approximately 5000 ppm) considerable heating of the bed can occur, which if not removed can cause the adsorber efficiency to rapidly decrease. In addition, granular carbon is a good insulator, which inhibits heat dissipation from the interior of the bed. In some cases, especially ketone recovery, the temperature rise can cause auto-ignition of the carbon bed. Monitoring of bed temperatures and leaving the bed slightly wet after steam regeneration are techniques used to avoid bed fires.

Pressure

Adsorption capacity increases with an increase in the partial pressure of the vapor. The partial pressure of a vapor is proportional to the total pressure of the system. Any increase in pressure will increase the adsorption capacity of a system (see Figure 5-4). The increase in capacity occurs because of a decrease in the mean free path of vapors at higher pressures. Simply, the molecules are packed more tightly together. More molecules have a chance to "hit" the available adsorption sites, increasing the number of molecules adsorbed.

Gas Velocity

The contact or residence time between the contaminant stream and adsorbent is determined by the gas velocity through the adsorber. The residence time directly affects capture efficiency. The slower the contaminant stream flows through the adsorbent bed, the greater the probability of a contaminant molecule hitting an available site. Once a molecule has been captured it will stay on the surface until the physical conditions of the system are changed. To achieve 90% + capture efficiency most carbon adsorption systems are designed for a maximum air flow velocity of 30 m/min (100 ft/min) through the adsorber. A lower limit of at least 6 m/min (20 ft/min) is maintained to avoid flow distribution problems, such as channeling.

Gas velocity through the adsorber is a function of the diameter of the adsorber for a given volume of contaminant gas. By specifying a maximum velocity through the adsorber, the minimum diameter is also specified. For example, if 300 m^3/min of contaminant gas is to be treated, and the maximum velocity through the adsorber is to be 30 m/min, then the adsorber must have a cross sectional area of at least 10 m^2.

The gas flow rate through the adsorber also affects the pressure drop. Increasing the flow rate increases the pressure drop. Within the above stated maximum and minimum flow rates, the allowable pressure drop usually dictates the required tower diameter and flow rate. The pressure drop across the bed also depends on the depth of adsorbent. This will be discussed in the following section.

Bed Depth

Providing a sufficient depth of adsorbent is very important in achieving efficient gas removal. If the adsorber bed depth is shorter than the required mass transfer zone, breakthrough will immediately occur rendering the system ineffective. Computing the length of the MTZ is very difficult since it depends upon six factors: the adsorbent particle size, gas velocity, adsorbate concentration, fluid properties of the gas stream, temperature, and pressure of the system. The MTZ can be estimated from experimental data using Equation 5-6 (Kovach, 1978). To obtain the necessary data, vendors will usually test a small portion of the exhaust stream on a pilot adsorber column.

(Eq. 5-6)

$$MTZ = \frac{1}{1 - X_S} D \left(1 - \frac{C_B}{C_S}\right)$$

Where: D = bed depth, m
C_B = breakthrough capacity, %
C_S = saturation capacity, %
X_S = degree of saturation in the MTZ, % (usually assumed to be 50%)
MTZ = length of MTZ, m

The above equation is used mainly as a check to ensure that the proposed bed depth is longer than the MTZ. Actual bed depths are usually many times longer than the length of the MTZ. The additional bed depth allows for adequate cycle times. Equation 5-6 can be rearranged to solve for the breakthrough capacity:

(Eq. 5-7) $$C_B = \frac{(X_S)(C_S)(MTZ) + C_S(D - MTZ)}{D}$$

The total amount of adsorbent required is usually determined from the adsorption isotherm, as illustrated in Example 5-2. Once this has been set, the bed depth can then be estimated by knowing the tower diameter and density of the adsorbent. Example 5-3 illustrates how this is done. Generally, the adsorbent bed is sized to the maximum length allowed by the pressure drop across the bed. Data on the pressure drop per meter of bed depth for typical carbons is presented in Figure 5-10 (Turk, 1977). The pressure drop per meter of bed depth is plotted versus the gas flow rate, with the carbon mesh size as a parameter. From the figure, an adsorber with a flow rate of 40 cm/s (80 ft/min) using 4 × 10 mesh carbon will have a pressure drop of approximately 5 kPa per meter (6 in. H_2O per foot) of bed depth. Therefore, if the pressure drop across the bed is limited to 4.5 kPa (18 in. H_2O) then the total bed depth should not exceed 0.9 meters (3 ft).

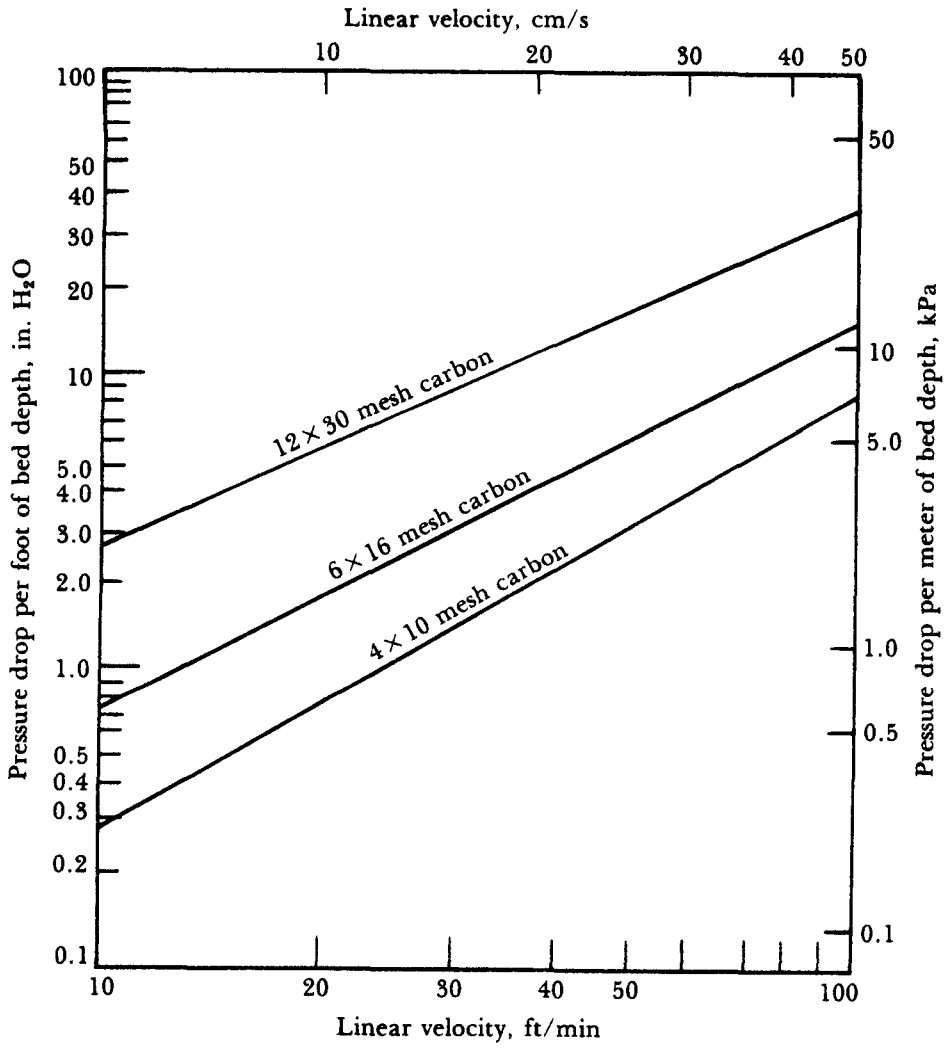


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Figure 5-10. Pressure drop vs. flow rate through granular carbon beds.

Humidity

As stated previously activated carbon will preferentially adsorb nonpolar hydrocarbons over polar water vapor. The water vapor molecules in the exhaust stream exhibit strong attractions for each other rather than the adsorbent. At high relative humidities, over 50%, the number of water molecules increases such that they begin to compete with the hydrocarbon molecules for active adsorption sites. This reduces the capacity and the efficiency of the adsorption system. Exhaust streams with humidities greater than 50% may require installation of additional equipment to remove some of the moisture. Coolers to remove the water are one solution. Dilution air with significantly less moisture in it than the process stream has also been used. Also, the contaminant stream may be heated to reduce the humidity as long as the increase in temperature does not greatly affect adsorption efficiency.

Contaminants

In addition to humidity; particulate matter, entrained liquid droplets, and organic compounds which have high boiling points can also reduce adsorber efficiency if present in the air stream. Any micron-sized particle of dust or lint which is not filtered can cover the surface of the adsorbent. This greatly reduces the surface area of the adsorbent available to the gas molecule for adsorption. Covering of active adsorption sites by an inert material is referred to as *blinding* or *deactivation*. To avoid this situation almost all industrial adsorption systems are equipped with some type of particulate matter removal device.

Entrained liquid droplets can also cause operational problems. Liquid droplets that are nonadsorbing act the same as particulate matter. The liquid covers the surface, blinding the bed. If the liquid is the same as the adsorbate, high heats of adsorption occur. This is especially a problem in activated carbon systems where liquid organic droplets carried over from the process can cause bed fires from the heat released. Some type of entrainment separator may be required when liquid droplets are present.

For activated carbon systems, other contaminants are high boiling point organic compounds, usually in excess of 260°C (500°F). High boiling point (high molecular weight) compounds have such an affinity for the carbon that it is extremely difficult to remove them by standard desorption practices. These compounds also tend to react chemically on the carbon surface forming solids or polymerization products which are extremely difficult to desorb. Loss of carbon activity in this manner is called *chemical deactivation*.

Adsorbent Regeneration Methods

Periodic replacement or regeneration of the adsorbent bed is mandatory in order to maintain continuous operation. When the adsorbate concentration is high and/or cycle time is short (less than 12 hours) replacement of the adsorbent is not feasible

and in-situ regeneration is required. Regeneration is accomplished by reversing the adsorption process, usually by increasing the temperature or decreasing the pressure. Commercially, four methods are used in regeneration:

Thermal swing: The bed is heated so that the adsorption capacity is reduced to a lower level. The adsorbate leaves the surface of the carbon and is removed from the vessel by a stream of purge gas. Cooling must be provided before the subsequent adsorption cycle begins. Steam regeneration is an example of thermal swing regeneration.

Pressure swing: The pressure is lowered at a constant temperature to reduce the adsorbent capacity.

Inert purge gas stripping: The stripping action is caused by an inert gas that reduces the partial pressure of the contaminant in the gas phase, reversing the concentration gradient. Molecules migrate from the surface into the gas stream.

Displacement cycle: The adsorbates are displaced by some preferentially adsorbed material. This method is usually a last resort for situations in which the adsorbate is both valuable and is heat sensitive, and for which pressure swing regeneration is ineffective (Bethea, 1978).

Table 5-3 compares the effectiveness of the various regeneration methods (Wood, 1964). As can be seen from this table, steam regeneration was most effective for the test conditions. This is also true for most industrial applications.

Table 5-3. Regeneration of one pound of activated carbon loaded with 20% ether.

Regeneration method	Regeneration conditions	Expelled ether (%)
Thermal swing	Heating at 100°C (212°F) for 20 min	15
Pressure swing	Vacuum of 50 mm Hg at 20°C (68°F) for 20 min	25
Combination	Gas circulation at 130°C (266°F) for 20 min	45
Thermal swing	Direct steam at 100°C (212°F) for 20 min	98

Thermal Swing—Steam Stripping

Because it is simple and relatively inexpensive, steam stripping is the most common desorption technique. Several additional advantages to using steam for desorption are:

- At high pressure, the steam's temperature (100°C) is high enough to desorb most solvents of interest without damaging the carbon or the desorbed vapors. Desorbed vapors can be polymerized or cracked, sometimes forming undesirable compounds.
- Steam readily condenses in the adsorber bed releasing its (the steam) heat of condensation, aiding in desorption.
- Many organic compounds can be easily separated and recovered from the effluent stream by condensation or distillation.

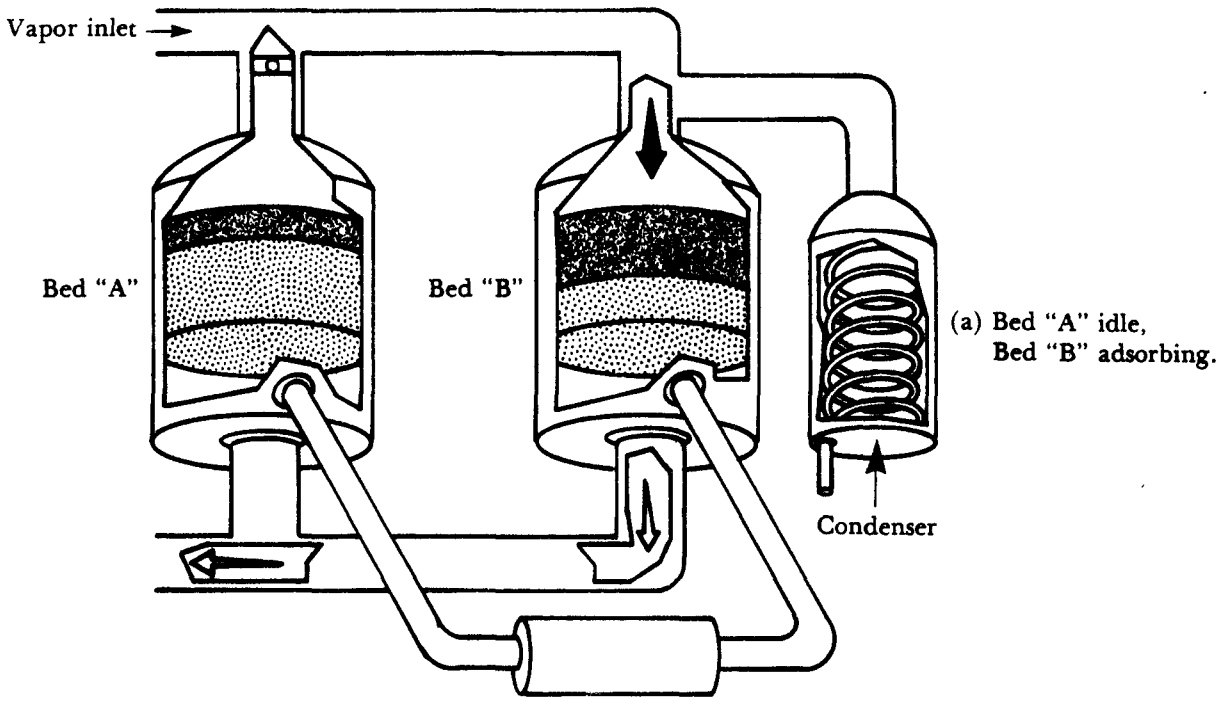
- Residual moisture in the bed is removed easily by a stream of cool dry air (either pure or process effluent).
- Steam is a more concentrated source of heat than hot air so it is very effective in raising the temperature of the adsorber bed very quickly.

The amount of steam required for regeneration depends on the adsorbate loading of the bed. The longer a carbon bed is steamed, the more adsorbate will be desorbed. It is usually not cost effective to try to desorb all of the adsorbed vapors from the bed. Acceptable working capacities can be achieved by using less steam and leaving a small portion of adsorbate in the bed. During the initial heating period no vapors are desorbed. This is because a fixed amount of steam is first required to raise the temperature of the cold bed to the desorption temperature. After this initial period a substantial amount of adsorbate vapor is released, until a plateau is reached. The plateau represents the optimum steam requirement, usually in the range 0.25 to 0.35 kg of steam/kg of carbon (Parmele, 1979). In these systems, steam is usually supplied at pressures ranging from 21 to 103 kPa (3 to 15 psig).

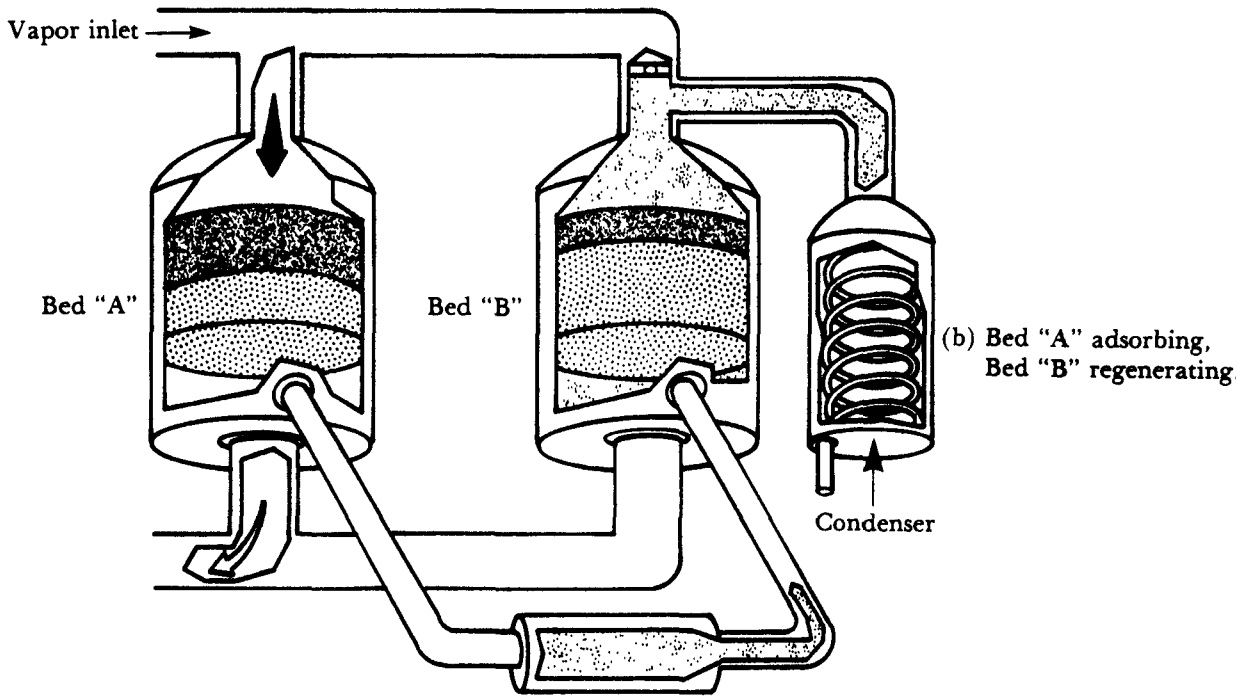
A typical two bed adsorption system is shown in Figure 5-11. Regeneration steam usually passes up through the bed countercurrent to the flow of solvent laden vapors. Since the bed is switched to the desorption mode before breakthrough, the outlet end of the bed remains adsorbate free, providing a safety margin for subsequent cycles. The steam usage can range anywhere from 0.3 to 10 kg of steam per kg of solvent removed.

Some disadvantages are associated with steam regeneration. Problems arising are:

- The effluent from the condenser could pose a water pollution problem unless the condensate is sent to a waste water treatment facility.
- Some organic compounds are subject to hydrolysis and/or other reactions with water which may produce corrosive substances. Corrosive substances can greatly reduce the life of the adsorption equipment unless expensive corrosive resistant materials are used.
- A hot wet carbon bed will not effectively remove organic vapors. Cooling and drying of the bed may be needed to ensure adequate removal efficiencies at the beginning of a subsequent cycle.



(a) Bed "A" idle, Bed "B" adsorbing.



(b) Bed "A" adsorbing, Bed "B" regenerating.

Figure 5-11. Two bed adsorption system.

Pressure Swing—Vacuum Desorption

Pressure swing or vacuum desorption has one primary advantage over thermal desorption. Desorption is accomplished by a change in pressure rather than temperature so no time is required to initially heat up or cool the carbon bed. This adiabatic (no change in temperature) pressure swing allows the bed to be in the adsorbing cycle longer. Units may also be sized smaller since there is no increase in air volume due to heating of the bed. Both of these conditions allow for higher throughputs or smaller adsorber designs than can be accommodated by thermal swing desorption systems. In addition, the desorbed vapors may be recovered directly without the need for additional downstream processing equipment.

The principle disadvantages of a pure pressure swing cycle are the high operating and construction costs. In pressure/vacuum systems the adsorber vessel and valving must be constructed of materials capable of withstanding vacuums of 9.5 kPa (28 in. Hg). Unless the adsorber is initially operated at elevated pressures (so that the pressure swing can be accomplished by reducing the vessel to atmospheric pressures) a vacuum producing system is required. Vacuum systems that operate cyclically may require more operating attention than other regeneration systems. To be effective, pressure regeneration systems must be designed so that a small decrease in pressure will result in a drastic shift in the direction of mass transfer.

Adsorption Control Systems

Adsorption control systems can be classified as either regenerable or nonregenerable. Nonregenerable systems are normally used to control exhaust streams with low pollutant concentrations, below 1.0 ppm. Generally these pollutants are highly odorous or to some degree toxic. When these systems reach the breakthrough point the bed is taken off stream and replaced with a fresh bed. The used carbon can then be sent back to the manufacturer for reactivation. Regenerable systems are used for higher pollutant concentrations such as in solvent recovery operations. Once the bed reaches the breakthrough point in a regenerable system, the pollutant vapors are directed to a second bed while the first has the vapors desorbed.

Nonregenerable Adsorption Systems

Nonregenerable adsorption systems are manufactured in a variety of configurations. Bed areas are sized to control the air flow through them at between 6.0 to 18 m/min (20 to 60 ft/min). They usually consist of thin adsorbent bed depth, ranging in thickness from 1.25 to 10.0 cm (0.5 to 4.0 inches). These thin beds have a low pressure drop, normally below 62 Pa (0.25 in. H_2O) dependent on the bed thickness, gas velocity, and particle size of the adsorbent. Service time for these units can range from 6 months for "heavy" odor concentrations to up to 2 years for

trace concentrations or intermittent operations (EPA, 1973). They are used mainly as air purification devices for small air flows in offices, laboratory exhaust, and other small exhaust streams.

The shapes of these thin bed adsorbers are flat, cylindrical, or pleated. The granules of activated carbon are retained by porous support material, usually perforated sheet metal. An adsorber system usually consists of a number of *retainers* or *panels* placed in one frame. Figure 5-12 shows a nine panel thin bed adsorber. The panels are similar to home air filters except that instead of containing steel wool they contain activated carbon as the filter. Figure 5-13 illustrates a pleated cell adsorber. The pleated cell is one continuous retainer of activated carbon, rather than individual panels. The cylindrical canisters (Figure 5-14) are usually small units designed to handle low flow rates of approximately 0.12 m/s (25 cfm). Cylindrical canisters are made of the same materials as the panel and pleated adsorbers except their shape is round rather than square. Panel and pleated beds are dimensionally about the same size, normally 0.6 meters square (2 ft by 2 ft) with the carbon depth ranging from 0.2 to 0.6 m (8 in. to 2 ft). Flat panel beds are sized to handle higher exhaust flow rates, approximately 9.4 m/s (2000 cfm), while pleated beds are limited to flow rates of 4.7 m/s (1000 cfm). Typical flow rate values are listed in Table 5-4.

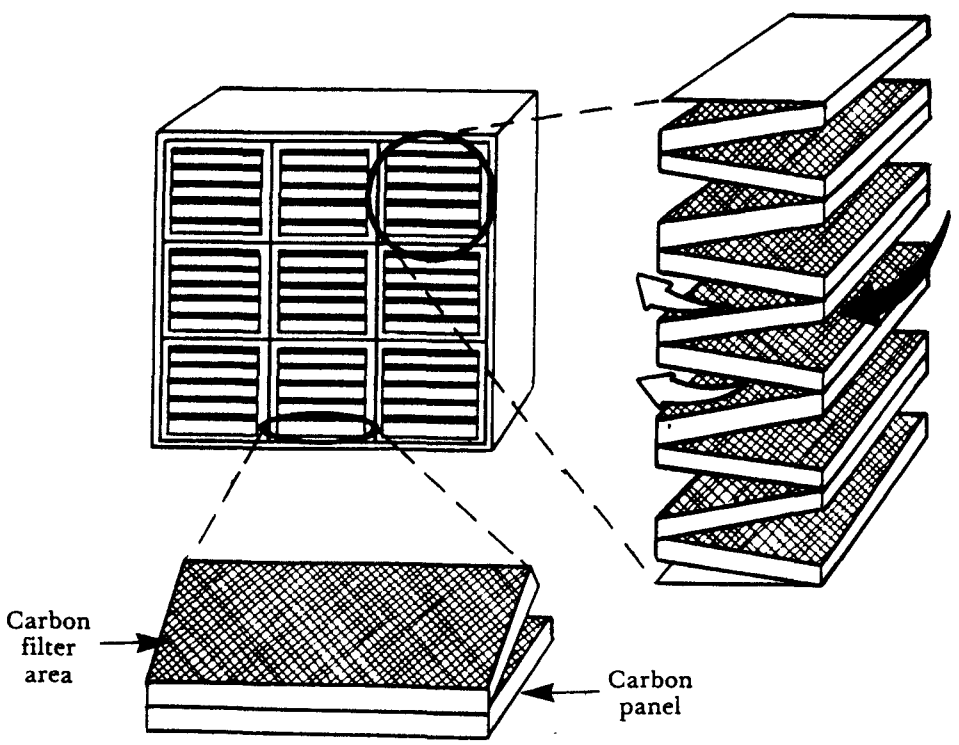


Figure 5-12. Thin bed adsorber: nine cell system.

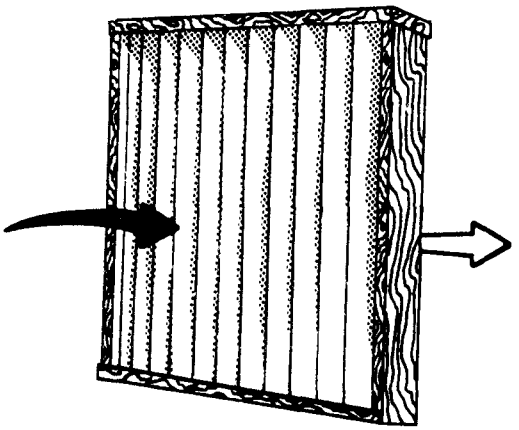


Figure 5-13. Pleated thin bed.

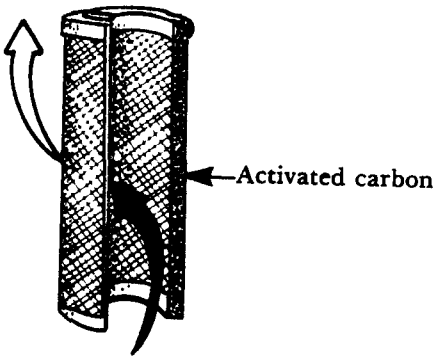


Figure 5-14. Canister.

Table 5-4. Adsorption filters.

Filter shape	Size	Flow rate
Multiple panel cell	~0.6 m^2	9.4 m/s
Pleated cell	~0.6 m^2	4.7 m/s
Cylindrical canister	~0.002 m diameter ~0.005 m length	0.12 m/s

In addition to thin bed systems, thick bed nonregenerable systems are also available. One system that can be used is essentially just a 55-gallon drum. The bottom of the drum is filled with gravel to support a bed of activated carbon weighing approximately 330 kg (150 lb). A typical unit is shown in Figure 5-15. These units are used to treat small flow rates (0.5 m/s or 100 cfm) from laboratory hoods, chemical storage tank vents, and chemical reactors.

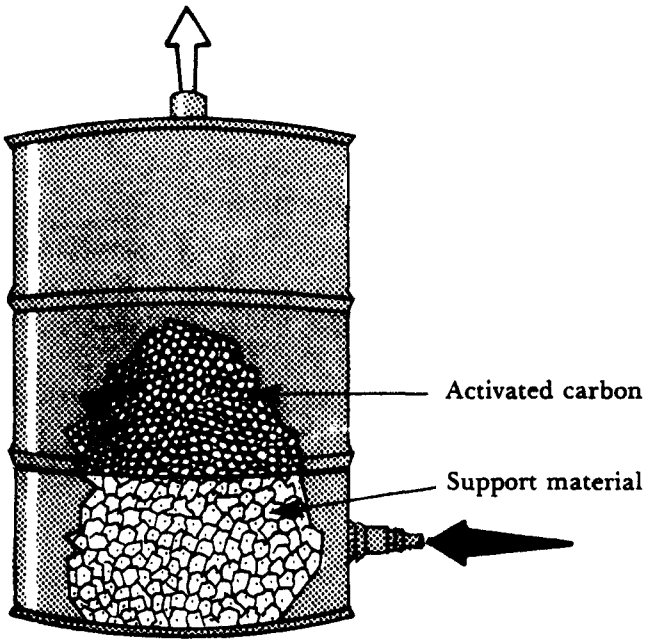


Figure 5-15. Canister.

Regenerable Adsorption Systems

A large regenerable adsorption system can be categorized as a fixed, moving or fluidized bed. The name refers to the manner in which the vapor stream and adsorbent are brought into contact. The choice of a particular system depends on the pollutants to be controlled and the recovery requirements. The most common adsorption system for controlling air pollutants is the fixed carbon bed. These systems are used to control a variety of organic vapors and are usually regenerated by direct steaming of the bed. The organic compounds may be recovered by condensing the exhaust from the regeneration step and separating out the water and solvent.

Fixed Bed Adsorbers

Fixed bed adsorption systems generally involve multiple beds. One or more beds treat the process exhaust while the other beds are either being regenerated or cooled. A typical three bed adsorption system is shown in Figure 5-16. The solvent-laden air stream is first pretreated to remove any solid or liquid particles which could blind the carbon bed and decrease its efficiency. The solvent-laden air stream then usually passes down through the fixed carbon bed. Upward flow through the bed is usually avoided (unless flow rates are low (<500 cfm) to eliminate the risk of entraining carbon particles in the exhaust stream).

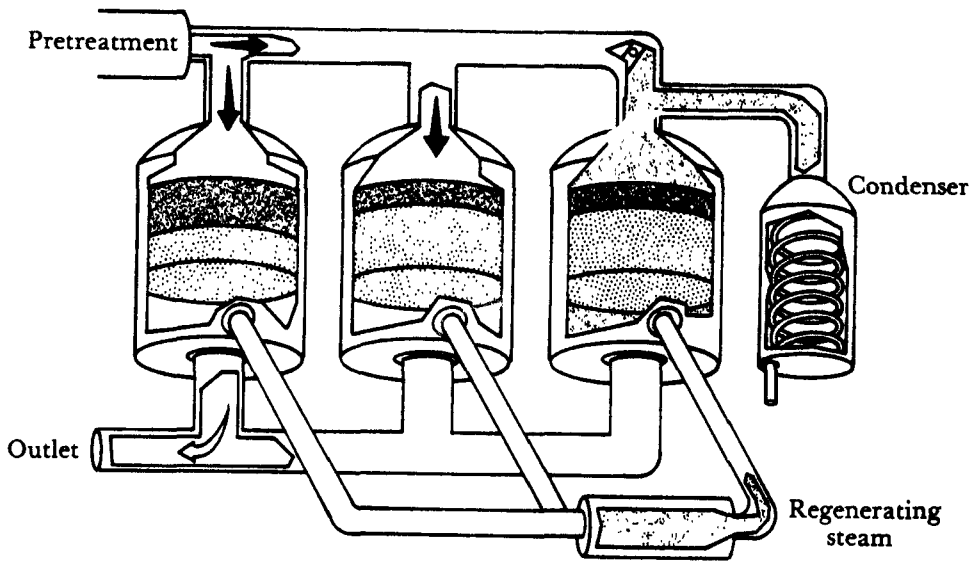


Figure 5-16. Three bed system.

After a predetermined length of time, referred to as the cycle time, the solvent-laden air stream is directed to the second adsorber by a series of valves. Steam is then injected into the first bed to remove the adsorbed vapors. The steam and desorbed vapors are then usually sent to a recovery system. If the solvents are immiscible in water, they can be separated by condensing the exhaust and decanting off the solvent. If the solvents are miscible in water, distillation may be required. Before the first adsorber is returned to service, cooling and drying of the carbon should be provided. This will ensure against immediate breakthrough occurring from the "hot, wet" carbon bed. This can be accomplished by venting the solvent-laden air stream through the hot, wet adsorber, then to the on-line adsorber to maintain a high removal efficiency.

Regenerable fixed carbon beds are usually between 0.3 and 1.2 m (1 to 4 ft) thick. The maximum adsorbent depth of 1.2 m is based on pressure drop considerations (Vic Mfg. Co.). Superficial gas velocities through the adsorber range from 6.0 to 30.0 m/min (20 to 100 ft/min) with 30.0 m/min being a maximum

permissible flow rate. Pressure drops normally range from 750 to 3730 Pa (3 to 15 in. H_2O) depending on the gas velocity, bed depth and carbon particle size (Bethea, 1978). For specific applications, graphs similar to that in Figure 5-11 are supplied by the carbon manufacturer to compute the pressure drop.

The two types of fixed bed adsorbers are distinguished by bed orientation in relation to air flow. The first is referred to as a *vertical flow adsorber*. The bed length is vertical as is the direction of air flow. The air stream usually flows downward. This system is shown in Figure 5-16. These units are suited to handle flows up to 1.4 to 2.0 m^3/s (3000 to 4000 cfm) per adsorber. Figure 5-17 shows a three bed vertical system used to recover 34 kg/m (75 lbs/hr) of trichloroethylene from a vapor degreasing operation. Each vessel is 1.2 m (48 in.) in diameter, contains 255 kg (560 lbs) of carbon approximately 10 cm (4 in.) deep and handles 1.7 m^3/s (3500 scfm) air flow. One vessel is being desorbed each hour using 102 kg (225 lbs) of steam at 103 kPa (15 psig). The pressure drop across the system is approximately 3 kPa (12 in. H_2O) (Vic Mfg. Co. with permission).

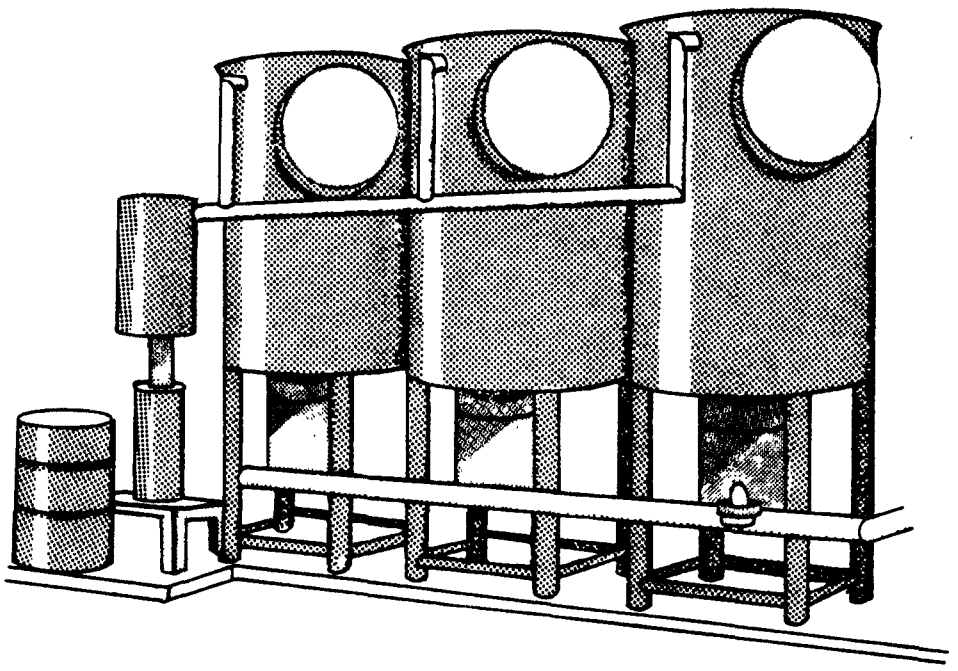


Figure 5-17. Three bed vertical system.

For larger flow rates, *horizontal flow adsorbers* are used. Structurally they are more suited to handle the larger air volumes. In horizontal flow units, the bed length is horizontal as is the direction of the incoming air stream. The air stream flows across the bed and down. This system is illustrated in Figure 5-18. Adsorbers of this type are manufactured as a package system capable of handling flow rates up to 1150 m^3/s (40,000 cfm). Larger units must be engineered and fabricated for the specific application. Figure 5-19 shows a three bed horizontal system used to recover 1180 kg/hr (2600 lb/hr) of toluene from a rotogravure operation. Each vessel is 7 m (22 ft) long and 3 m (10 ft) in diameter and contains 9000 kg (20,000 lbs) of carbon packed to a 1 m (3 ft) bed depth. The system handles 21 m^3/s (44,000 scfm) air flow with a pressure drop of 6 kPa (24 in. H_2O). One vessel is desorbed each 45 minutes using 3600 kg/hr (8000 lbs/hr) of steam at 103 kPa (15 psig) pressure (Vic Mfg. Co. with permission).

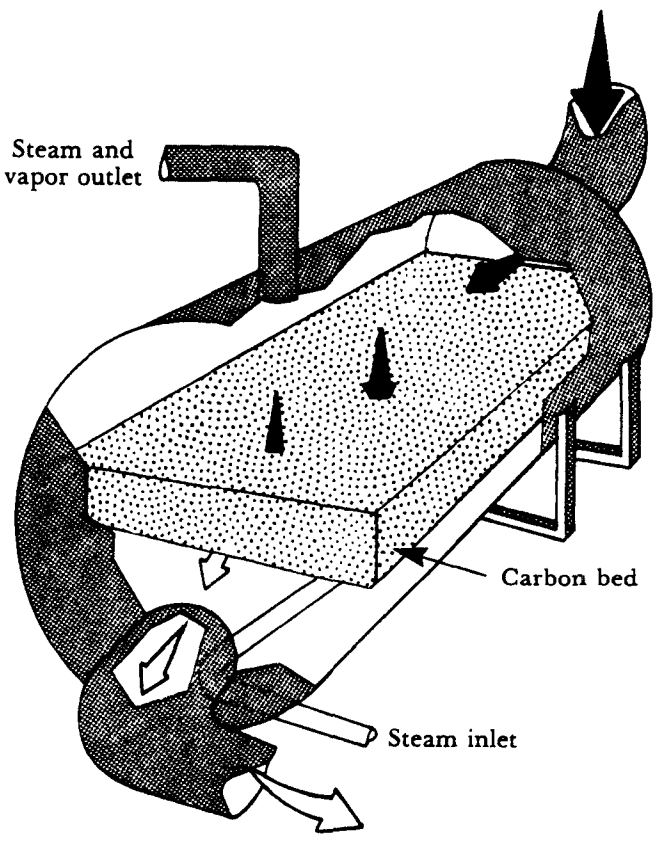


Figure 5-18. Horizontal bed.

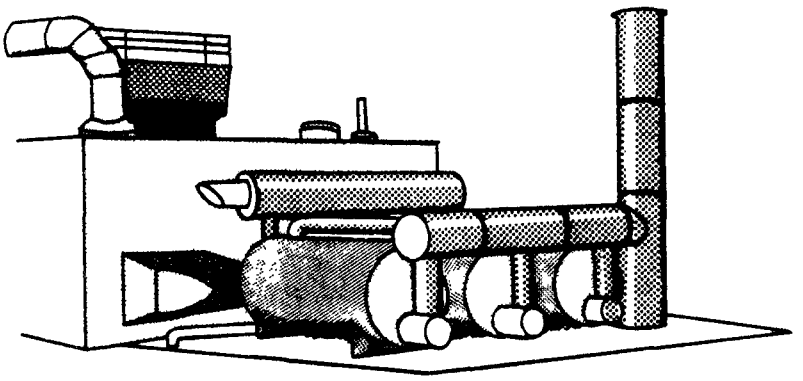


Figure 5-19. Three bed horizontal system.

Moving Bed Adsorbers

Moving bed systems can also be used to obtain a higher degree of utilization of the carbon bed than is possible with a fixed bed. In moving bed systems, the solvent-laden vapor stream passes only through the unsaturated portion of the carbon bed. This reduces the distance (thus pressure drop) the air stream travels through the bed.

One design of a moving (rotary) bed system is illustrated in Figure 5-20 (Sutcliffe Speakman Co., 1963). The device consists of four cylinders which are in constant rotation. The granular carbon is held in place between two cylinders made of steel screening or perforated sheet metal. This bed is placed between inner and outer cylinders which are impervious to air flow except at slots near their ends.

The slots on the outer cylinder act as solvent-laden inlets. They permit the air stream to pass into the annular section where the carbon is located. The solvent-laden air stream passes through the carbon bed and purified air exits out the inner slots. The carbon bed is broken into sections. The cylinders rotate such that when the proper degree of saturation is reached the bed is desorbed. Desorption occurs by injecting steam in through the slots on the inner cylinder. Steam and desorbed vapors exit through the slots on the outer cylinder. During each rotation of the cylinder, each segment of the carbon bed undergoes both an adsorption and desorption cycle.

Because of the continuous adsorbing and desorbing process, bed utilization is improved. The air stream is no longer required to pass through the top, saturated portion; or the bottom, idle portion of the bed. The air stream passes only through the active, mass transfer portion of the bed. Therefore, shorter and more compact beds may be used which reduce the pressure drop. The disadvantages are wear on moving parts and maintaining air tight seals on moving parts.

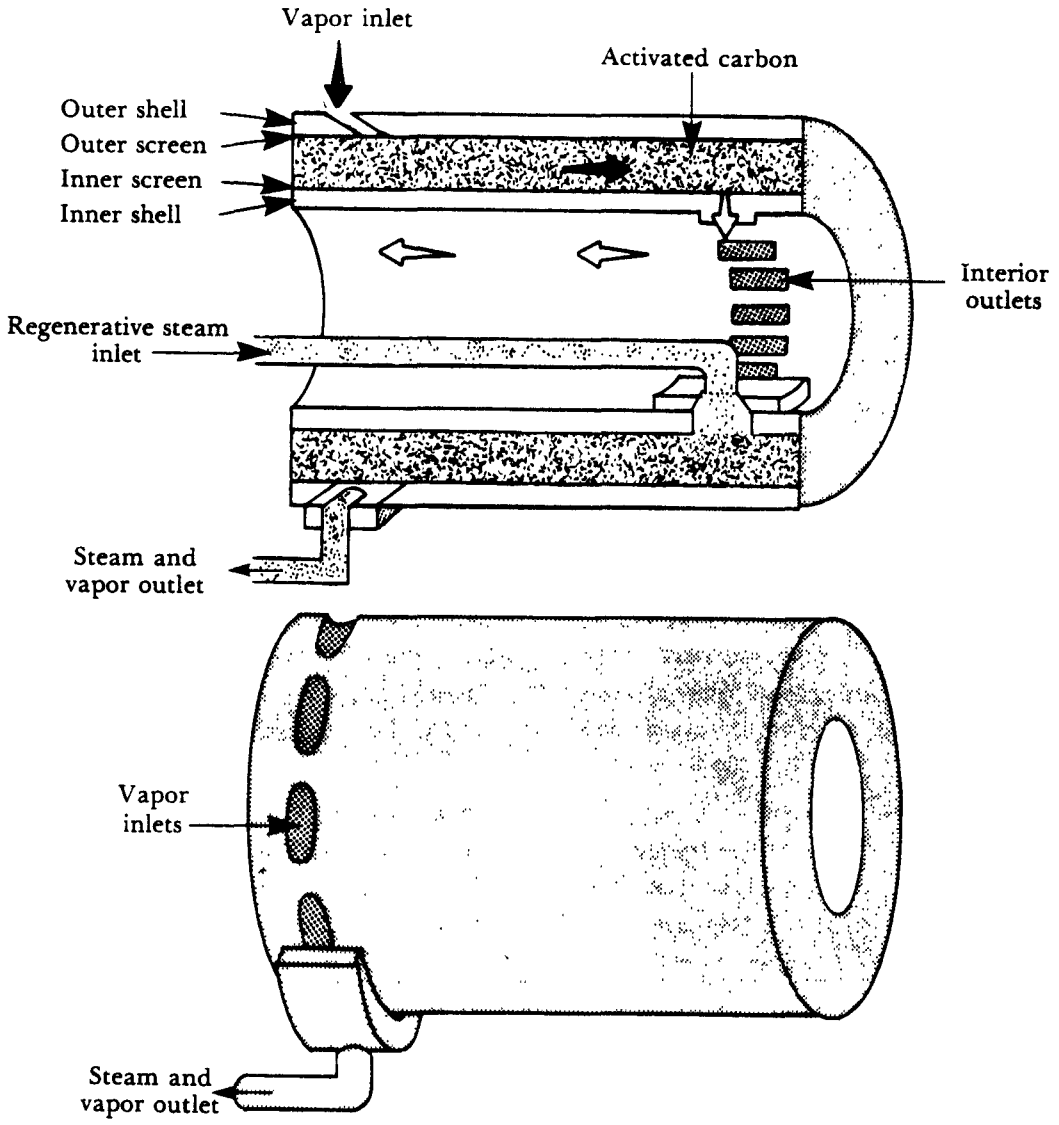


Figure 5-20. Rotary bed system.

Fluidized Bed Adsorbers

A fluidized bed adsorption system operates in the same physical manner as a tray scrubber. Instead of liquid flowing down the column from tray to tray, granular activated carbon is used. Figure 5-21 shows one recently developed fluidized bed system which is being marketed by Union Carbide. The solvent-laden air stream is introduced at the middle of the tower. Then it passes up through the tower fluidizing the activated carbon in a series of trays. The carbon then flows down through the vessel from tray to tray until it reaches the desorption section. Regeneration is accomplished in the bottom half of the vessel and the activated carbon is air conveyed back to the top of the tower.

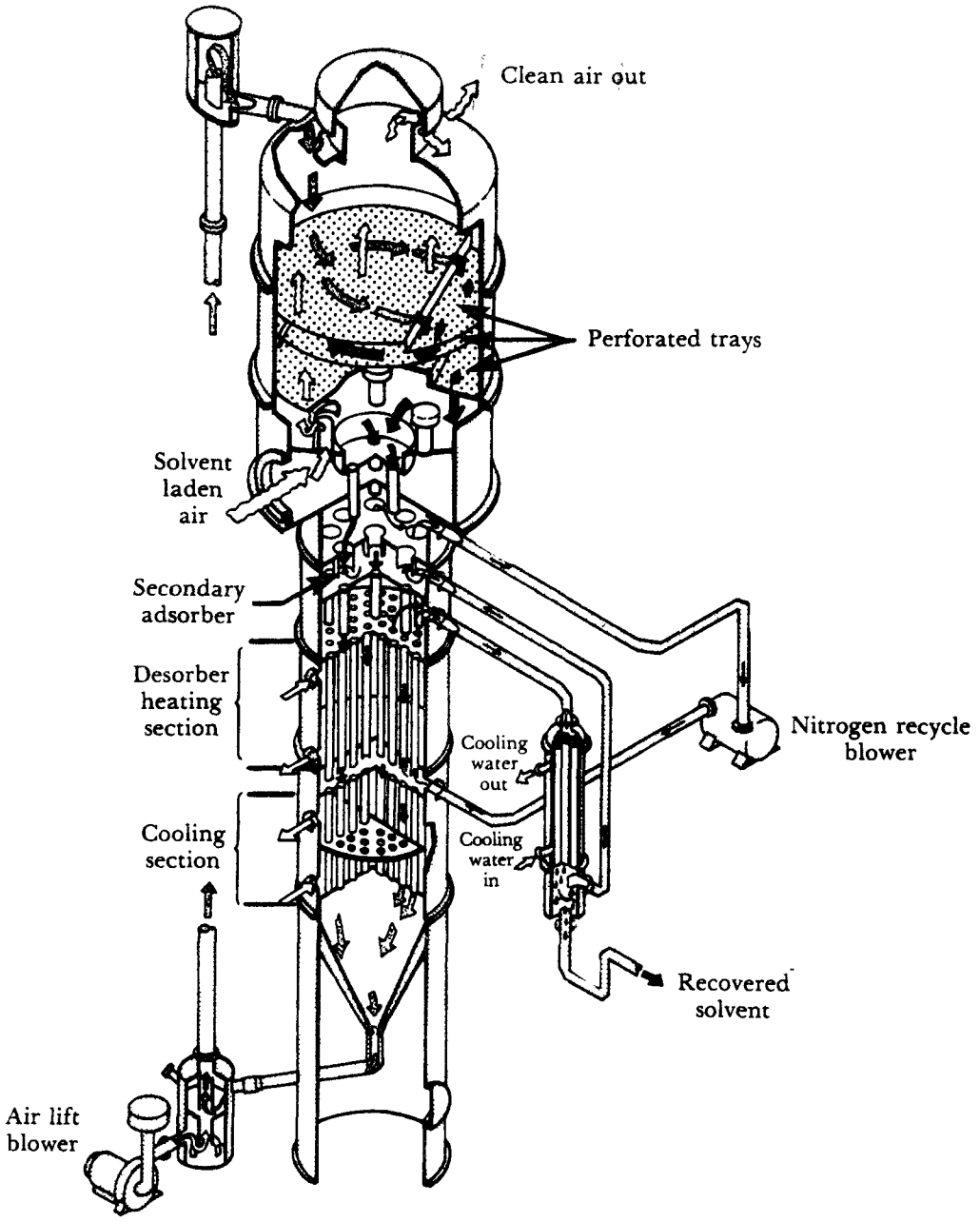


Figure 5-21. Fluidized bed adsorber.

As with the moving bed, the fluidized bed also provides continuous operation and more efficient utilization of the adsorbent. The need for multiple vessels is eliminated, which can greatly reduce the cost of the system. Gas velocities around 1 m/s (196 ft/min) are needed to fluidize the bed. These are 2 to 4 times the velocities achieved in fixed bed systems. This allows for use of a much smaller vessel for comparable air flow and helps to achieve uniform gas distribution.

The main disadvantage with fluidized bed adsorption systems is the high attrition (wear) losses of the granular activated carbon. Recently a new "beaded" activated carbon was developed in Japan. The beaded shape is inherently stronger and has better fluidity properties than granular carbon. The beaded carbon has been used in a number of installations (mostly in Japan) and is reported to reduce the attrition losses to 2 to 5% per year as compared to 10% for fluidized granular carbon (Union Carbide).

The following example illustrates the use of the principles and general rules of practice discussed in this chapter in designing an adsorption system.

Example 5-3

A solvent degreaser is designed to recover toluene from a 3.78 m^3/s (8000 acfm) air stream at 25 °C (77 °F) and atmospheric pressure. The company is planning to use a two bed carbon adsorption system with a cycle time of 4 hours. The maximum concentration of toluene is kept below 50% of the lower explosive limit for safety purposes. Given Figure 4-4 in the 415 Student Workbook, the adsorption isotherm for toluene, and the additional operational data, estimate:

1. the amount of carbon required for a 4-hour cycle
2. square feet of surface area required based on a 0.508 m/s (100 fpm) maximum velocity
3. depth of the carbon bed

Given: LEL for toluene = 1.2%
molecular weight of toluene = 92.1 kg/kg mol
carbon density = 480 kg/m^3 (30 lb/ft^3)

Solution:

1. To compute amount of carbon required, first, calculate the toluene flow rate.

$$(3.78\ m^3/s)(50\%)(1.2\%) = 0.023\ m^3/s \text{ toluene}$$

To determine the saturation capacity of the carbon, calculate the partial pressure of toluene at the adsorption conditions.

$$p = YP$$

$$= \left(\frac{0.023\ m^3/s}{3.78\ m^3/s}\right)(14.7\ \text{psia})$$

$$= 0.089\ \text{psia}$$

From Figure 4-4 in the 415 Student Workbook, the saturation capacity of the carbon is 40% or 40 kg toluene per 100 kg of carbon at 0.089 psia.

The flow rate of toluene is:

$$\left(0.023\ \frac{m^3}{s}\right)\left(\frac{kg\ mol}{22.4\ m^3}\right)\left(\frac{273\ K}{350\ K}\right)\left(\frac{92.1\ kg}{kg\ mol}\right)$$

$$= 0.074\ kg/s \text{ of toluene}$$

The amount of carbon at saturation for a 4-hour cycle is:

$$\left(0.074\ \frac{kg}{s}\ \text{toluene}\right)\left(\frac{100\ \text{kg carbon}}{40\ \text{kg toluene}}\right)\left(\frac{3600\ s}{hr}\right)(4\ hr)$$

$$= 2664\ \text{kg of carbon}$$

The working charge of carbon can be estimated by doubling the saturation capacity. Therefore, the working charge is:

$$\text{working charge} = (2)(2664\ \text{kg of carbon})$$
$$= 5328\ \text{kg of carbon}$$

2. The square feet of superficial surface area is the surface area set by the maximum velocity of 0.508 m/s (100 fpm) through the adsorber.

   The required surface area is:

$$A = \frac{Q}{\text{maximum velocity}}$$

$$= \frac{3.78\ m^3/s}{0.508\ m/s}$$

$$= 7.44\ m^2$$

For a horizontal flow adsorber this would correspond to a vessel approximately 2 m (6.6 ft) in diameter and 4 m (13.1 ft) in length to give 8.0 m^2 (87 ft^2) surface area. This would supply more than the required area.

The flow rate is too high to be handled by a single vertical flow adsorber. An alternative would be to use three vessels, two adsorbing while one is being regenerated. Each vessel then must be sized to handle 1.89 m^3/s (4000 acfm). The area required for a limiting velocity of 0.508 m/s is:

$$\text{Area to handle half flow} = \frac{1.89\ m^3/s}{0.508\ m/s}$$

$$= 3.72\ m^2$$

This cross sectional area corresponds to a vessel diameter of:

$$d_r = \left(\frac{4}{\pi}A\right)^{0.5}$$

$$= \left[\left(\frac{4}{\pi}\right)(3.72)\right]^{0.5}$$

$$= 2.18\ m\ (7\ ft)$$

3. The volume that the carbon would occupy in the horizontal bed system is:

$$\text{Volume of carbon} = \text{weight} \times \frac{1}{\text{density}}$$

$$= 5328 \text{ kg} \times \frac{\text{m}^3}{480 \text{ kg}}$$

$$= 11.1 \text{ m}^3$$

Note: for the three bed vertical system the volume of each bed would be half this or 5.55 m^3.

4. The depth of carbon for the horizontal bed is given by:

$$\text{Depth of carbon} = \frac{\text{Volume of carbon}}{\text{Cross sectional area of adsorber}}$$

$$= \frac{11.1 \text{ m}^3}{7.44 \text{ m}^2}$$

$$= 1.49 \text{ m}$$

Note: the depth for the three bed vertical system is the same, since **both** the volume and area are halved.

This bed depth is greater than the recommended maximum of 1.2 m (4 ft). This depth would correspond to a pressure drop of approximately 10 kPa (40 in. H_2O) assuming the carbon to be 4 × 10 mesh from Figure 5-11. In practical design situations the operating parameters (such as bed depth, vessel size, cycle time, etc.) would be varied to reduce this excessive pressure drop and to determine the most cost effective alternatives.

The above solution is based on a number of adsorber design maximum and minimum rules of thumb. It is intended as a guide to illustrate how to "red flag" any parameters which may be greatly exceeded.

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Chapter 6
Condensation

Introduction

Condensation is the process of reducing a gas or vapor to a liquid. Any gas can be reduced to a liquid by lowering its temperature and/or increasing its pressure. The most common approach is to reduce the temperature of the gas stream, since increasing the pressure of a gas is very costly (EPA, 1973).

Condensers are simple, relatively inexpensive devices that normally use water or air to cool and condense a vapor stream. Since these devices are usually not capable of reaching low temperatures (below 80°F), high removal efficiencies of most gaseous pollutants are not obtained unless the vapors will condense at high temperatures. Condensers are typically used as pretreatment devices. They are used ahead of incinerators, absorbers, or adsorbers to reduce the total gas volume to be treated by more expensive control equipment. Used in this manner, they help reduce the overall cost of the control system.

Condensation Principles

When a hot vapor stream contacts a cooler surface, heat is transferred from the hot gases to the cooler surface. As the temperature of the vapor stream is cooled, the average kinetic energy of the gas molecules is reduced. Also, the volume that these vapors occupy is reduced. Ultimately the gas molecules are slowed down and crowded together so closely that the attractive forces (van de Waals' forces) between the molecules cause them to condense to a liquid.

The two conditions which aid condensation are: low temperatures so that the kinetic energy of the gas molecules are low; and high pressures so that the molecules are brought close together. The actual conditions at which a particular gas molecule will condense depends on its physical and chemical properties.

Condensation occurs when the partial pressure of the pollutant in the gas stream equals its vapor pressure as a pure substance at operating conditions.

Condensation of a gas can occur in three ways: first, at a given temperature, the system pressure is increased (compressing the gas volume) until the partial pressure of the gas equals its vapor pressure; second, at a fixed pressure, the gas is cooled until the partial pressure equals its vapor pressure; or third, by using a combina-

tion of compression and cooling of the gas until its partial pressure equals its vapor pressure. These processes are illustrated in Figure 6-1, a typical vapor pressure diagram for a pure substance.

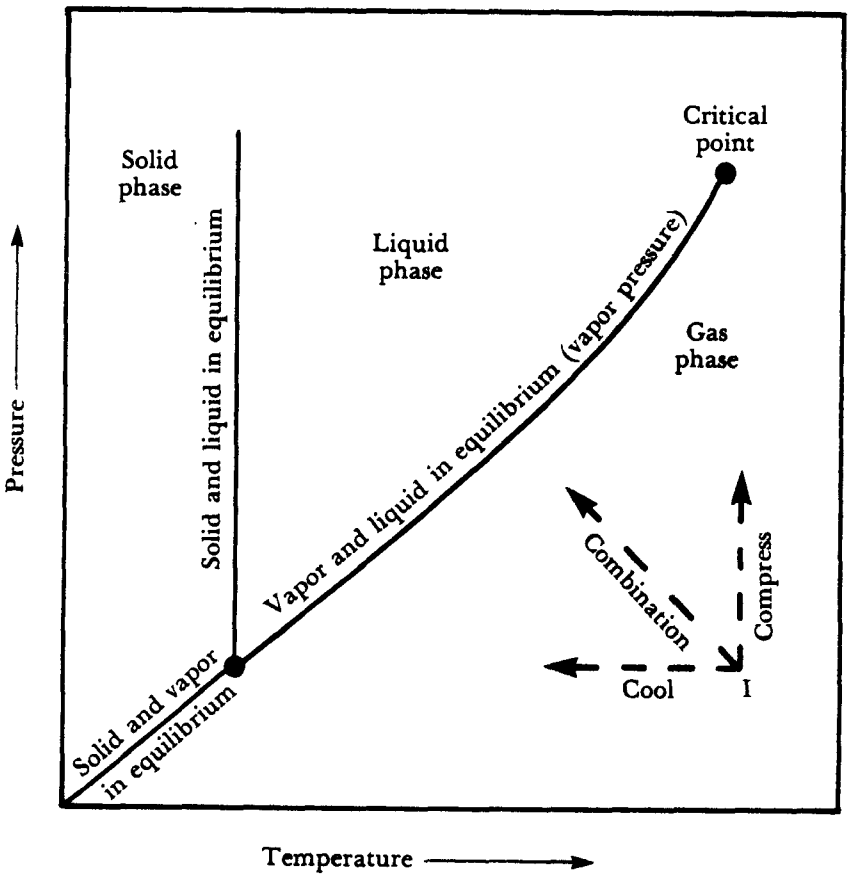


Figure 6-1. Typical vapor pressure curve.

In Figure 6-1, point I is the initial temperature and pressure of a gas. The dotted lines indicate the paths a quantity of gas would follow to reach the vapor pressure curve. Points on the vapor pressure curve are also referred to as *dew points*. The dew point is defined as the condition at which gas is ready to condense into the first drop of liquid.

Also on Figure 6-1, the critical point is plotted. Each substance has a critical temperature and critical pressure. The critical temperature is important in that it is a maximum temperature above which the gas will not condense, no matter how great a pressure is applied. The pressure required to liquify a gas at its critical temperature is the critical pressure.

Once the gas conditions (temperature, pressure, and volume) equal those on the vapor pressure line, liquid begins to condense. From this point on, the gas-liquid mixture follows the vapor pressure line. If the mixture is cooled continuously, the partial pressure of the remaining gas will always equal the vapor pressure. This is

important since even though the contaminated gas is being condensed, it still has a certain partial pressure indicating that uncontrolled vapors are being emitted from the condenser. For most practical applications, the vapor-liquid equilibrium restricts the use of condensers as primary air pollution control devices. Unless very low temperatures or high pressures are attained, condensers are not capable of reducing the pollutant concentration to within acceptable emission limits.

Practically, temperature is the only process variable which governs the effectiveness of a condenser. In industrial applications, increasing the system pressure is very costly and therefore rarely used for condensation. At the operating pressure of the system, the outlet temperature from the condenser determines the maximum removal efficiency. Therefore, condensers cannot be used in the same manner as other gaseous pollutant control devices. For example, condensers cannot be used in series like adsorbers or absorbers to further reduce outlet concentration unless the outlet temperature of the second condenser is lower than the previous one. Increasing gas residence time or decreasing flow rates in the condenser does not add to the theoretical achievable efficiency as these operations do in incinerators, adsorbers, and absorbers.

Condensers

Condensers fall into two basic categories; *contact* and *surface condensers*. In a contact condenser the coolant and vapor stream are physically mixed. They leave the condenser as a single exhaust stream. In a surface condenser, the coolant is separated from the vapors by tubular heat transfer surfaces. The coolant and condensed vapors leave the device by separate exits. Surface condensers are commonly called *shell-and-tube heat exchangers*. The temperature of the coolant is increased, so these devices also act as heaters.

Direct Contact Condensers

Contact condensers are simple devices such as *spray towers*, *steam* or *water jet ejectors*, and *barometric condensers*. These devices bring the coolant, usually water, into direct contact with the vapors as illustrated in Figure 6-2. The liquid stream leaving the condenser contains the coolant plus the condensed vapors. If the vapor is soluble in the coolant then absorption also occurs. Absorption increases the amount of contaminant that can be removed at the given conditions.

(a) spray contact condenser

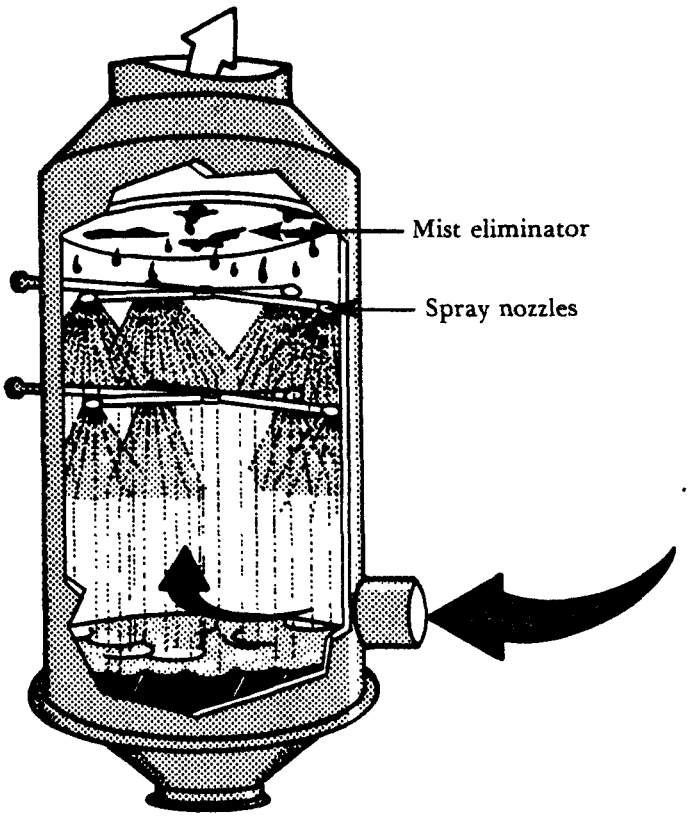
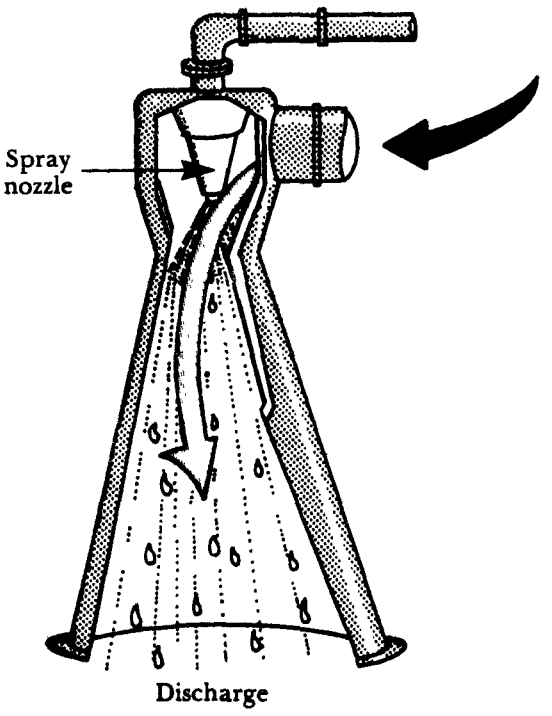
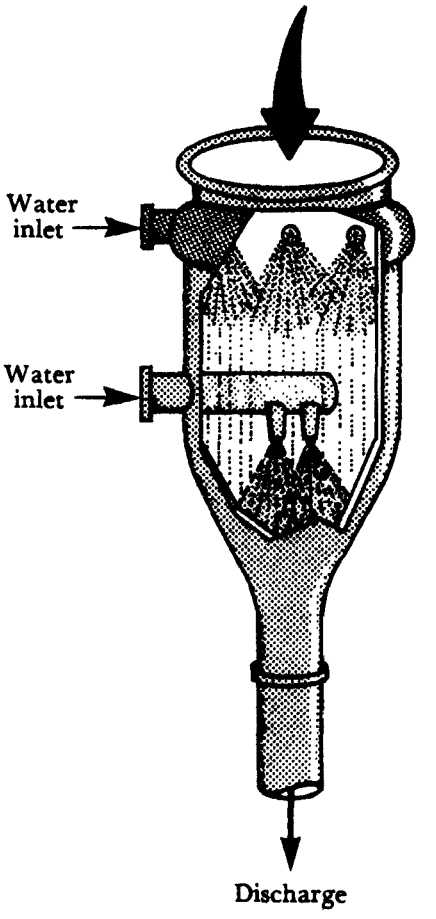


Figure 6-2. Direct contact condensers:
(a) spray, (b) jet ejector, and (c) barometric.

(b) jet ejector condenser



(c) barometric condenser



Spray tower condensers (Figure 6-2a) are normally the same as the spray absorbers discussed in Chapter 4. The vapors enter the bottom of the tower while coolant is sprayed down over them. Baffles are usually added to ensure adequate contact between coolant and vapors. Ejectors (Figure 6-2b) and barometric condensers (Figure 6-2c) operate in a similar manner. The difference being that they use liquid sprays to move the vapor stream. In both of these devices the coolant is sprayed into a venturi throat creating a vacuum which moves the vapor stream through the condenser.

Surface Condensers

Surface condensers are usually in the form of *shell-and-tube heat exchangers* (Figure 6-3). The device consists of a circular or oval cylindrical shell into which the vapor stream flows. Inside the shell are numerous small tubes through which the coolant flows. Vapors contact the cool surface of the tubes, condense, and are collected, while noncondensed vapors are sent for further treatment.

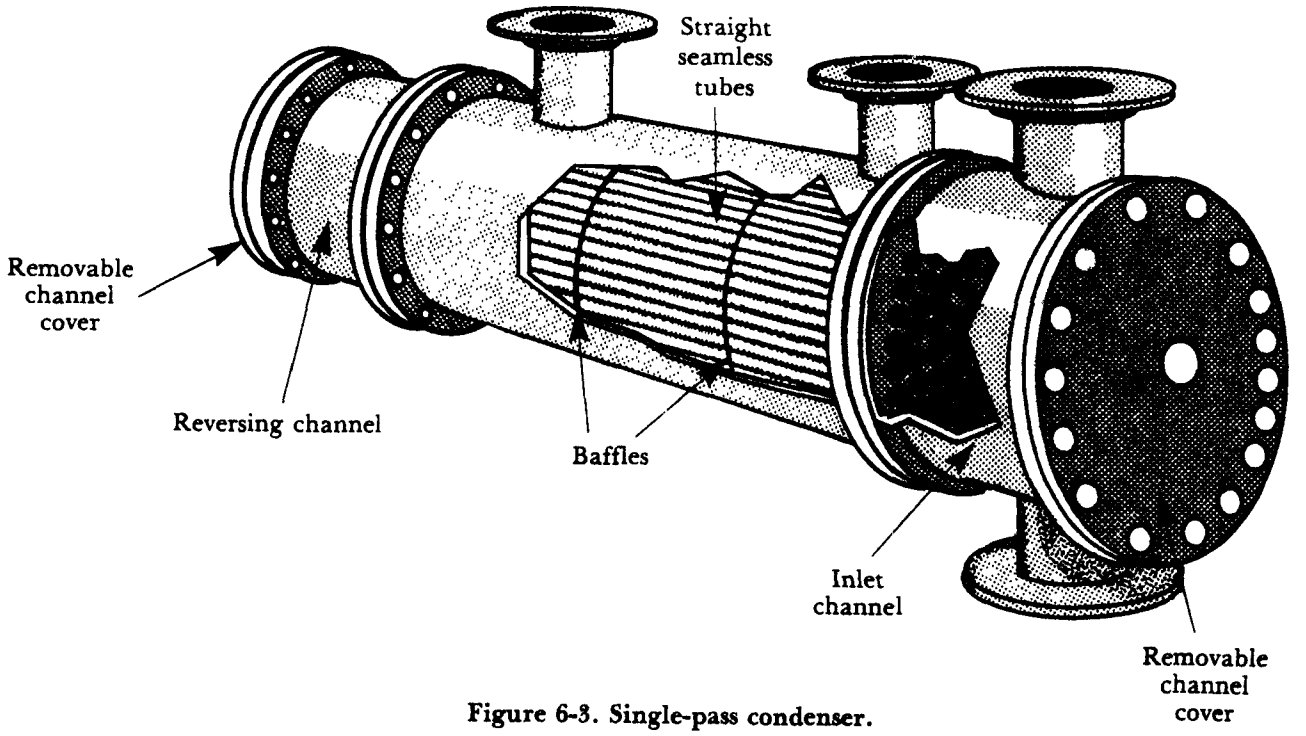


Figure 6-3. Single-pass condenser.

Figure 6-3 is a diagram of a *single-pass heat exchanger,* where the entire stream of coolant flows through all the parallel tubes. The cooling liquid enters and makes **one pass** through the tubes or **tube side** of the exchanger. The uncondensed vapor stream enters and makes one pass on the **shell side** of the exchanger. The single pass exchanger is limited in that it requires a large number of tubes and low gas velocities through the exchanger to provide adequate heat transfer. By using a *multipass system*, shorter tube lengths, higher gas velocities through the exchanger, and improved heat transfer can be achieved (McCabe, 1967).

Figure 6-4 illustrates two types of multipass heat exchangers, referred to as *1-2* and *2-4* heat exchangers. The first digit refers to the number of passes the vapor makes on the shell side, while the second digit indicates the number of tube side passes. Both of these designs give improved performance over the single-pass exchanger. The 2-4 heat exchanger is capable of higher gas velocities and better heat transfer than the 1-2 heat exchanger. Adding more passes does have disadvantages however. These disadvantages are: the exchanger construction is more complicated; friction losses are increased due to the higher velocities; and exit and entrance losses are multiplied.

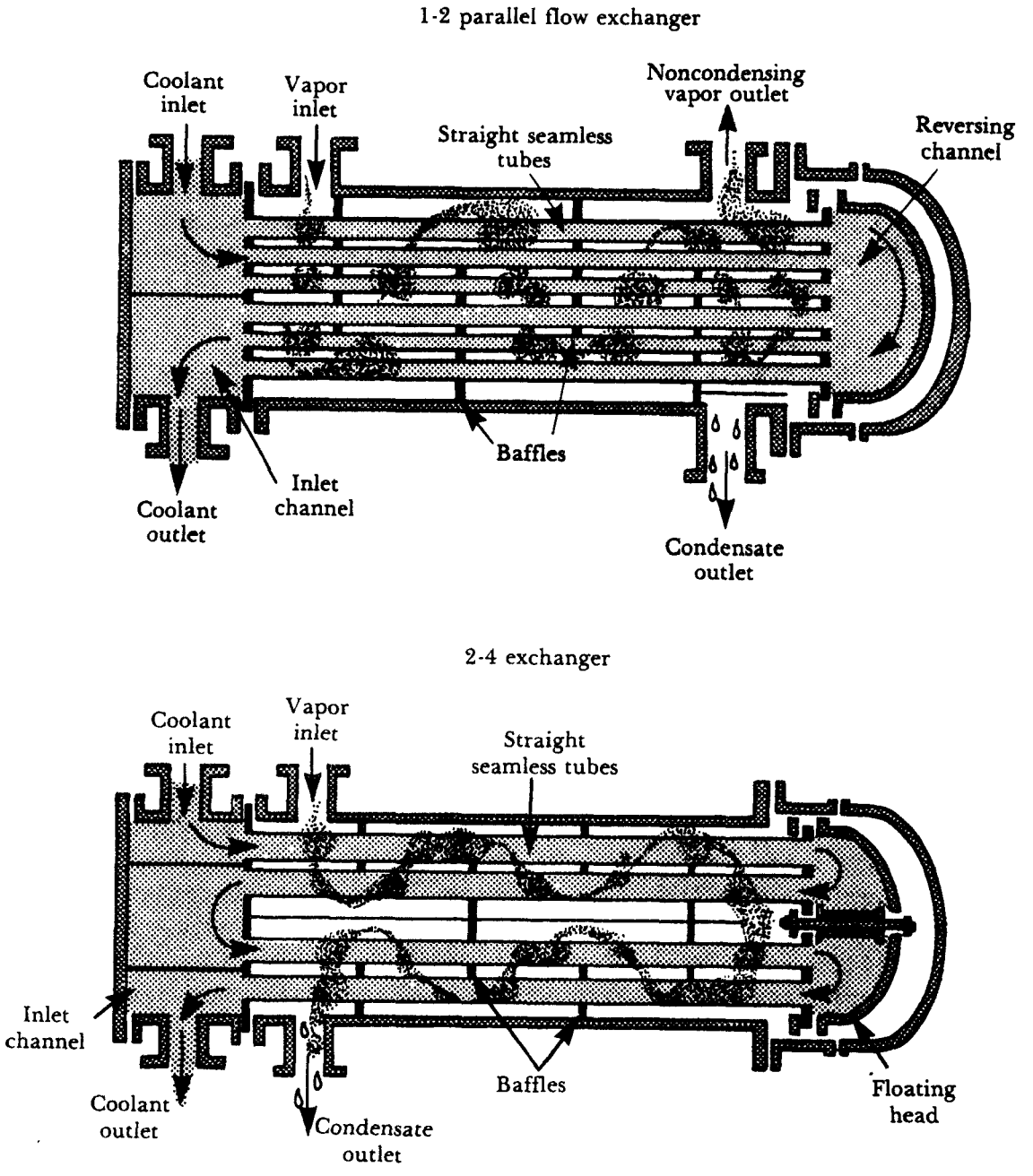


Figure 6-4. Simplified air flow in multipass exchangers.

Condensation applications normally require large temperature differentials between vapor and coolant. These temperature variances can cause the tubes to expand or contract. This expansion stress can eventually cause the tubes to buckle or pull loose from the shell, destroying the condenser. Floating head construction is commonly used to avoid condenser expansion stress damage. In a floating head, one end of the tube bundle is mounted so that it is structurally independent from the shell as shown in Figure 6-4. This allows the tubes to expand and contract within the shell.

Water is generally the coolant used in condensers. However, short supply and expense to treat water make it an uneconomical choice in some cases. In these cases, air-coolers are used. The specific heat of air is only about 0.25 Btu/lb•°F, approximately one-fourth that of water. Therefore, air condensers must be very large compared to water condensers.

To conserve space and reduce the cost of equipment in these cases, heat exchangers with *extended surfaces* can be used. In these devices, the outside area of the tube is multiplied or extended by adding *fins* or *disks*. Figure 6-5 illustrates two types of finned tubes. In extended surface condensers, the vapor is condensed inside the tube while air flows around the outside contacting the extended surfaces.

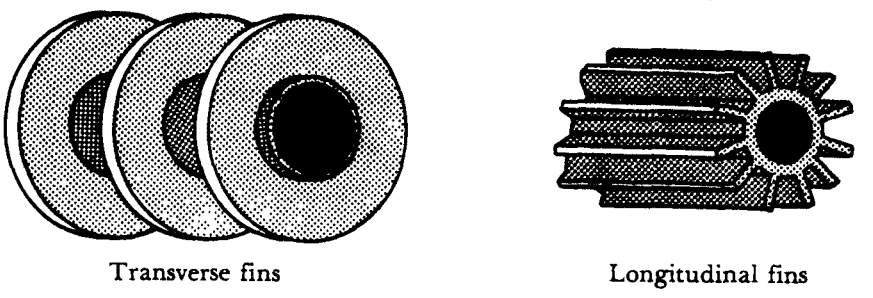


Figure 6-5. Extended surface tubes.

Comparison of Contact and Surface Condensers

Since in contact condensers coolant is merely sprayed on the vapors, these systems are simpler in design, less expensive, and more flexible in application than surface condensers. However, contact condensers require more coolant, and due to direct mixing, produce 10 to 20 times the amount of wastewater (condensate) than surface condensers. Since the wastewater from a contact condenser is contaminated with vapors, it cannot be reused posing a water disposal problem. If the condensed vapors have a recovery value, surface condensers are usually used since the condensate can be recovered directly.

Design of Condensers

Condensers reduce the temperature (heat content) of a hot vapor stream by contacting it with a cooler liquid or air stream. Heat is being transferred from a hot to a colder fluid; thus the process is termed a *heat transfer* or *heat exchange* operation. Condensers are designed using the same basic principles and empirical relationships used for heat exchangers.

Procedures used to design simple heat exchangers are well defined and can be found in numerous texts, such as: Perry (1973); McCabe (1967); or Kern (1950). In adapting these procedures to designing a condenser, an additional variable complicates the calculations. Heat exchanger designs are based on heat transfer theory. In a condenser, however, both **heat** and **mass** are being transferred. As the vapor stream passes through the condenser, **both** its temperature and its composition are changing. Even the simplest of condenser design procedures found in Perry (1973) or Bell (1972) are complicated and require trial and error solutions. A rigorous procedure for the design of condensers will not be presented in this chapter. Simplified heat balance calculations are used to estimate the important parameters. The following equations are not intended to be used as a design method, but only as rough estimates for evaluation purposes.

Heat Balance

The first step in analyzing any heat transfer process is to set up a heat balance relationship. For a condensation system, the heat balance can be expressed as:

Heat in = Heat out

Heat required to reduce vapors to the dew point	+	Heat required to condense vapors	=	Heat needed to be removed by the coolant

This heat balance is written in equation form as:

(Eq. 6-1) $$q = \dot{m}\bar{C}_p(T_{G1} - T_{dew\ point}) + \dot{m}H_v = L\bar{C}_p(T_{L2} - T_{L1})$$

Where: q = heat transfer rate, Btu/hr*
$\dot{m}$ = mass flow rate of vapor, lb/hr
L = mass flow rate of liquid coolant, lb/hr
$\bar{C}_p$ = average specific heat of a gas or liquid, Btu/lb•°F
T = temperature of the streams; G for gas and L for liquid coolant, °F
H_v = heat of condensation or vaporization, Btu/lb

*The units used in this chapter are in English, since most readily available heat transfer data is in English units.

In Equation 6-1, the mass flow rate ($\dot{m}$) and inlet temperature (T_{G1}) of the vapor stream are set by the process exhaust stream. The temperature of the coolant entering the condenser (T_{L1}) is also set. The average specific heats ($\bar{C}_p$) of both streams, the heat of condensation (H_v), and the dew point temperature can be obtained from chemistry handbooks. Therefore, only the amount of coolant (L) and its outlet temperature are left to be determined. If either one of these terms are set by process restrictions (i.e. only x pounds an hour of coolant are available or the outlet temperature of coolant must be below a set temperature), then the other term can be solved for directly.

Equation 6-1 is applicable for direct contact condensers and should be used **only** to obtain **rough estimates**. Equation 6-1 has a number of limitations: the specific heat ($\bar{C}_p$) of a substance is dependent on temperature; and the temperature throughout the condenser is constantly changing. Also, the dew point of a substance is dependent on its concentration in the gas phase; and since $\dot{m}$ is constantly changing (vapors being condensed) the dew point temperature is constantly changing. Finally, no provision is made for cooling the vapors below their dew point. An additional term would have to be added to the left side of Equation 6-1 to account for this amount of cooling.

Surface Condensers

In a surface condenser or heat exchanger, heat is transferred from the vapor stream to the coolant through a heat exchange surface. The rate of heat transfer depends upon three factors: total cooling surface available, resistance to heat transfer, and mean temperature difference between condensing vapor and coolant. This can be expressed mathematically by:

(Eq. 6-2)

$$q = UA\Delta T_m$$

Where: U = overall heat transfer coefficient, Btu/°F•ft²•hr
A = heat transfer surface area, ft²
ΔT_m = mean temperature difference, °F

U, the overall heat transfer coefficient is a measure of the total resistance heat experiences while being transferred from a hot body to a cold body. In a shell-and-tube condenser, cold water flows through the tubes causing vapor to condense on the outside surface of the tube wall. Heat is transferred from the vapor to the coolant. The ideal situation for heat transfer would be where heat is transferred from the vapor to the coolant without any heat loss (heat resistance). Figure 6-6 shows the typical heat resistances occurring during heat transfer in a shell-and-tube condenser. Every time heat moves through a different medium, it encounters a different and additional heat resistance. These heat resistances (Figure 6-6) are: throughout the condensate, through any scale or dirt on the outside of the tube (fouling), through the tube itself, and through the film on the inside of the tube (fouling). Each of these resistances are individual heat transfer coefficients and must be added together to obtain an overall heat transfer coefficient.

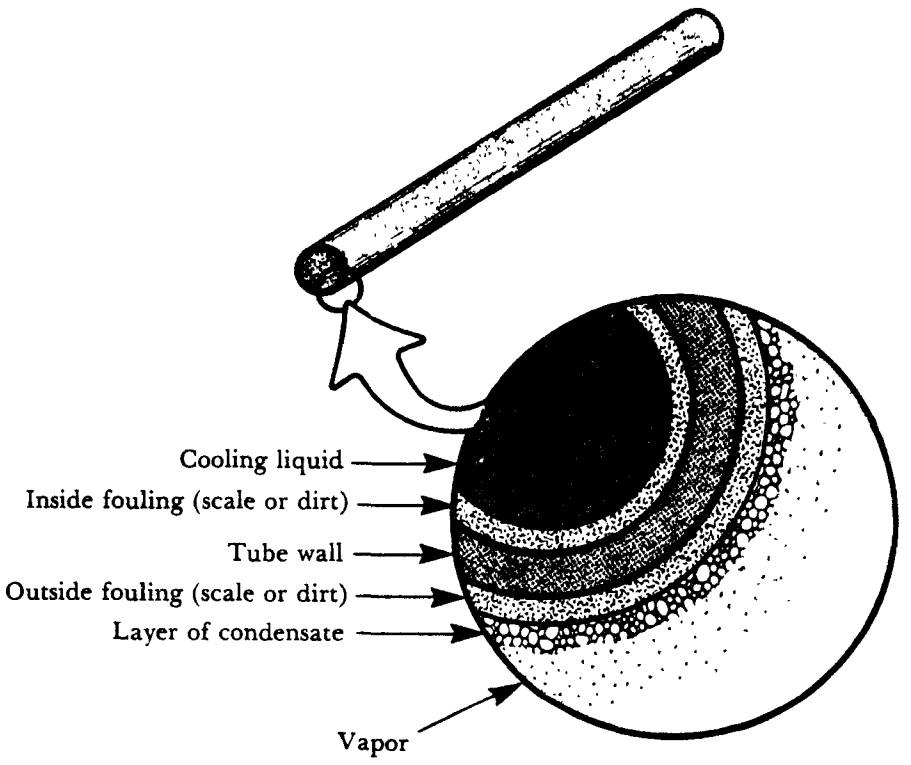


Figure 6-6. Resistance to heat transfer around a cooling tube.

A number of correlations exist to determine the individual heat transfer coefficient (Perry, 1973 and Kern, 1950). These correlations are usually presented in terms of dimensionless numbers such as the Reynolds Number. These correlations consider variables including geometry of the condenser, gas and liquid density and viscosity, and the mechanism of heat transfer (convection, conduction and radiation). Calculation of individual heat transfer coefficients require numerous data from the system. An estimate of an overall heat transfer coefficient can be used for preliminary calculations. Table 6-1 lists some typical values for overall heat transfer coefficients. Table 6-1 should be used only for preliminary estimating purposes.

Table 6-1. Typical overall heat transfer coefficients in tubular heat exchangers.

Condensing vapor* (shell side)	Cooling liquid (tube side)	U (Btu/°F•ft²•hr)
Alcohol vapor	Water	100-200
High boiling hydrocarbons (vacuum)	Water	20-50
Low boiling hydrocarbons	Water	80-200
Organic solvents	Water	100-200
Organic solvents with high percent of noncondensables present	Water or brine	20-60
Naptha	Water	50-75
Stabilizer reflux vapors	Water	80-120
Sulfur dioxide	Water	150-200
Tall oil derivatives, vegetable oil vapors	Water	20-50
Steam	Feedwater	400-1000

*Note: for water-water (liquid-liquid) heat exchanger (no phase change) the values for U range between 200-250.

In a surface heat exchanger the temperature difference between the hot vapor and the coolant usually varies throughout the length of the exchanger. Therefore, a mean temperature difference (ΔT_m) must be used. For the special cases where the flow of both streams is completely cocurrent, the flow of both streams is completely countercurrent, or the temperature of one of the fluids remains constant (as is the case in condensing a pure liquid), the log mean temperature difference can be used. The temperature profiles for these three conditions are illustrated in Figure 6-7. The log mean temperature for countercurrent flow can be expressed as:

(Eq. 6-3)

$$\Delta T_m = \Delta T_{lm} = \frac{\Delta T_1 - \Delta T_2}{\ln\left(\frac{\Delta T_1}{\Delta T_2}\right)}$$

Where: ΔT_{lm} = log mean temperature

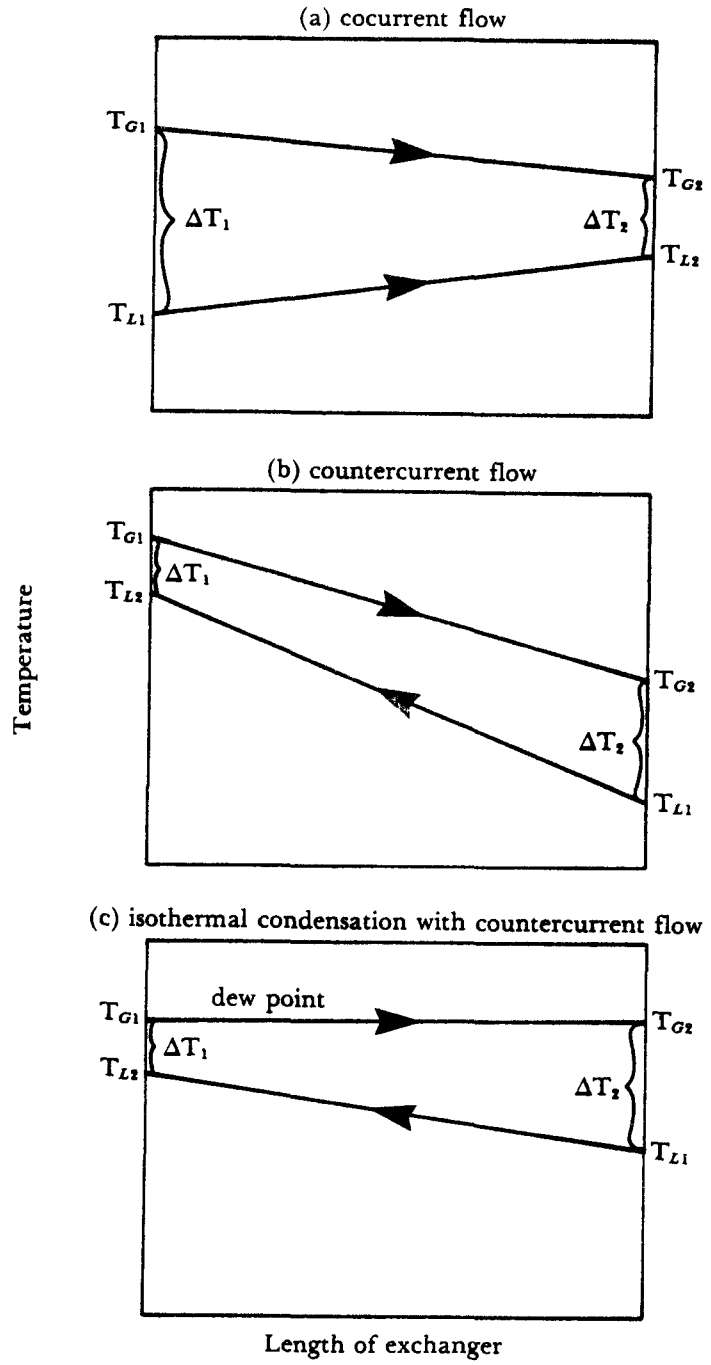


Figure 6-7. Temperature profiles in a heat exchanger for countercurrent flow, cocurrent flow, and isothermal condensation with countercurrent flow.

The value calculated from Equation 6-3 is used for single pass heat exchangers or condensers. For multiple pass exchangers a correction factor to the log mean temperature must be included. However, for the special case of isothermal condensation (no change in temperature) of a **single component** vapor, $T_{G1} = T_{G2}$ and is equal to the dew point temperature. No correction factor is needed (Perry, 1973).

In order to size a condenser, Equation 6-2 must be rearranged to solve for the surface area.

(Eq. 6-4)

$$A = \frac{q}{U\Delta T_{lm}}$$

Where: A = surface area of a shell-and-tube condenser, ft^2
q = heat transfer rate, Btu/lb
U = overall heat transfer coefficient, $Btu/°F \cdot ft^2 \cdot hr$
ΔT_{lm} = log mean temperature, °F

For the exact determination of the surface area, U in Equation 6-4 should be calculated from the individual resistances (coefficients). For a rough estimate the values in Table 6-1 can be used.

Equation 6-4 is valid **only** for *isothermal* condensation of a *single* component. This implies that the pollutant is a pure vapor stream comprised only of one specific hydrocarbon such as benzene, not a mixture of hydrocarbons. Nearly all air pollution applications involve multicomponents since air and any other gas are two-components. This complicates the design procedure for a condenser.

If the incoming vapor stream is above its dew point (superheated), it must first be cooled to its dew point. It is generally considered conservative to size the condenser as if the entire heat load (desuperheating and condensing) were transferred by condensation (Perry, 1973). In this case, the heat load, q, is computed by using the left side of Equation 6-1. The dew point temperature of the gas is used in calculating the log mean temperature.

Solving for the surface area, A, this method will give the size of the condenser that will cool the vapor to the dew point and condense it, although this method will not include the heat removal required for subcooling. The liquid must be subcooled to ensure that vapors will not readily volatilize from the condensate. Subcooling the liquid is accomplished by maintaining a liquid level which covers some of the bottom tubes in the condenser. This condition is called *flooding*.

The single phase (liquid) subcooling section can be treated separately, giving an area that is added onto the area needed for condensation. The heat balance and surface area are determined on the basis of cooling the liquid from its dew point temperature to the subcooled outlet temperature.

Different situations can occur when trying to condense vapors containing more than one component. In one case, all the components will condense at the coolant temperature. In another case, a few of the components may condense, while the other components will not condense as pure components but be very soluble in the condensate. Finally, there are mixtures where one or more of the components will not condense and are insoluble in the liquid condensate.

No matter which of these situations occur, features common to all three can affect condenser design. Features which must be considered are: multicomponent condensation is always **nonisothermal**; both liquid and vapor sensible heat removal must be provided for; and the composition of both phases is continually changing (Bell, 1972). This last feature is what complicates the design of multicomponent condensers. In multicomponent condensers, the liquid and gas phases are continually changing. This results in diffusional resistances occuring in both the liquid and vapor phases in addition to the resistance to heat transfer. In order to design the condenser, temperature and concentration profiles and diffusion coefficients must be evaluated at various points in the condenser. Methods for designing multicomponent condensers will not be discussed here, but are presented in Perry (1973) and in Bell (1972).

In nearly all condensation applications related to air pollution control, the gas to be treated is a mixture of condensable and noncondensable vapors. Noncondensable gases offer resistance to the rate of condensation in terms of both heat and mass transfer since the condensable vapor must diffuse to the cool surface. Air is usually the contaminant vapor but other gases that do not condense or dissolve in the liquid will also affect heat transfer. Unless specifically accounted for, the presence of noncondensable fluids will lower the values of the overall heat transfer coefficients that are listed in Table 6-1.

For preliminary rough estimates of condenser size, the procedure in Example 6-1 for single component condensation can be used for condensing multicomponents. In choosing a heat transfer coefficient, the smallest value should be used to allow as much overdesign as possible.

Example 6-1

In a rendering plant, tallow is obtained by removing the moisture from animal matter in a cooker. Exhaust gases from the cookers contain essentially steam, however the entrained vapors are highly odorous and must be controlled. Condensers are normally used to remove most of the moisture prior to incineration, scrubbing, or carbon adsorption.

The exhaust flow rate from the continuous rendering cooker is 20,000 acfm at 250°F. The exhaust gases are 95% moisture with the remaining portion consisting of air and obnoxious organic vapors. The exhaust stream is sent first to a shell-and-tube condenser to remove the moisture and then to a carbon adsorption unit. If the coolant water enters at 60°F and leaves at 120°F, estimate the required surface area of the condenser. The condenser is a horizontal, countercurrent flow system with the bottom few tubes flooded to provide subcooling.

Solution:

1. First, compute the pounds of steam condensed per minute.

   $20{,}000 \text{ acfm} \times 0.95 = 19{,}000 \text{ acfm steam}$

   From the ideal gas law:

   $PV = nRT$

   $$n = \frac{PV}{RT} = \frac{(1 \text{ atm})(19{,}000 \text{ acfm})}{\left(0.73 \dfrac{\text{atm} \cdot \text{ft}^3}{\text{lb mole} \cdot {}^\circ\text{R}}\right)(250 + 460\,{}^\circ\text{F})}$$

   $= 36.66 \text{ lb mole/min}$

   $$\dot{m} = \left(36.66 \frac{\text{lb mole}}{\text{min}}\right)\left(\frac{18 \text{ lb}}{\text{lb mole}}\right)$$

   $= 660 \text{ lb/min of steam to be condensed}$

2. Solve the heat balance to determine q for cooling the superheated steam and condensing only.

   $$q = \begin{matrix}\text{heat needed to} \\ \text{cool steam to} \\ \text{condensation temperature}\end{matrix} + \begin{matrix}\text{heat of} \\ \text{condensation}\end{matrix}$$

   $= \dot{m}\bar{C}_p \Delta T + \dot{m}H_v$

   The average specific heat ($\bar{C}_p$) of steam @ $250\,{}^\circ\text{F} \approx 0.45$ Btu/lb•°F. The heat of vaporization of steam @ $212\,{}^\circ\text{F} = 970.3$ Btu/lb. (Both these values were obtained from Perry, 1973).

   Substituting into the equation:

   $$q = \left(660 \frac{\text{lb}}{\text{min}}\right)\left(0.45 \frac{\text{Btu}}{\text{lb}\,{}^\circ\text{F}}\right)(250 - 212\,{}^\circ\text{F}) + \left(660 \frac{\text{lb}}{\text{min}}\right)\left(970.3 \frac{\text{Btu}}{\text{lb}}\right)$$

   $= 11{,}286 \text{ Btu/min} + 640{,}398 \text{ Btu/min}$

   $= 651{,}700 \text{ Btu/min}$

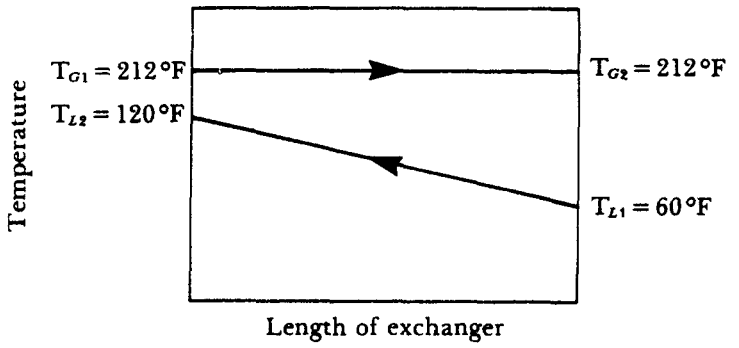
3. Now using Equation 6-4 to estimate surface area for this part of the condenser:

   $$A = \frac{q}{U \Delta T_{lm}}$$

   For a countercurrent condenser the log mean temperature is given by:

   $$\Delta T_{lm} = \frac{(T_{G1} - T_{L2}) - (T_{G2} - T_{L1})}{\ln \dfrac{T_{G1} - T_{L2}}{T_{G1} - T_{L1}}}$$

Remember that the desuperheater-condenser section is designed using the saturation temperature in calculating the log mean temperature difference.



$$\Delta T_{lm} = \frac{(212-120)-(212-60)}{\ln \dfrac{212-120}{212-60}}$$

$$= 119.5°F$$

The overall heat transfer coefficient, U is assumed to be 100 Btu/°F•ft²•hr from Table 6-1 for condensing steam. This value is chosen from the line in the table for stabilizer reflux vapors. Stabilizer reflux vapors are a mixture of hydrocarbon and steam similar to that exhausted from a rendering dryer. The median value of 100 is used as a best guess to approximate the conditions.

Substituting the appropriate values into Equation 6-4:

$$A = \frac{(651{,}700 \text{ Btu/min})\left(\dfrac{60 \text{ min}}{\text{hr}}\right)}{(100 \text{ Btu/°F•ft}^2\text{•hr})(119.5°F)}$$

$$= 3272 \text{ ft}^2$$

4. To estimate the total size of the condenser, we need to allow for subcooling of the water (212°F→160°F). (160°F is a safe margin).

   Refiguring the heat balance for cooling the water:

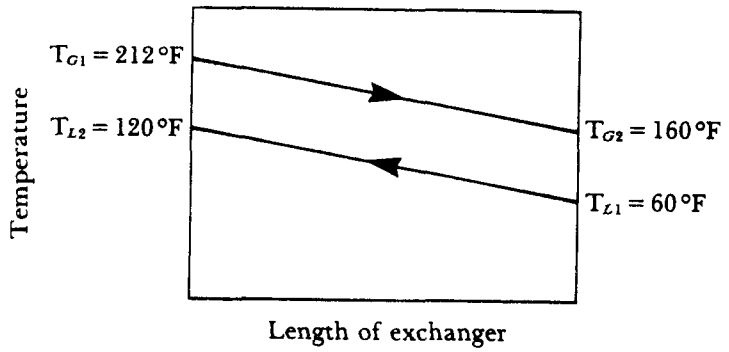
   $q = \dot{m}\overline{C}_p\Delta T$

   Where: $\dot{m} \approx 660$ lb/min (assuming all the steam is condensed)
   $\overline{C}_p$ for water = 1 Btu/lb•°F

   $$q = \left(660 \frac{\text{lb}}{\text{min}}\right)\left(1 \frac{\text{Btu}}{\text{lb•°F}}\right)(212-160)$$

   $$= 34{,}320 \text{ Btu/min}$$

The new log mean temperature is:



$$\Delta T_{lm} = \frac{(212-120)-(160-60)}{\dfrac{212-120}{160-60}}$$

$$= 96\,°F$$

$$A = \frac{q}{U\Delta T_{lm}}$$

U for cooling water with a water coolant is 200 Btu/°F•ft²•hr from the footnote at the bottom of Table 6-1.

$$A = \frac{34{,}320 \text{ Btu/min}}{(200 \text{ Btu/°F•ft}^2\text{•hr})(96\,°F)}$$

$$= 1.79 \text{ ft}^2 \text{ or } 2 \text{ ft}^2$$

5. The total area needed is:

$$A = 3272 + 2$$
$$= 3274 \text{ ft}^2$$

As illustrated by this example, the area for subcooling is usually very small compared to the area required for condensing.

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Chapter 7

Control of Nitrogen Oxide Emissions

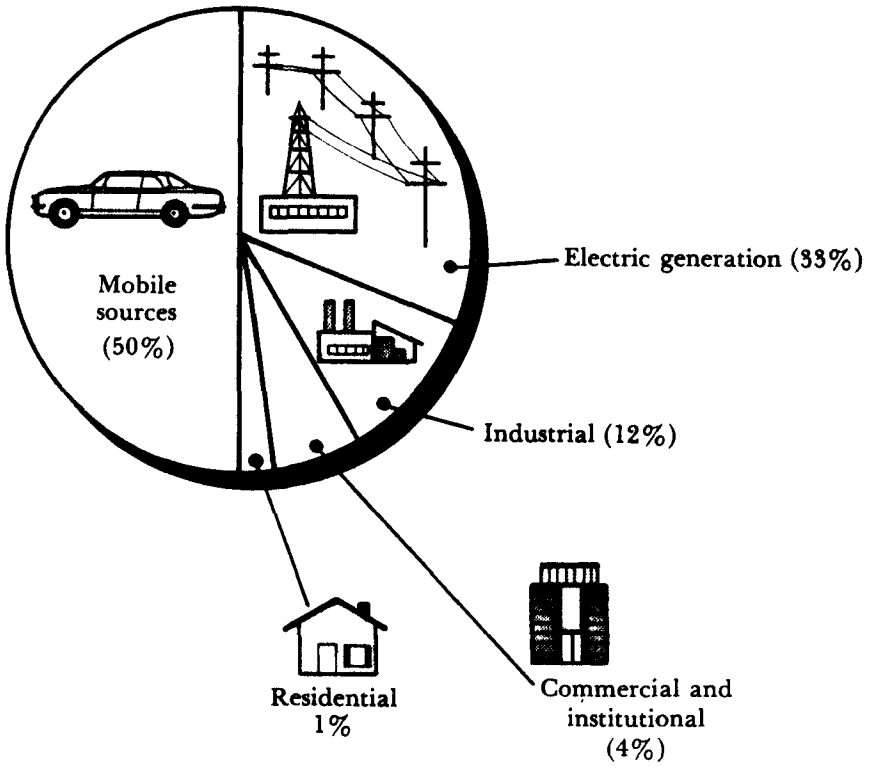
Introduction

More than 20 million metric tons of nitrogen oxides (NO_x) are emitted into the atmosphere each year as a result of burning fossil fuels (EPA, 1980). Fires and other natural occurrences emit 10 times the amount that anthropogenic nitrogen oxides emit, and tend to be dispersed over very large areas. Nitrogen oxides emitted from combustion processes can create local ambient levels that are 10 to 100 times greater than natural concentrations (EPA, August 1980). Controlling nitrogen oxides is important because of their harmful effects on public health and welfare, and their involvement in the formation of photochemical smog.

Sources of Nitrogen Oxides

Approximately one half of the total nitrogen oxide emissions come from motor vehicles. The other half comes from fossil fuel combustion by utility and industrial boilers and other industrial furnaces and processes (Figure 7-1). EPA has estimated that stationary source nitrogen oxide emissions will grow to 70% of the total by 1985 (EPA, August 1980). Part of this growth is due to the increased use of coal for generating electricity. There is more nitrogen in coal than in most other fossil fuels. Nitrogen oxides are released as coal is burned.

Of the stationary sources producing nitrogen oxides, utility and industrial boilers are the most significant; emitting approximately 60% of the total emissions (EPA, January 1978). Only about 2% of the NO_x emissions come from industrial processes such as nitric acid plants. The remaining stationary source NO_x emissions are from industrial, commercial, and residential heating. Emissions from nitric acid plants are usually reduced by control methods such as *wet scrubbing* or *catalytic reduction*. NO_x emissions from combustion sources are reduced by *combustion modifications, flue gas treatment*, or *fuel denitrification*. These reduction techniques will be discussed later in this chapter.



Source: EPA, February 1980

Figure 7-1. Manmade nitrogen oxide emissions.

Effects

NO_x emissions are known to cause human health problems and environmental damage. Exposure to nitrogen oxides is believed to increase the risks of acute respiratory disease and susceptibility to chronic respiratory infection. Nitrogen dioxide (NO_2) contributes to heart, lung, liver and kidney damage. Exposure to high NO_2 concentrations can be fatal. At lower NO_2 levels, 25 to 100 parts per million, acute bronchitis and pneumonia can occur. Eye and skin irritations can result from occasional exposure to low concentrations of NO_2 (EPA, August 1980). Nitrogen oxides are also toxic to vegetation. Reduced growth and plant destruction have been observed as a result of exposure to NO_x emissions.

Probably the most important effect of NO_x is the contribution to photochemical smog. NO_x reacts with hydrocarbons in the presence of sunlight to form photochemical oxidants, or ozone (O_3). Ozone can cause severe damage to the lungs and can aggravate asthma and other respiratory diseases. In addition to ozone, other smog pollutants such as peroxyacetylnitrate (PAN) can cause coughing, eye irritation, headaches, and throat irritation.

Nitrogen oxides also contribute to the global acid rain problem. Through a series of complex reactions, nitrogen oxides can be converted to nitric acid and nitrates which may be deposited in the form of rain and snow. EPA has estimated that nearly one-half of the acidity in acid rain is due to nitric acid (EPA, August 1980).

New Source Performance Standards

EPA has promulgated *New Source Performance Standards* (NSPS) for *fossil-fuel-fired steam generators* (FFFSG) with heat input greater than 250×10^6 Btu/hr (73 MW thermal). These standards establish nitrogen oxide emission limits for utility boilers. There are two standards: one for steam generators installed after August 17, 1971; and one for steam generators installed after September 18, 1978 (Table 7-1). These standards were set to reduce the NO_x emissions from the largest combustion source, utility boilers. Standards for industrial boilers are forthcoming in the near future.

Table 7-1. New Source Performance Standards for NO_x emissions from fossil-fuel-fired steam generators rated at greater than 250×10^6 Btu/hr or 73 MW (thermal) heat input.

NO_x emissions	Metric units (ng/J)	English units (lb/10^6 Btu)	Fuel
Subpart D; new sources after August 17, 1971	86	0.2	Gaseous
	130	0.3	Liquid
	300	0.7	Solid (except lignite)
Subpart Da; new sources built after September 18, 1978	86	0.2	Gaseous (except coal derived)
	130	0.3	Liquid (except coal derived)
	210	0.5	Subbituminous coal
	260	0.6	Bituminous/anthracite coal and lignite
	340	0.8	Lignite mined in ND, SD, and Montana burned in a slag-top furnace

Research Activities

The EPA's Office of Research and Development (ORD) is currently developing inexpensive methods to reduce NO_x emissions from combustion sources. Most recent studies for NO_x control technology are being done by the EPA's Industrial Environmental Research Laboratory in Research Triangle Park, NC (IERL-RTP). Additional research is being conducted at the Environmental Sciences Research Laboratory in Research Triangle Park, NC (ESRL-RTP) and through the Headquarters Office of Environmental Engineering and Technology in Washington, DC (EPA, August 1980).

The most promising control methods for reducing NO_x emissions from combustion sources are:

Before burning:
- fuel denitrogenation (denitrification)

During burning:
- combustion modifications
- catalytic combustion

After burning:
- flue gas treatment
- catalytic emission control

Fuel Denitrogenation

One control technology for reducing NO_x emissions is to remove the nitrogen contained in the fuel. Amounts of nitrogen vary in a fossil fuel. Coal, shale, and residual fuel oil contain a larger amount of nitrogen than either distillate oil or natural gas. The nitrogen in the fuel can be emitted as NO_x when the fossil fuel is burned in the furnace.

Nitrogen is removed from coal, shale, or heavy fuel oils by liquifying the fuels and mixing with hydrogen gas. The mixture is heated and a catalyst is used to cause the nitrogen in the fuel and the hydrogen to unite. This reaction produces two products; ammonia and a cleaner fuel. Researchers are developing better catalysts and finding ways to reduce the deposition of carbon on the catalyst surface (EPA, August 1980). Carbon deposits reduce the effectiveness of the catalyst life (see Chapter 3). This technology can reduce the nitrogen content in both natural fuels and synthetic liquid or gaseous fuels (made from shale and coal). This could become an increasingly important technology with the development and use of synthetic fuels in the future.

Formation of Nitrogen Oxides in Combustion Sources

When fossil fuels are burned with air in a furnace, some of the oxygen (O_2) and nitrogen (N_2) present combine to form oxides of nitrogen. Most of the oxides form according to the following reaction:

$$N_2 + O_2 \rightleftharpoons 2NO$$

Once the NO forms, the rate of decomposition is very slow and NO does not dissociate into N_2 and O_2 in any appreciable amounts.

The NO formed can react with more oxygen to form NO_2.

$$NO + \tfrac{1}{2} O_2 \rightleftharpoons NO_2$$

In large combustion furnaces, the majority of NO_x formed, approximately 95%, is in the form of NO. The main factors involved in NO_x formation are: **the flame temperature; the length of time combustion gases are maintained at that temperature;** and **the amount of excess air present in the flame.** Flame temperature in a utility boiler furnace is approximately 1650°C (3000°F). At this high temperature NO is formed in great abundance (sometimes greater than 1000 ppm). The residence time usually available in combustion equipment is too short for an appreciable fraction (usually less than 5%) of the NO to be oxidized to NO_2. The bulk of the NO_2 is formed in the atmosphere after release from the stack rather than in the confines of the combustion furnace. In this manual, NO_x refers to the total NO and NO_2 being emitted from major combustion sources.

In small residential heating units, furnace temperatures are usually less than 1090°C (2000°F). These relatively low temperature-operating units produce an exhaust gas containing low NO_x emissions; usually less than 10 ppm (Public Health Service, 1970).

Thermal NO_x and Fuel NO_x

Nitrogen oxide emissions are formed by two chemical processes occurring during combustion: the thermal NO_x process and the fuel NO_x process.

Thermal NO_x results from intense heat during combustion causing the nitrogen content of the combustion air to be oxidized. Its rate of formation is highly sensitive to the flame temperature and to a lesser extent to the local concentration of oxygen at the flame. Virtually all thermal NO_x is formed in the region of the flame which is at the highest temperature (EPA, January 1978). In theory, the formation of thermal NO_x can be reduced in four ways: reduce the nitrogen level (of air) at peak temperature; reduce the oxygen level at peak temperature; reduce the peak flame temperature; and reduce the time of exposure at peak temperature (EPA, January 1978). Since the concentration of nitrogen in air-fuel mixtures is relatively fixed, this tactic is not really applicable. Therefore, thermal NO_x is reduced in **field practice** by reducing oxygen levels, peak flame temperatures, and residence time in the NO_x producing section of the furnace. This is accomplished by various combustion modification techniques such as the use of *low excess air, staged combustion, reduced air preheat,* or *flue gas recirculation.*

Fuel NO_x occurs as the nitrogen contained in the fuel is oxidized. Therefore, it is dependent on the nitrogen content of the fuel. In fuels such as coal and heavy oil that are high in nitrogen content, approximately 20 to 60% of the fuel-bound nitrogen is oxidized (EPA, January 1978). Its rate of formation is strongly affected by the local oxygen concentration present in the flame and also by the mixing rate of the fuel and air. Thus, like thermal NO_x, fuel NO_x is dominated by the local combustion conditions. One way to reduce fuel NO_x emissions is to reduce the nitrogen content in the fuel. This is not always possible and therefore combustion modification techniques are used to reduce NO_x emissions. These include combining the use of low excess air firing, optimum burner designs, two stage combustion or high air preheat (secondary air preheat).

Combustion Modifications

As previously discussed, nitrogen oxide emissions result from operating conditions in the furnace, the amount of nitrogen in the combustion air, and the amount of nitrogen in the fuel. Combustion conditions in the furnace can be modified to reduce NO_x emissions. Some of the more widely used combustion modification techniques will be discussed in this section.

A typical furnace with boiler tube sections is shown in Figure 7-2. The main sections are the *fire wall* or *water wall tubes, superheater, convection tubes, economizer,* and *air preheater.* These boiler sections transfer heat in the furnace to produce steam for generation of electricity or process steam in an industrial plant. The furnace is surrounded by water walls or fire wall tubes. Water fed to the tubes is heated and is turned into steam. Steam is collected in steam drums located in the main convection section of the furnace. Convection tubes are used to convert any water collected in the steam drum back into steam. The superheater is a section of tubes used to superheat the steam to a higher temperature prior to going to the turbine. The economizer is used to heat the feedwater makeup to the boiler. The air preheater is used to preheat the air used for combustion. The furnace is designed to use as much of the heat as possible before exhausting flue gas out the stack.

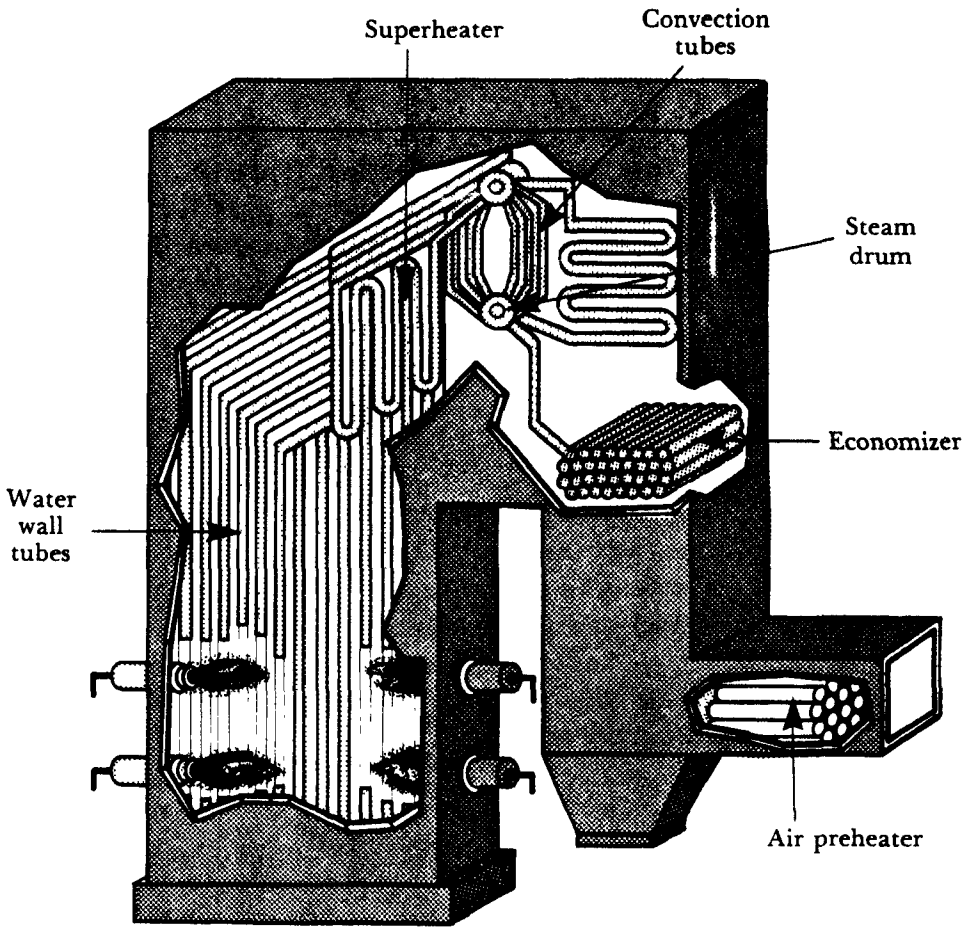


Figure 7-2. Typical boiler.

Low Excess Air

In a combustion system, a certain amount of excess air is required to ensure complete combustion of the fuel (Chapter 3). The more efficient the burners are for air and fuel mixing, the less amount of excess air required for complete combustion (Figure 7-3). The minimum amount of excess air is limited by the production of smoke and unburned fuel leaving the furnace.

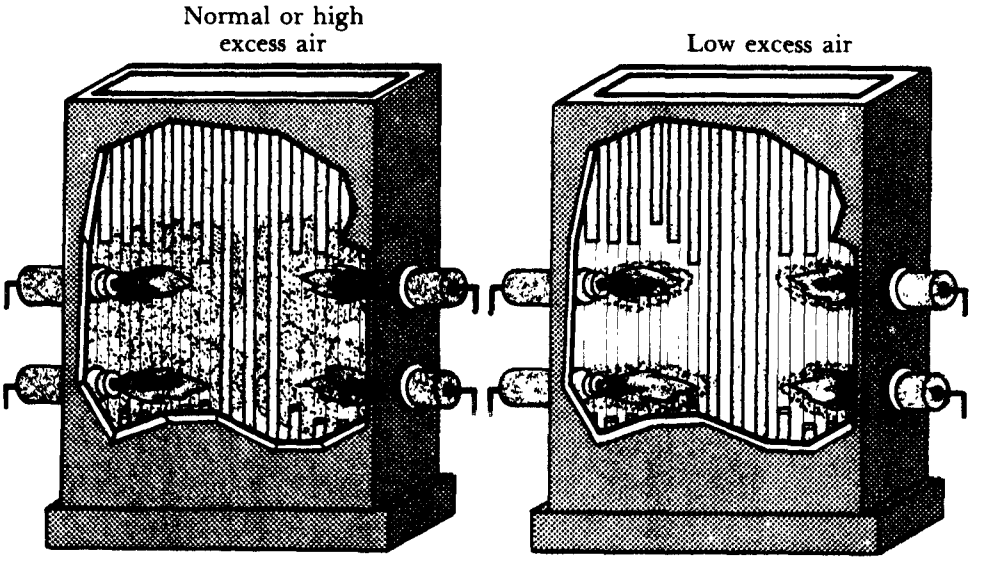


Figure 7-3. Excess air.

The level of excess air in an industrial or utility boiler will usually range from 5% to as high as 50 or 100%. However, most large boilers today operate with excess air less than 5%. Small residential heating furnaces, because of their unsophisticated design, usually operate with approximately 80 to 100% excess air. Gas turbines operate at very high excess air conditions, 300 to 400%. Most of this excess air is added as secondary air at lower temperatures resulting in low NO_x emissions.

NO_x emissions are reduced from many combustion furnaces by low excess air firing. The local flame zone concentration of oxygen is reduced, thus reducing both thermal and fuel NO_x. This method is easy to implement and actually increases the efficiency of the furnace slightly. However, there are problems with this combustion modification. Low excess air firing can produce smoke and high CO emissions. Also, fouling and slagging of boiler tube surfaces can occur if various coals and residual oils are burned.

NO_x reductions averaging between 16 and 20% are achieved on gas and oil fired utility boilers when the excess air is reduced to a level between 2 and 7% (EPA, January 1978). NO_x reductions averaging around 20% can be achieved on coal fired utility boilers if the excess air is reduced to the 20% level or lower (EPA, January 1978). For most utility boiler applications, operating at low excess air conditions is considered a routine procedure.

Staged Combustion

During staged combustion, air and fuel mixtures are combusted in two separate zones. In one zone, the fuel is fired with less than a stoichiometric amount of air. This creates a **fuel rich** local zone in the regions of the primary flame. The second zone is an **air rich** zone where the remainder of the combustion air is introduced to complete the combustion of the fuel (Figure 7-4). The heat in the primary flame zone is not as intense as with normal firing because combustion is incomplete. The air mixed with the fuel is sub-stoichiometric in the NO_x forming region of the flame, thus creating a low NO_x condition. This modification is also referred to as *off-stoichiometric combustion.*

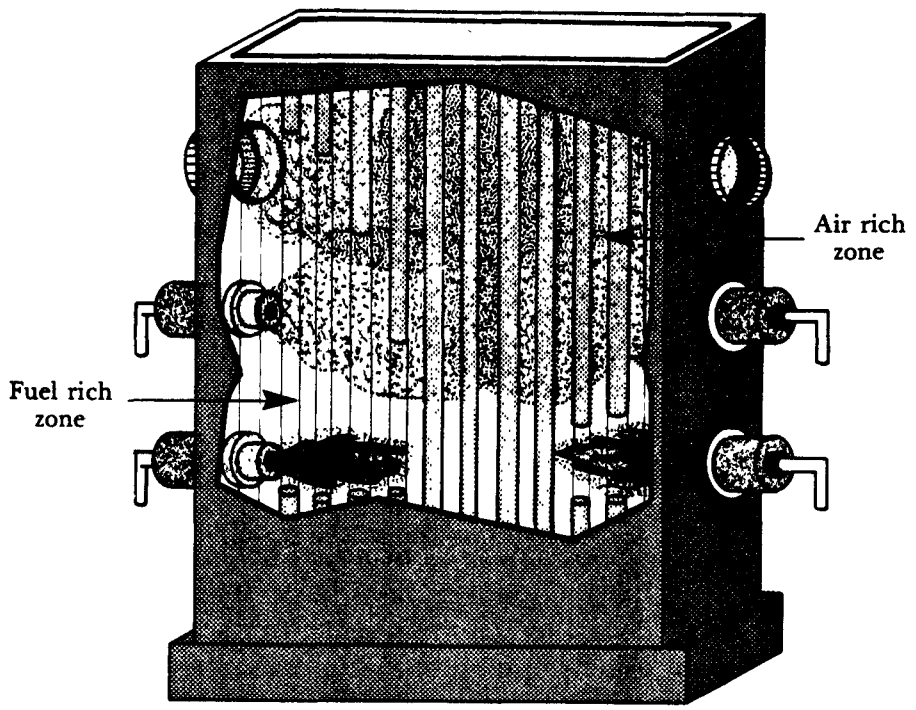


Figure 7-4. Staged combustion.

Staged combustion reduces NO_x emissions by a combination of several factors. First, a lack of available oxygen for NO_x formation in the fuel rich stage is due to off-stoichiometric firing. Second, the flame temperature may be lower in the first stage than with single stage combustion. Third, the peak temperature in the second stage (air rich) is lower.

Staged combustion is an effective technique for controlling both thermal and fuel NO_x due to its ability to control the mixing of fuel with combustion air. The NO_x reduction effectiveness depends on good burner operation to prevent convective boiler tube fouling, unburned hydrocarbon emissions, and poor ignition characteristics which occasionally occur at an excessively fuel rich boiler operation. Fire side boiler tube corrosion may occur when burning some coals or heavy oils under staged combustion conditions (EPA, September 1975).

In staged combustion the flame is long, yellow, and slightly smokey as opposed to the short and intense flame observed on normal firing. Fuel combustion (second stage) also extends further into the furnace, sometimes causing excessive temperatures in the convective and superheater sections of the boiler.

Staged combustion is accomplished by using modifications to the boiler such as *over fire air ports* (OFA), or *burners out of service* (BOOS). Over fire air ports are separate ports located above the burner. The burners are operated fuel rich with the remainder of the combustion air coming in through the over fire air ports (Figure 7-5). These are sometimes referred to as *NO_x ports*. In some boilers, a number of the burners are operated fuel rich, others air rich in a staggered configuration. This is called *biased firing*. In the case where some burners are operated on air only, this modification is called burners out of service (Figure 7-6).

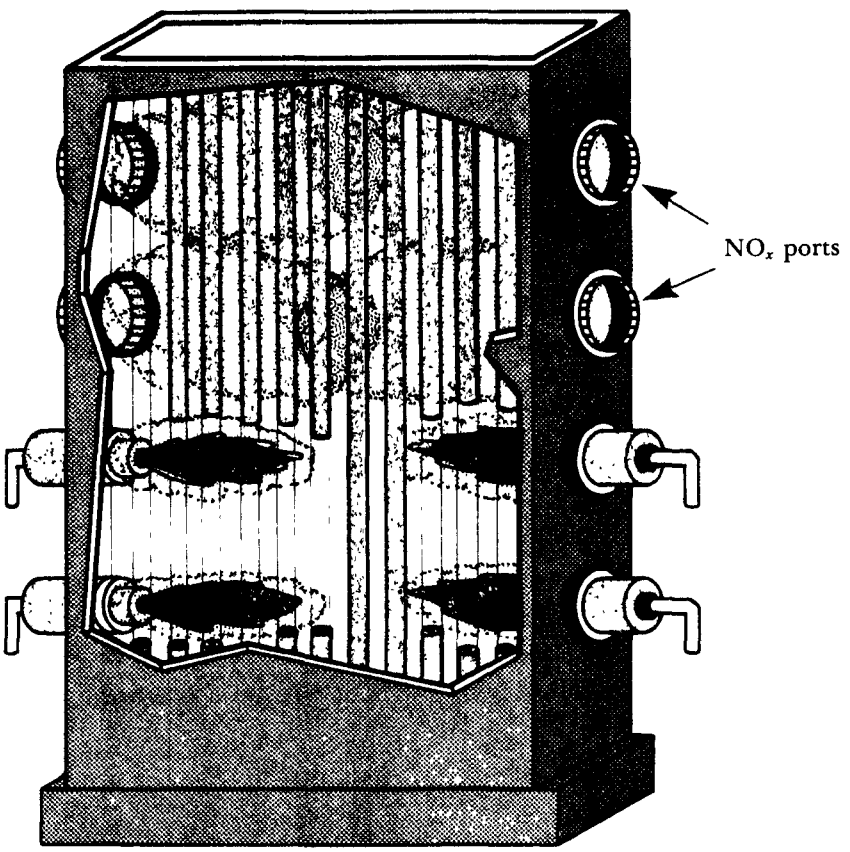


Figure 7-5. Over fire air.

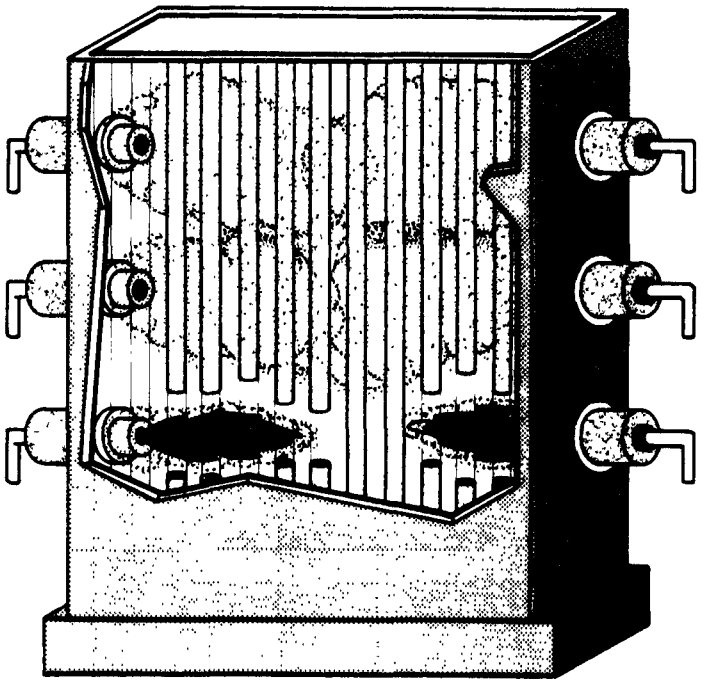


Figure 7-6. Burners out of service.

On existing boilers, a steam load reduction will result with burners out of service if the active fuel burners do not have the capacity to supply fuel for a full load. Most utility boilers installed since 1971 have been designed with over fire air ports so that all fuel burners are active during the staged combustion operation (EPA, January 1978). Using staged combustion modifications on oil and gas fired boilers reduces NO_x emissions by approximately 30 to 40% (EPA, 1978). Modifying existing coal boilers has reduced NO_x emissions 30 to 50% (EPA, August 1980).

Staged combustion can also be accomplished by careful control of air and fuel mixing in the burner. EPA (IERL) is working on the development of new coal burners based on a staged combustion design that may reduce NO_x as much as 85% (EPA, August 1980). These *low* NO_x burners will be discussed in more detail later in this chapter.

Flue Gas Recirculation

Flue gas recirculation (FGR) has been used to reduce thermal NO_x emissions from large oil and gas fired boilers. A portion (10 to 30%) of the flue gas exhaust is recycled back into the main combustion chamber by removing it from the stack breeching and mixing it with the secondary air windbox (Figure 7-7). In order for FGR to be effective in reducing NO_x emissions, the gas must enter directly into the combustion zone. This recirculated gas lowers the flame temperature and dilutes the oxygen content of the combustion air, thus lowering thermal NO_x emissions.

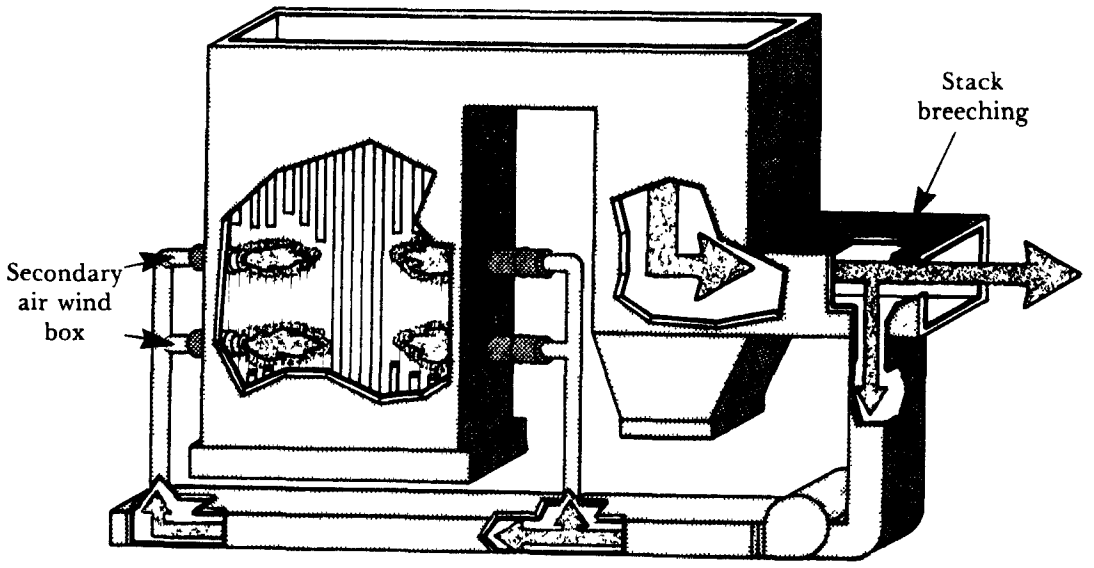


Figure 7-7. Flue gas recirculation.

Some operational problems can occur using flue gas recirculation. Possible flame instability, loss of heat exchanger efficiency; and for small package boilers, condensation on internal heat transfer surfaces limit the usefulness of gas recirculation (EPA, January 1978).

Flue gas recirculation requires greater capital expenditures than low excess air and staged combustion modifications. High temperature fans (forced or induced draft), ducts, and large space are required for recirculating the gas. However, the costs can be reasonable if the boiler is already equipped with gas recirculation for steam temperature control (EPA, September 1974).

NO_x reduction of approximately 40 to 50% is possible with recirculation of 20 to 30% of the exhaust gas in gas and oil fired boilers. Since FGR is used to reduce thermal NO_x, it is not effective for reducing NO_x emissions generated from burning a high nitrogen content fuel such as coal. At high rates of recirculation ($\geq 30\%$) the flame can become unstable, thus increasing carbon monoxide and hydrocarbon emissions.

The combustion modification technique (FGR) should not be confused with flue gas recirculation used by utilities to control boiler tube (superheat) temperature. In this utility practice, flue gas is taken from the stack breeching and mixed into the hopper bottom section of the boiler that is located below the combustion chamber. Since flue gas is not recirculated into the main combustion chamber, this utility practice is not useful in reducing NO_x emissions.

Low NO_x Burners

New low NO_x burners have been developed by several manufacturers to reduce NO_x emissions. Burners control mixing of fuel and air in a pattern to keep flame temperature low and dissipate the heat quickly. Some burners are designed to control the flame shape to minimize the reaction of nitrogen and oxygen at peak flame temperatures. Other designs have fuel rich and air rich regions to reduce flame temperature and oxygen availability.

The Dual Registered Low NO_x burner is shown in Figure 7-8. The mixture of pulverized coal and primary air is controlled to slightly delay combustion of the fuel. The remainder of the combustion air (secondary air) is introduced through two concentric air zones which surround the coal nozzle. The flame produced is elongated compared to the short intense flame produced in a conventional burner. The peak flame temperature is reduced, thus lowering thermal NO_x emissions; and the oxygen available in the flame is lower, thus reducing fuel NO_x emissions. Field tests of this burner have shown a 40 to 60% reduction in NO_x emissions (EPRI, 1979).

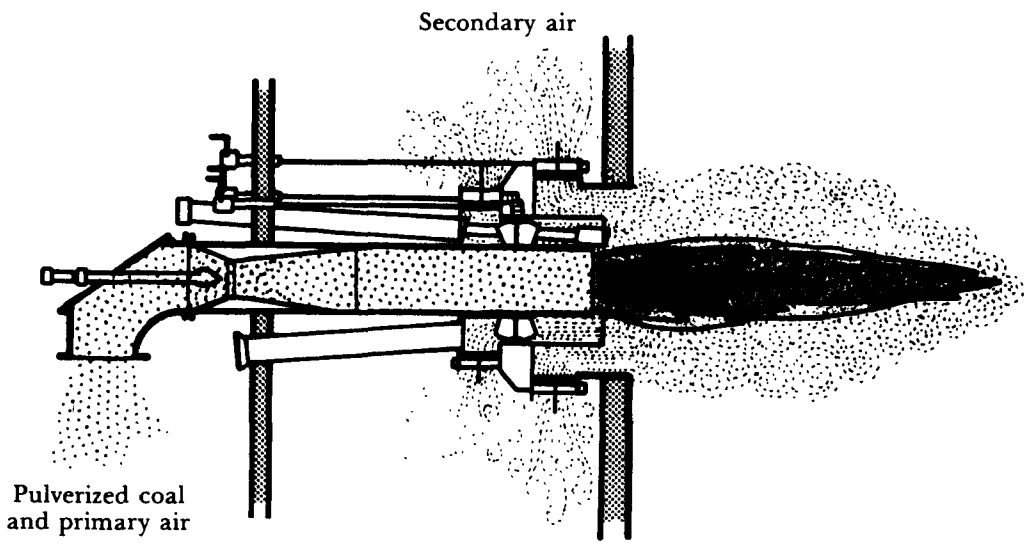


Figure 7-8. Schematic of a low NO_x burner.

EPA is developing a low NO_x burner that produces a fuel rich primary combustion zone and controls air and fuel mixing. Additional air is introduced from the periphery of the burner to complete combustion in a secondary zone. These conditions lead to a reducing atmosphere and a preferential conversion of the nitrogen in the coal to molecular nitrogen (N_2) emissions. This burner design may reduce NO_x emissions as much as 85% (EPA, August 1980).

Other Combustion Modification Techniques

Other combustion modifications can also reduce NO_x emissions from combustion sources. Some of these include *reduced air preheat, load reduction, steam and water injection,* and *catalytic combustion.*

Reduced air preheat and load reduction are used sparingly in large boilers due to the energy penalty involved and the relatively low emission reduction occurring. Steam and water injection are used mainly for NO_x emission reduction in gas turbines and internal combustion engines. Steam or water is injected into the combustion area to lower the peak flame temperature and thus reduce thermal NO_x emissions. In catalytic combustion, a catalyst is used to achieve oxidation of the fuel rather than using high flame temperatures. These systems have been used in gas turbines to reduce NO_x emissions well below 10 ppm (EPA, August 1980).

Flue Gas Treatment

Nitrogen oxide emissions can be reduced by treating the flue gas after it leaves the combustion zone. Flue gas treatment processes include *Exxon Thermal DeNO$_x$, Selective Catalytic Reduction* (SCR), *Shell UOP,* and *Wet Simultaneous NO$_x$ and SO$_2$ reduction.* These processes have been used in Japan to reduce NO_x emissions from utility boilers and pilot projects are currently being tested in the United States at various utilities. Full scale units are expected to be installed in the next few years.

Exxon Thermal DeNO$_x$ Process

A process using ammonia injection to reduce NO_x emissions has been developed by Exxon Research and Engineering Co. called Thermal DeNO$_x$. Ammonia is injected into the post combustion zone of the boiler (Figure 7-9). The ammonia reacts with NO_x (which is 95% NO) to reduce the oxides to molecular nitrogen and water:

(Eq. 7-1)

$$4NH_3 + 4NO + O_2 \rightarrow 4N_2 + 6H_2O$$

The above reaction is extremely temperature dependent. In a boiler, this reaction successfully takes place at approximately 950°C (1740°F). At higher temperatures (above 1090°C) the ammonia is oxidized, forming additional NO_x. At lower temperatures (below 850°C) the ammonia passes through the boiler unreacted (Lyon and Tenner, 1978). This temperature range can be altered somewhat with the addition of hydrogen. Hydrogen shifts the lower temperature range where the reaction occurs at approximately 700°C. The hydrogen does not widen the range, but just changes it. At H_2:NH_3 ratios of 2:1, the reaction will proceed rapidly at 700°C (Lyon and Tenner, 1978). By selecting various H_2:NH_3 ratios, the NO_x reduction can be accomplished at intermediate temperatures.

This process also depends on the ammonia and NO_x concentrations in the system. As can be seen from Equation 7-1, four moles of NH_3 are needed to reduce 4 moles of NO, or, a 1:1 mole ratio. Injection of NH_3:NO in 1 to 1.5 mole ratios will reduce NO_x emissions between 60 and 70%. Using a mole ratio of 0.5 will give NO_x reductions of approximately 40%.

Ammonia is mixed with a carrier gas of air or steam to provide good mixing of NH_3 with the flue gas. This ammonia mixture is injected at various points in the convection and superheater sections of the boiler (Figure 7-9). Multipoint injection grids are used (to inject ammonia) to compensate for varying temperatures in the convection and superheater sections. Temperature differences result from changing electric loads to the generator.

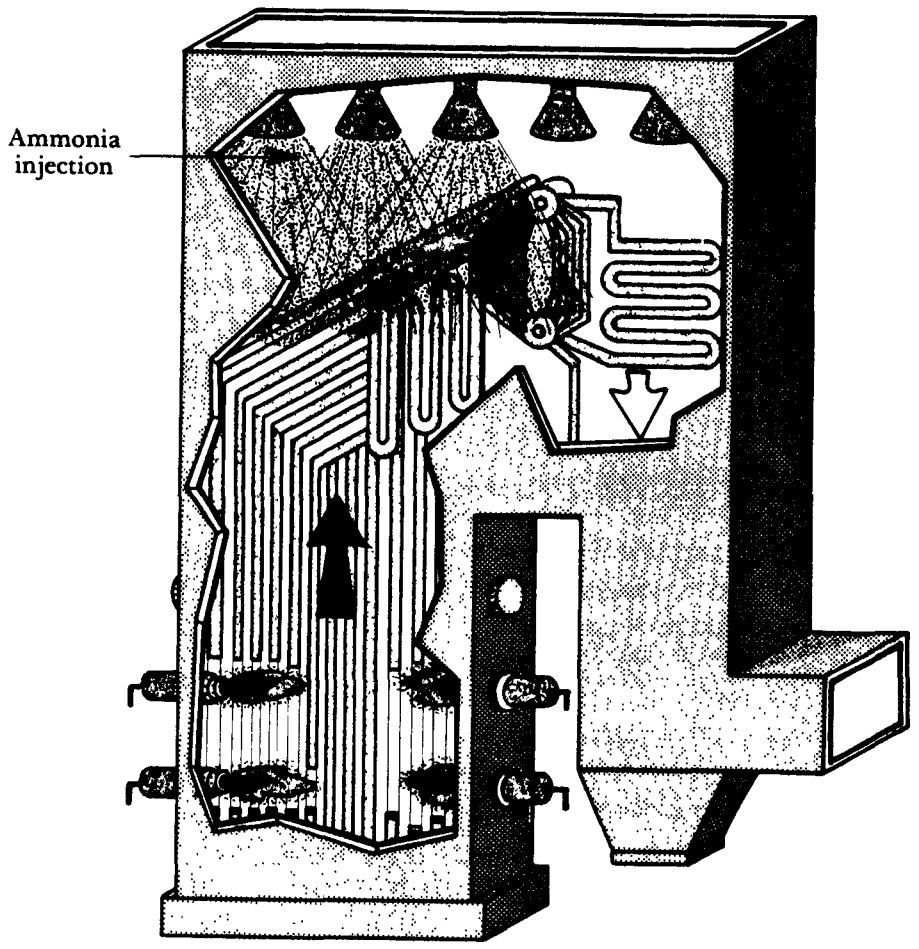


Figure 7-9. Exxon Thermal $DeNO_x$ process.

NO_x emissions in oil, gas, and small coal fired boilers can be reduced more than 60% by using ammonia injection. This process could be used to supplement the NO_x emission reductions accomplished by combustion modifications, when very low levels of NO_x are required. Strict NO_x emission regulations adopted by the California Air Resources Board and the South Coast Air Quality Management District of California require very low NO_x emissions from utility boilers. The Los Angeles Department of Water and Power is planning to install the Exxon Thermal $DeNO_x$ process on their Haynes No. 4 unit, a 227 MW generator. The capital investment to retrofit the Thermal $DeNO_x$ process is estimated to be approximately 18 to 25 $/kW. Operating and maintenance costs are estimated at 6 to 9 $/kW. Both cost estimates are on a 1979 cost basis (EPA, Winter 1979-1980).

Selective Catalytic Reduction

The Selective Catalytic Reduction (SCR) is a dry process used to reduce NO_x emissions from fossil-fuel-fired boilers. The SCR process has been used extensively in Japan to achieve a 90% reduction in NO_x emissions (Mobley, 1979). This process is based on the preferential reaction of NH_3 with NO_x rather than with SO_2 in the flue gas. The reactions are expressed as:

(Eq. 7-2) $$4NH_3 + 4NO + O_2 \rightarrow 4N_2 + 6H_2O$$

and

(Eq. 7-3) $$4NH_3 + 2NO_2 + O_2 \rightarrow 3N_2 + 6H_2O$$

Equation 7-2 represents the predominant reaction occurring since 95% of NO_x emissions in combustion flue gas are in the form of NO.

This process involves injecting NH_3 into the flue gas and passing this mixture through a catalytic reactor (Figure 7-10). NO_x emissions are reduced to harmless molecular nitrogen (N_2) and water vapor (H_2O). Ammonia is injected on an NH_3:NO mole ratio of 1:1 attaining a 90% NO_x emission reduction with less than 20 ppm NH_3 leaving the reactor (Mobley, 1979).

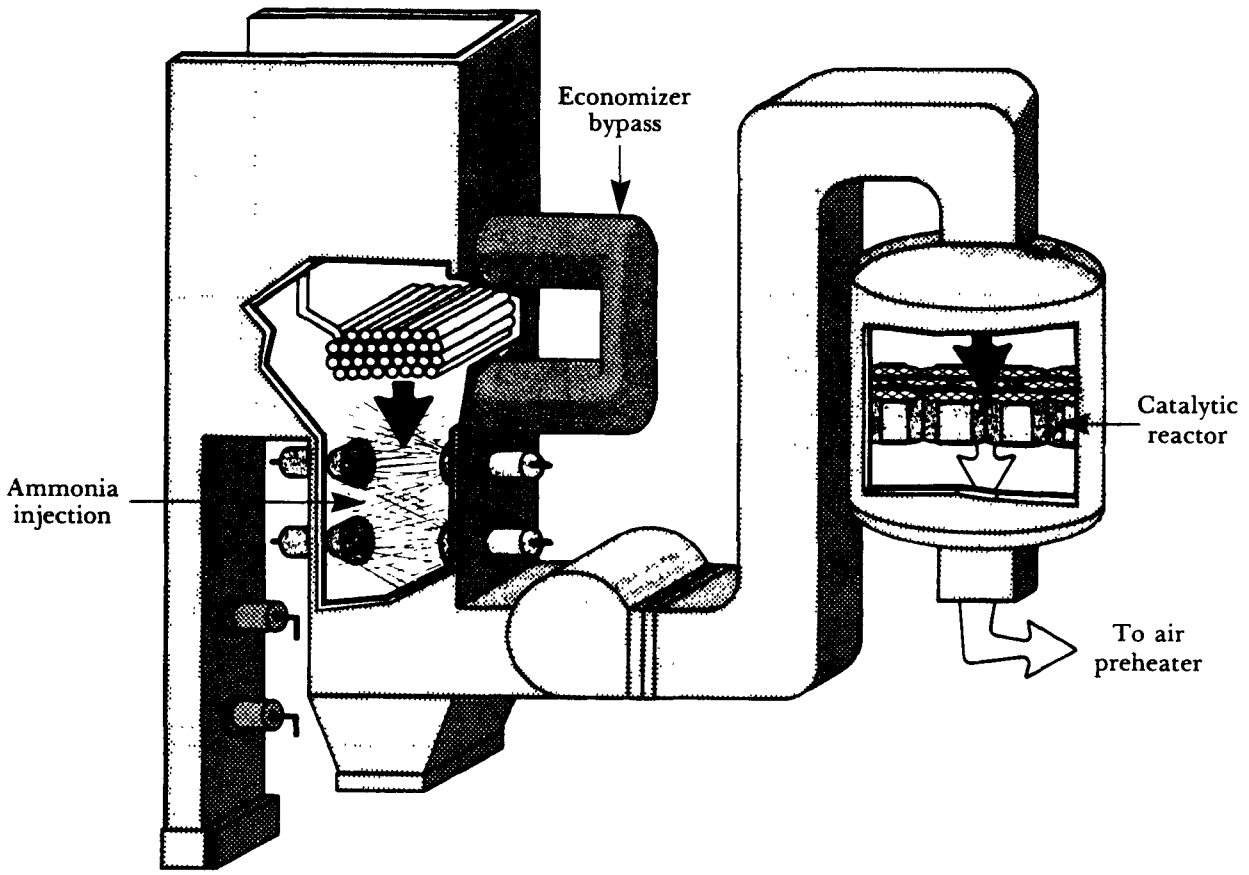


Figure 7-10. Selective Catalytic Reduction (SCR) process.

The optimum temperature for successful NO_x reduction in the catalytic reactor is between 300 and 400°C. The reactor is usually located between the boiler economizer and air preheater. A bypass around the economizer is used when temperatures begin to fall below 300°C. The reactor can be located before or after the baghouse or electrostatic precipitator used to collect particulate matter.

A number of materials have been used for catalysts. Initially, a platinum metal on an alumina (Al_2O_3) support was used for NO_x control on gas fired boilers. Sulfur oxides, particularly SO_3 and SO_2, poison the alumina. Other catalysts must be used which resist SO_x deterioration when burning fuel containing sulfur. Titanium dioxide (TiO_2) and vanadium (V_2O_5) catalysts are resistant to SO_x attack. Therefore, most catalysts on oil and coal fired boilers contain TiO_2 or V_2O_5 (Mobley, 1980). Other active metals including C, CO, Cr, Fe, Mn, Mo, Ni and W have been used as catalysts. The oxides and sulfates of these metals have also been

used as catalysts (Mobley, 1979). The exact compositions and constituents of most catalysts are proprietary information.

Catalyst shape and reactor design vary, depending on the manufacturers' design and the boiler application. Natural gas fired boilers use catalysts made of spherical pellets, rings or cylinders that are packed in a *fixed bed reactor* (Figure 7-11) which is similar to the packed tower described in Chapter 4. Since high concentrations of particulate matter will plug a fixed bed reactor, a *moving bed reactor* or a *parallel reactor* is used on oil or coal fired boilers. In a moving bed reactor, the catalyst is charged at the top of the reactor and moves down through the vessel while the flue gas passes through the catalyst in a cross flow motion (Figure 7-12). The spent catalyst is discharged at the bottom of the reactor. Particulate matter is then removed by screening, and the catalyst is regenerated before being returned to the reactor. Parallel flow reactors (Figure 7-13) were designed for use on coal fired boiler flue gas containing a high particulate matter concentration. The flue gas flow is parallel to the catalyst surface. Parallel catalyst shapes are either tubular, honeycomb, or plates (Figure 7-14). Catalysts are usually made into unit cells that are 12 inches square. These unit cells are stacked into the reactor to form the grids or layers of the catalyst (Figure 7-15).

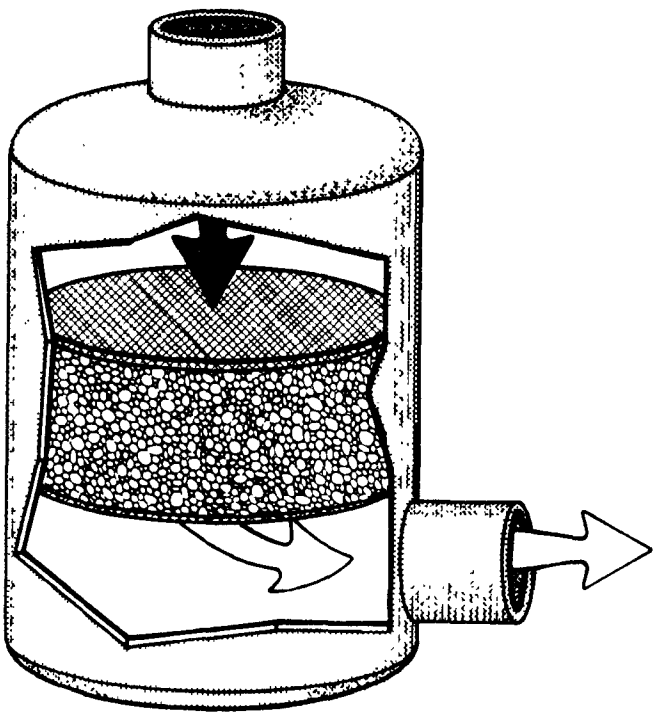


Figure 7-11. Fixed bed reactor.

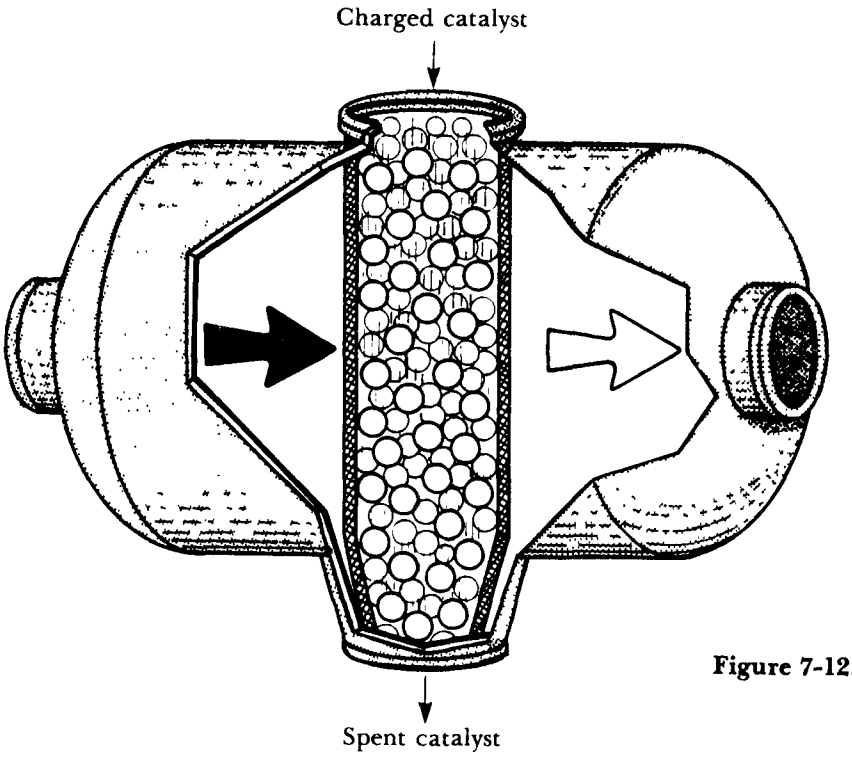


Figure 7-12. Mobile bed reactor.

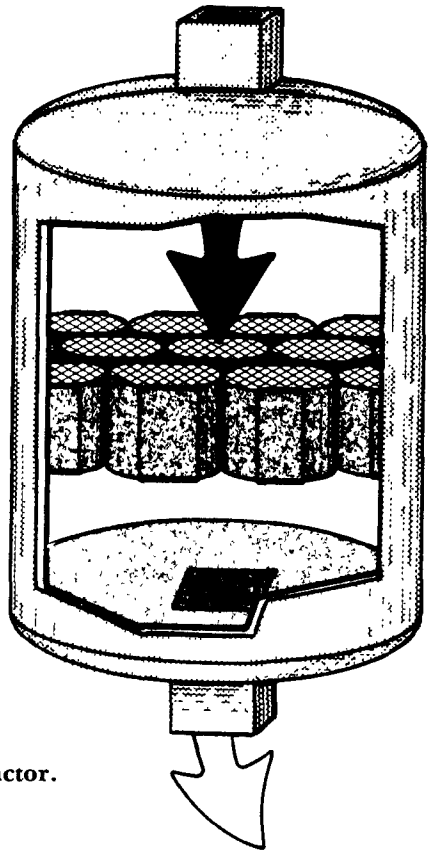


Figure 7-13. Parallel flow reactor.

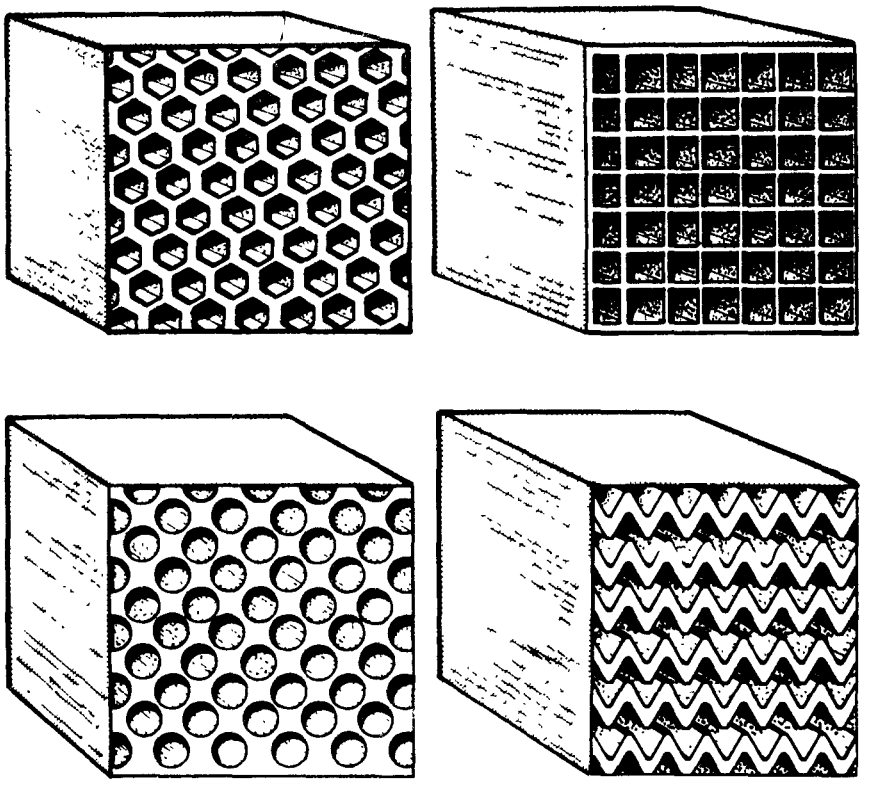


Figure 7-14. Parallel flow catalysts.

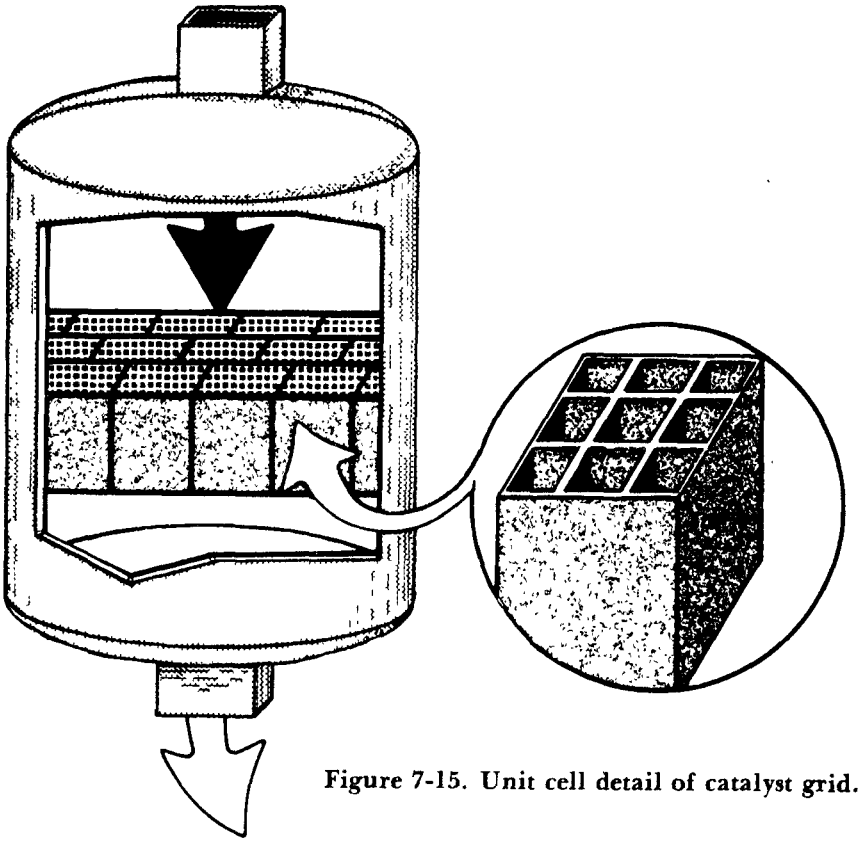


Figure 7-15. Unit cell detail of catalyst grid.

The biggest problem with the SCR process is the formation of solid ammonium sulfates $(NH_4)_2SO_4$ and liquid ammonium bisulfate (NH_4HSO_4) (Mobley, 1980). This problem occurs when using the SCR process on high sulfur oil and coal fired boilers. Ammonia reacts with SO_3 to form corrosive compounds of $(NH_4)_2SO_4$ and NH_4HSO_4 that coat the air preheaters. Increased soot blowing in the furnace and water washing of the air preheaters helps remove these materials.

In Japan, the SCR process has been used to reduce NO_x on over fifty boilers or furnaces burning gas or oil and is being installed on four coal fired utility boilers. Expected startup of these facilities is mid to late 1981. In the U.S., a pilot plant (0.5 MW) SCR process sponsored by the EPA has been installed at the Georgia Power Company. The Southern California Edison Company is installing the SCR process on an oil fired 100 MW unit at their Huntington Beach Station. A NO_x reduction of 90% is expected in order to meet California Air Resources Board and South Coast Air Quality Management District regulations.

The cost of the SCR process is estimated on a basis of a new 500 MW boiler firing eastern bituminous coal with a heating value of 10,500 Btu/lb, a sulfur content of 3.5% and an ash content of 16%. The capital cost for 90% NO_x reduction (in 1979) is 42 $/kW and the annualized cost is 2.2 mills/kWh (Mobley, 1980).

Shell UOP Process

Shell UOP is a dry process that simultaneously removes both NO_x and SO_x emissions. This process can also be designed to remove either compound separately (Pohlenz, 1979).

The process uses a copper oxide (CuO) catalyst supported on alumina. These catalysts are located in two or more parallel passage reactors as shown in Figure 7-16. Flue gas containing both NO_x and SO_x is introduced into the reactor where the SO_x reacts with copper oxide to form copper sulfate $(CuSO_4)$. At the same time ammonia is being injected which reacts with the NO_x. The copper sulfate, and to a lesser extent the copper oxide, act as catalysts for the NO_x-NH_3 reaction.

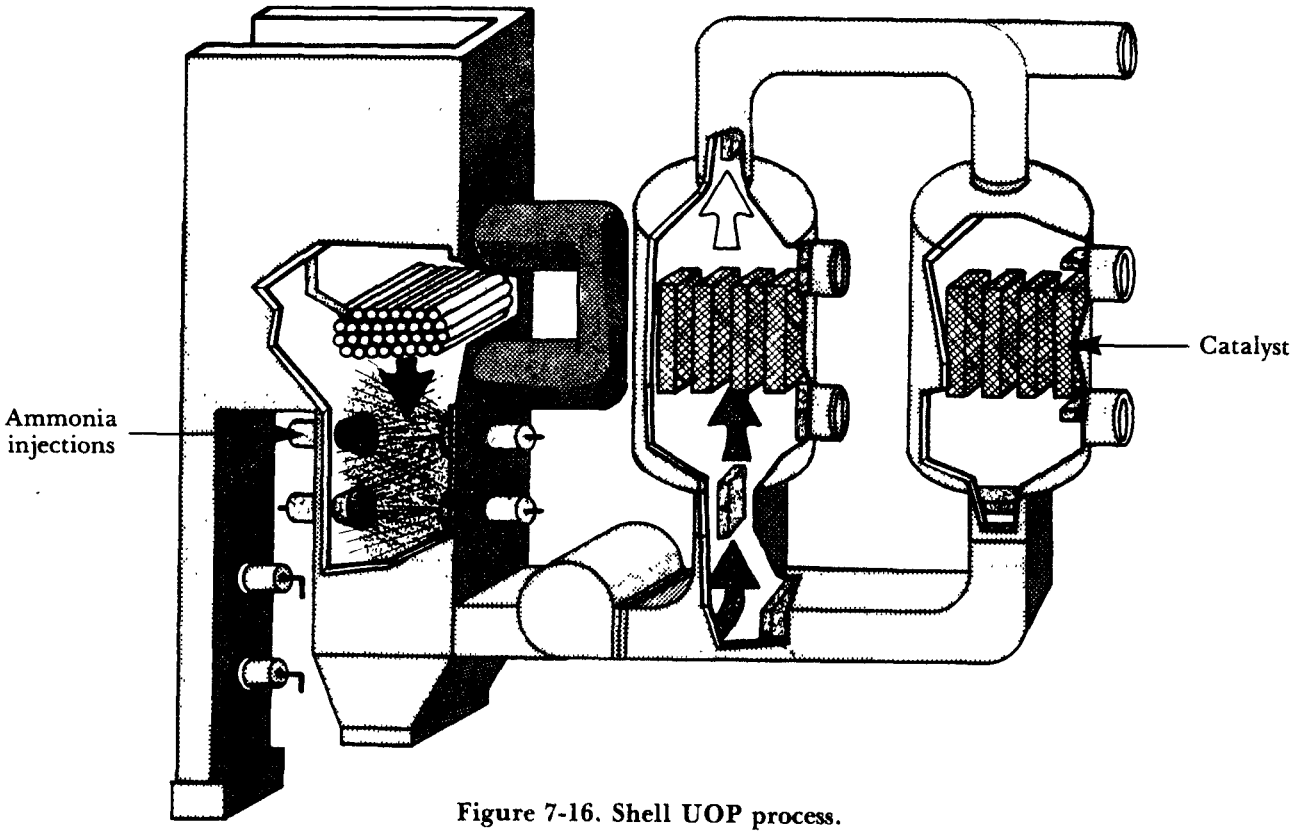


Figure 7-16. Shell UOP process.

The following reactions occur in the reactor:

(Eq. 7-4) $$CuO + \tfrac{1}{2}O_2 + SO_2 \rightarrow CuSO_4$$

(Eq. 7-5) $$4NO + 4NH_3 + O_2 \xrightarrow{CuSO_4} 4N_2 + 6H_2O$$

SO_2 NO_x reduction

When the reactor catalyst is saturated with $CuSO_4$, the flue gas is redirected to a fresh reactor and the spent catalyst is regenerated (Figure 7-17). Hydrogen is used to regenerate the catalyst by reducing the $CuSO_4$ to copper and producing a concentrated SO_2 gas stream. The SO_2 gas is then used to produce sulfuric acid or elemental sulfur for commercial sale. The copper in the reactor is oxidized to CuO and the process is ready to be put on line again. The reactions that take place in the reactor during catalyst regeneration are:

(Eq. 7-6) $$CuSO_4 + 2H_2 \rightarrow Cu + SO_2 + 2H_2O$$

(Eq. 7-7) $$Cu + \tfrac{1}{2}O_2 \rightarrow CuO$$

} Catalyst regeneration

The Shell UOP process can be operated as a NO_x emission reduction process by eliminating the regeneration cycle. The process can be operated as a SO_x emission reduction process by eliminating the ammonia injection.

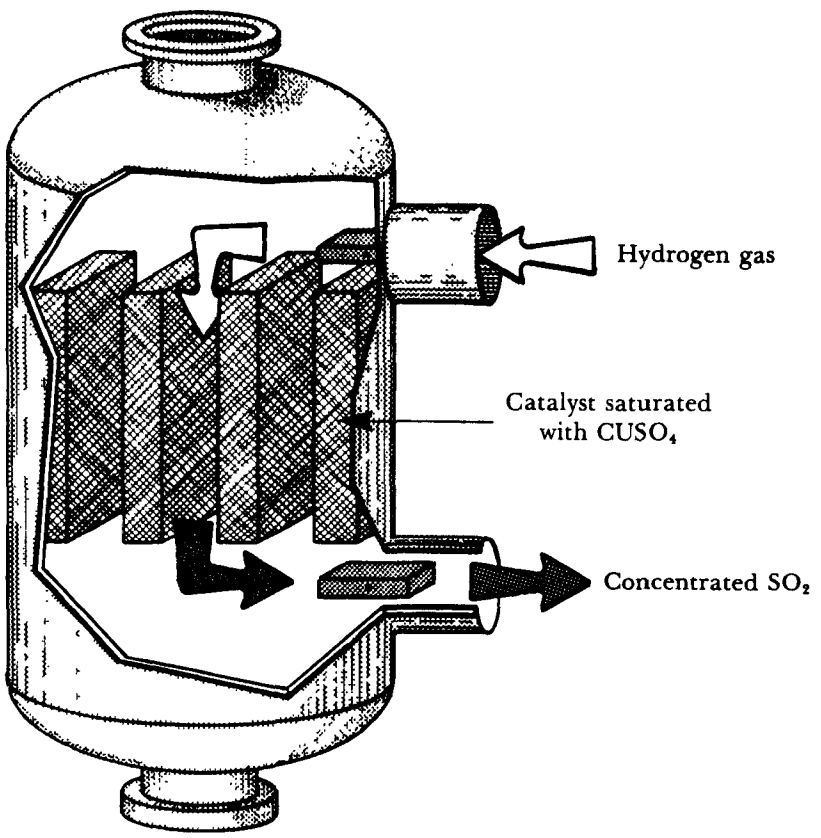


Figure 7-17. Catalyst regeneration (Shell UOP).

This system can be designed to use moving packed beds, fixed packed beds, or fluid systems. Shell UOP Inc. has developed a type of fixed bed reactor in which the flue gas flows through open channels called a parallel passage reactor (Figure 7-18). The catalyst (CuO) is placed between thin layers of wire gauze. Flue gas flows along the catalyst surfaces and the NO_x and SO_x gas diffuses into the catalyst. This design reduces reactor plugging by particulate matter. The NO_x/SO_x emission reduction is achieved in less than one-half second (Pohlenz, 1979).

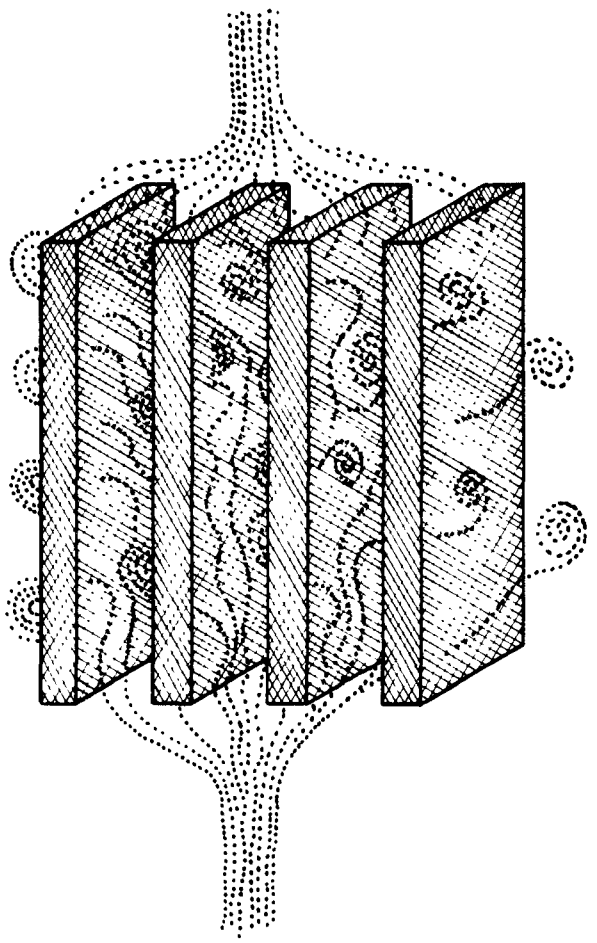


Figure 7-18. Detail of catalyst (Shell UOP).

The reactors are operated at temperatures between 350 and 450°C during the NO_x reduction cycle. The pressure drop through the parallel passage reactor is approximately 75 to 150 mm H_2O depending on the gas flow rate and NO_x reduction desired. The ammonia injection rate is one mole of NH_3 injected for one mole of NO_x reduced.

The Shell UOP process has been operated on large boilers in Japan for both SO_2 and NO_x emission reduction. Units have operated to remove at least 90% SO_2 and 70% NO_x emissions (Mobley, 1980). EPA has sponsored a pilot plant demonstration unit at the Big Bend Station of the Tampa Electric Company in North Ruskin, Florida. This demonstration unit (0.5 MW) is installed on a coal burning boiler and is designed for 90% SO_2 and 90% NO_x removal efficiencies. The cost estimates for a 500 MW utility boiler burning 3.5% sulfur and 16% ash coal are given on a 1979 basis. The capital cost is 134 $/kW for SO_2 and NO_x removal and the annualized cost is 6.4 mills/kWh (Mobley, 1980).

Wet NO_x and SO_x Processes

Wet processes for the simultaneous reduction of NO_x and SO_x emissions have been developed and installed on large oil fired boilers in Japan. These processes use absorbers to reduce both SO_2 and NO_x emissions. These processes are not cost competitive compared to other flue gas treatment processes used in the United States at this time. The estimated capital cost is 200 $/kW ($SO_2$ and NO_x) and annualized cost is 11.3 mills/kWh (Mobley, 1980).

Other NO_x Reduction Techniques

Glass Manufacturing

A project to reduce NO_x emissions during the manufacture of glass is being conducted by ORD's Industrial Environmental Research Laboratory in Cincinnati. In conventional glassmaking processes, the silicon and other materials for producing glass are heated in a furnace. Over 50% of the energy used to heat the furnace is lost in exhaust gases. Experiments are under way to alter the furnace design so that the exhaust gas is used to preheat the raw materials for glassmaking. Preheating has a number of advantages in the glassmaking industry. It reduces the amount of fuel needed in the furnace, the amount of air drawn into the combustion process, the temperature required in the furnace to melt the silicon material, and the length of time that the material must remain in the furnace. Reductions in fuel use, air intake, and combustion temperature have resulted in the decrease of both fuel-bound NO_x emissions, and thermal NO_x emissions. The costs that the glass industry could save by using less fuel are an additional incentive to make use of the preheating process (EPA, February 1980).

Nitric Acid Production

About 70% of the nitric acid (HNO_3) produced in the U.S. is used to manufacture fertilizer. Other uses include the production of industrial explosives, separating gold and silver, pickling steel and brass, and photoengraving. Nitric acid is produced by oxidizing ammonia (NH_3). The oxidation is never totally complete, however, and uncontrolled emissions from nitric acid plants are typically on the order of 1000 to 3000 ppm NO_x.

Several NO_x control techniques are available and are being used. The IERL Branch Laboratory in Edison, New Jersey has been developing a technique called molecular sieve adsorption. NO_x is removed by converting NO to NO_2 and adsorbing the NO_2. This process results in NO_x concentrations of less than 50 ppm in the emission stream, and the NO_2 which is collected can be used to produce more nitric acid (EPA, Feburary 1980).

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Chapter 8

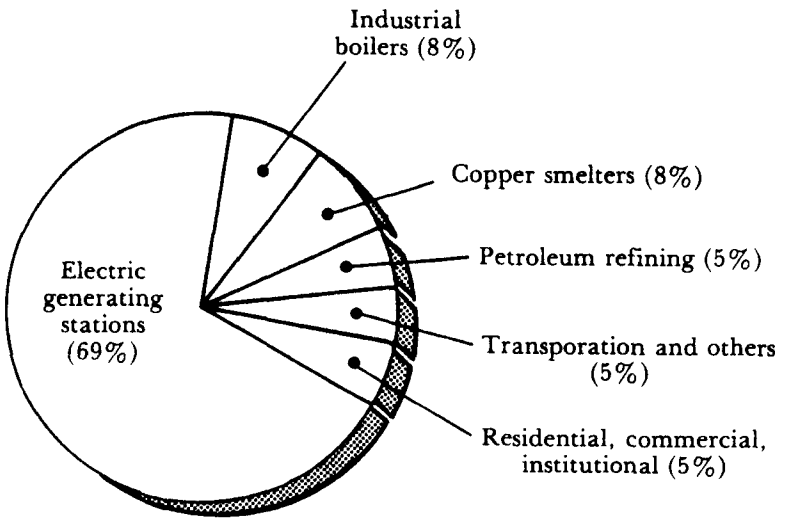
Control of Sulfur Oxide Emissions

Introduction

More than 25 metric tons of sulfur oxides (SO_x) are emitted into the atmosphere each year from manmade sources. These emissions can cause health problems, reduce visibility, and contribute to the global acid rain problem (EPA, August 1980).

Sources of Sulfur Oxides

More than two-thirds of all manmade sulfur oxide emissions result from fossil fuel combustion in electric generating plants (Figure 8-1). Approximately 8% of the total sulfur oxides emitted are from industrial boilers. The largest noncombustion sources are copper smelters, followed by petroleum refining operations. Other small SO_x sources include residential, commercial, and institutional heating furnaces, and transportation vehicles (EPA, August 1980).



Source: EPA, August 1980.

Figure 8-1. Sulfur oxide emissions sources.

The majority of sulfur dioxide (SO_2) emissions from utility boilers occurs when coal is burned. When burned, most of the sulfur (S) contained in coal is oxidized to SO_2. Other fuels used by utilities such as natural gas and oil contain varying amounts of sulfur. Much less SO_2 is emitted from boilers burning natural gas than from boilers burning high sulfur fuel oil. The consumption of coal for electricity

generation is expected to increase annually from 475 million metric tons in 1975, to between 600 and 1000 million metric tons in the year 2000 (EPA, August 1980). The SO_2 emissions from these sources are expected to increase from 20 to 41 million metric tons per year. Therefore, SO_x emissions must be controlled if the current SO_2 ambient air quality levels are expected to be maintained or improved.

Approximately 95% of all sulfur oxides emitted from combustion processes are in the form of sulfur dioxide (SO_2). SO_2 is a colorless gas that when cooled and liquified is used as a bleach, disinfectant, refrigerant, or preservative. The remaining sulfur oxide emissions are in the form of sulfur trioxide (SO_3), sulfates ($SO_4^=$), sulfites ($SO_3^=$) or sulfuric acid (H_2SO_4). SO_2 can form sulfates when oxidized in the atmosphere. Sulfur trioxide is not a stable compound and in the presence of water forms sulfuric acid, a component of acid rain (EPA, August 1980).

Health Effects

High concentrations of sulfur oxides cause breathing difficulty. SO_2 is very soluble and is absorbed in the nose and upper air passages during breathing. This can cause a choking effect known as pulmonary flow resistance. The degree of breathing difficulty is directly related to the amount of SO_2 in the ambient air. Chronic coughing and mucus secretion may result from repeated exposure to high levels of SO_2. The young, elderly, and individuals with chronic lung or heart diseases are most susceptible to the adverse effects of sulfur oxides (EPA, August 1980).

Visibility

Atmospheric visibility is reduced by the presence of small suspended particles in ambient air. Estimates of sulfate concentration in the total suspended particulate matter range from 30 to 50% (EPA, August 1980). Analysis of ambient air quality data in the South Coast Air Basin give estimates of sulfate and nitrate levels as high as 30% of the TSP concentration during air pollution episodes in southern California (Beachler, 1978). Various studies have shown a reduced visibility occurring in the northeastern United States and other areas in the past 25 years. EPA estimated that in some areas on a hot summer day, visibility has decreased from 15 to 8 miles (EPA, August 1980).

Acid Rain

Acid rain primarily consists of sulfuric acid (H_2SO_4) and nitric acid (HNO_3). Sulfuric acid (H_2SO_4) comprises 40 to 60% of acid rain, depending on regional emission patterns (EPA, August 1980). Rain in the northeastern United States averages between 10 and 100 times the acidity of normal rainwater. More than 90 lakes in the Adirondack Mountains in New York no longer contain fish due to the increased acidity of the lake water. In some cases, this acidity has caused toxic metals in the lake beds and surrounding soils to be released into the water, consequently killing fish (EPA, August 1980). Studies indicate damage to agricultural crops, statues and monuments has occurred due to acid rain.

New Source Performance Standards

EPA has promulgated *New Source Performance Standards* (NSPS) for fossil-fuel-fired steam generators (FFFSG) with heat input greater than 250×10^6 Btu/hr (73 MW thermal). These standards establish sulfur oxide emission limits for utility boilers. There are two standards: one for steam generators for which construction commenced after August 17, 1971; the other for steam generators for which construction commenced after September 18, 1978 (Table 8-1). The term "construction commenced" is defined in the Clean Air Act and does not necessarily mean that actual physical construction was initiated. It may be interpreted as the date that a permit process is initiated at the air pollution control agency. These standards also specified the degree of SO_2 reduction required for utility boilers.

Table 8-1. New Source Performance Standards for SO_2 emissions from fossil-fuel-fired steam generators rated greater than 250×10^6 Btu/hr or 73 MW (thermal) heat input.

SO_2 emissions	Metric units (ng/J)	English units (lb/10^6 Btu)	Fuel
Subpart D; new sources after August 17, 1981	340	0.8	Liquid
	520	1.2	Solid (coal)
Subpart Da; new sources built after September 18, 1978	340	0.8	Liquid or gaseous
	and 90% SO_2 reduction is required unless emissions are less than		
	86	0.2	
	520	1.2	Solid (coal)
	and 90% SO_2 reduction required unless SO_2 emissions are less than		
	260	0.6	
	then 70% SO_2 reduction is required		

The NSPS for steam generators built after September 18, 1978 require flue gas scrubbing to reduce SO_2 emissions. In generators burning coal, the SO_2 emissions must be less than 1.2 lb/10^6 Btu. Also a 90% reduction of SO_2 emissions is required. If the emissions are less than 0.6 lb/10^6 Btu (coal fired units) then only 70% SO_2 emission reduction is required. The purpose of this "two part" standard is to encourage the use of new technology for reducing SO_2 emissions from boilers using low sulfur coal (JAPCA, 1979). Generators burning low sulfur coal generally have SO_2 emissions less than 0.6 lb/10^6 Btu. For oil fired utility boilers the SO_2 emissions must be less than 0.8 lb/10^6 Btu and 90% SO_2 emission reduction (scrubbing) is required; unless the emissions are less than 0.2 lb/10^6 Btu. In this case, no scrubbing is required.

For example, a power plant ($>250 \times 10^6$ Btu/hr) cannot have SO_2 emissions exceeding 1.2 lb/10^6 Btu. In addition, the plant is required to reduce the SO_2 emissions by 90%. If the plant's emissions are less than 0.6 lb/10^6 Btu then only 70% scrubbing is required. It is possible that the SO_2 emissions will fall between 0.6 and 1.2 lb/10^6 Btu. In this case, 90% scrubbing is required. It is also possible

to meet the NSPS standard by scrubbing 85% **if and only if** the plant's emissions **do not exceed** 0.6 lb/10^6 Btu. A useful graphical representation of the 1978 NSPS for sulfur dioxide emission limitations from fossil-fuel-fired steam generators can be found in JAPCA (1980).

SO_2 Emission Reduction Technology

Sulfur oxide emissions from fossil-fuel-fired combustion sources can be reduced by five techniques: *fuel desulfurization, combustion of coal and limestone mixtures, coal gasification, coal liquefaction,* and *flue gas desulfurization.*

The EPA's Office of Research and Development is providing research and development projects on these technologies through grants, contracts, and cooperative agreements.

Fuel Desulfurization

One of the most straightforward ways to reduce SO_2 emissions from combustion sources is by burning fuel containing less sulfur. This might involve using low sulfur coal, low sulfur fuel oil, or natural gas instead of a high sulfur coal. The use of low sulfur coal (usually supplied from western U.S.) can reduce SO_2 emissions. In some cases, low sulfur coal has been used to meet State and Federal (1.2 lb/10^6 Btu) air pollution regulations. Low sulfur coal generally contains less than 1% sulfur; high sulfur coal contains between 3 and 5% sulfur. Low sulfur coal is more expensive and less available (in the eastern U.S.) than high sulfur coal. However, the sulfur content can be reduced by a process called fuel desulfurization.

Fuel desulfurization removes or reduces the sulfur content of coal before it is burned. Since coal is projected to be the major domestic fuel to meet future energy demands, fuel desulfurization can be a very useful method for reducing SO_2 emissions. Coal contains sulfur in two forms: mineral sulfur in the form of inorganic pyrite and organic sulfur which is chemically bound to the coal. The amount of inorganic (and organic) sulfur in coal varies from 10 to 90%. Mineral sulfur can be removed by physical coal cleaning, but organic sulfur requires chemical cleaning.

Physical coal cleaning depends on the differences in density of both coal and the impurities. Coal is crushed, washed, and then separated by settling processes using cyclones, air classifiers or magnetic separators. Approximately 40 to 90% of the pyritic sulfur content can be removed by physical coal cleaning (EPA, August 1980). Its effectiveness depends on the size of pyritic sulfur particles and the amount of pyritic sulfur contained in the coal.

Chemical coal cleaning methods that reduce the organic bound sulfur are currently under development. The Industrial Environmental Research Laboratory (IERL) of the EPA is involved in two chemical coal cleaning technologies: *microwave desulfurization* and *hydrothermal desulfurization.*

In microwave desulfurization the coal is crushed, then heated for 30 to 60 seconds by exposure to microwaves. Mineral sulfur selectively absorbs this radiation forming hydrogen sulfide gas (H_2S). The H_2S is usually reduced to elemental sulfur

by the Claus process. Another microwave process adds calcium hydroxide [$Ca(OH)_2$] to crushed coal. The organic sulfur converts to calcium sulfite ($CaSO_3$) when exposed to this radiation. The coal is washed with water to remove the $CaSO_3$ and other impurities. As much as 70% of the sulfur can be removed by the microwave process (EPA, August 1980).

Hydrothermal desulfurization, developed by Battelle Laboratories in Columbus, Ohio, is being refined through experiments sponsored by EPA's Office of Research and Development (ORD). Coal is crushed and mixed with a solution of sodium and calcium hydroxides [NaOH and $Ca(OH)_2$]. When this mixture is heated to 275°C in a pressurized vessel, most of the pyritic sulfur and 20 to 50% of the organic sulfur is converted to sodium and calcium sulfites (Na_2SO_3 and $CaSO_3$)(EPA, August 1980). The coal is rinsed to remove the sulfites and the water is processed to recycle the sodium and calcium hydroxides. This process is an expensive but effective method for removing sulfur from coal. ORD is directing current efforts toward reducing high process costs and finding alternative methods for drying the coal and recovering the sodium and calcium hydroxides.

Combustion of Coal and Limestone Mixtures

Sulfur oxides can be removed by burning coal and limestone mixtures in a boiler. Two promising burning technologies currently under development are: *fluidized bed combustion* and the use of *limestone coal pellets as fuel.*

In the fluidized bed combustion process, a grid supports a bed of coal and limestone (or dolomite) in the firebox of the boiler. Combustion air is forced upward through the grid suspending the coal and limestone bed in a fluid-like motion. Natural gas is used to ignite the pulverized coal. Once the coal is ignited, the gas is turned off. The sulfur in the coal is oxidized to SO_2 and consequently combined with the limestone to form calcium sulfate ($CaSO_4$). The $CaSO_4$ and flyash particulate matter are usually collected in a baghouse or electrostatic precipitator (see APTI Course 413 Student Manual). A 100,000 lb steam/hr fluidized bed boiler is now operating at Georgetown University, Washington, DC. The sulfur dioxide emissions from this unit have been reduced 75 to 90% (Milliken and Young, 1981). The nitrogen oxide emissions were reported to be less than 0.7 lb/10^6 Btu due to low combustion temperatures (815 to 870°C). Other demonstrations of fluidized bed combustion are being conducted at several sites including use of a 30 MW boiler at Rivesville, West Virginia (EPA, May 1979). This technology is expected to be available for commercial industrial boilers in the early 1980's.

Another way to reduce SO_2 emissions from combustion processes is by burning pellets made of limestone and coal. ORD research has shown that SO_2 emissions from conventional stoker and pulverized coal fired boilers can be reduced by burning coal and limestone pellets. Pellets are made by pulverizing coal and limestone and adding a binder forming small, cylinder shaped pellets which are about half the size of a charcoal briquette. These consist of approximately two-thirds coal, and one-third limestone. As the pellet burns, the calcium in the limestone absorbs the SO_2 generated by coal combustion and forms calcium sulfate.

Calcium sulfate emissions are collected in a baghouse or electrostatic precipitator. SO_2 emissions have been reduced on a pilot scale boiler by as much as 80% (Mobley and Linn, 1981).

Coal Gasification

Over 70 different processes have been developed for producing a combustible gas from coal. Three basic steps are common to all coal gasification processes: pretreatment, gasification, and gas cleaning. Coal pretreatment involves coal pulverizing and washing. The pulverized coal is gasified in a reactor with limited oxygen. Gasification produces either a low, medium, or high Btu gas by applying heat and pressure or by using a catalyst to break down the components of coal (EPA, August 1980). The gas produced contains carbon monoxide (CO), hydrogen (H_2), carbon dioxide (CO_2), water (H_2O), methane (CH_4), and contaminants such as hydrogen sulfide and char. Low and medium Btu gas contains more CO and H_2 than high Btu gas which contains a higher CH_4 content. Methane gas produces more heat when burned (see Table 3-1). The sulfur in the coal is converted to H_2S during gasification. H_2S is removed during the gas cleaning step, generally by a scrubbing process. H_2S is converted to elemental sulfur by partial oxidation and catalytic conversion (EPA, August 1980). The synthetic gas produced is sulfur free and can be burned without releasing harmful pollutants. Commercial operation of coal gasification units is expected in the mid 1980's.

Coal Liquefaction

A process for changing coal into synthetic oil is called coal liquefaction. Coal liquefaction is similar to coal gasification. Two basic approaches for liquefaction are used. One involves using a gasifier to convert coal to carbon monoxide, hydrogen and methane; followed by condensation to convert the gases to oils. The second approach uses a solvent or slurry to liquify pulverized coal and then processes this liquid into a heavy fuel oil. Some processes produce both a synthetic gas and synthetic oil.

Hydrogen is used to convert sulfur in the coal to hydrogen sulfide gas. Hydrogen sulfide is partially oxidized to form elemental sulfur and water. More than 85% of the sulfur is removed from coal by liquefaction (EPA, August 1980). EPA is currently working on improving these processes. Commercial operation of liquefaction processes is expected in the early 1990's.

Flue Gas Desulfurization

Flue gas desulfurization (FGD) is the most popular technology used for controlling sulfur oxide emissions from combustion sources. FGD technology is also used to reduce SO_x emissions from copper smelters. In this method SO_x gaseous emissions are generally removed by a post-combustion absorption process. This process uses an absorbent in a vessel and produces a sludge that must be discarded.

FGD scrubbing processes can either be *wet* or *dry*. Wet scrubbing processes use a liquid absorbent to absorb the SO_2 gases. Wet scrubbing can be further categorized into *nonregenerable* and *regenerable* processes. Nonregenerable processes produce a sludge that must be disposed of properly; sometimes referred to as throw away FGD processes. Regenerable processes generate a usable product in addition to removing sulfur oxides. Regenerated products include elemental sulfur, sulfuric acid, or gypsum wall-board. Dry scrubbing processes use a dry spray to absorb SO_2 gas and form dry particles. These dry particles are collected in a baghouse or electrostatic precipitator.

More than 50 different FGD processes have been developed. The majority of these systems operating in Japan and the United States are capable of reducing the sulfur oxide emissions from utility boilers by at least 90%. This chapter will address the most important processes used in the U.S. today.

Nonregenerable FGD Processes

Nonregenerable FGD processes are those which generate a sludge or waste product as a result of SO_2 emission reduction. The sludge must be disposed of properly in a pond or landfill. The three most common nonregenerable processes used in the U.S. today are *lime scrubbing, limestone scrubbing* and *double alkali scrubbing*. Approximately 75% of the current FGD installations in the U.S. are either lime or limestone scrubbing (EPA, March 1978). However, this number could change with additional use of regenerable processes to eliminate sludge disposal problems.

Lime Scrubbing

Lime scrubbing uses an alkaline slurry made by adding lime (CaO), usually 90% pure, to water. The alkaline slurry is sprayed in an absorber and reacts with the SO_2 in the flue gas. Calcium sulfite ($CaSO_3$) and calcium sulfate ($CaSO_4$) salts are formed in the reaction and are removed as sludge. The sludge produced can be stabilized to produce an inert landfill material or can be stored in sludge ponds.

A number of chemical reactions take place in the absorber. SO_2 can be absorbed in the water and form sulfite ($SO_3^=$) and sulfate ($SO_4^=$) ions. Lime (CaO) is slaked with water to produce a calcium slurry of CaO and H_2O or calcium hydroxide [$Ca(OH)_2$]. The calcium hydroxide/water mixture is a solution containing calcium ions (Ca^{++}) and hydroxide ions (OH^-). Calcium ions combine with sulfate and sulfite ions to produce a calcium sulfite and calcium sulfate sludge. The basic reactions occurring are:

$$SO_2 + CaOH + \tfrac{1}{2} H_2O \rightarrow CaSO_3 + H_2O$$

$$SO_2 + CaOH + \tfrac{1}{2} O_2 + \tfrac{1}{2} H_2O \rightarrow CaSO_4 + H_2O$$

Lime, in a stoichiometric ratio of 1.0 to 1.1, will remove 1 mole of SO_2 gas.

System Description

The equipment necessary for SO_2 emission reduction comes under four operations:

1. Scrubbing or absorption: accomplished with scrubbers, holding tanks, liquid spray nozzles, and circulation pumps.
2. Lime handling and slurry preparation: accomplished with lime unloading and storage equipment, lime processing and slurry preparation equipment.
3. Sludge processing: accomplished with sludge clarifiers for dewatering, sludge pumps and handling equipment, sludge solidifying equipment.
4. Flue gas handling: accomplished with inlet and outlet ductwork, dampers, fans, and stack gas reheaters.

The individual FGD systems vary depending on the FGD vendor and individual plant installations. For example, some FGD systems have used a scrubber and absorber for both particulate and SO_2 emission control (Figure 8-2). This system is used at Pennsylvania Power's Bruce Mansfield 1 and 2 in Shippingport, Pennsylvania. This FGD system is designed to remove 99.8% particulate matter and 92.1% SO_2 emissions on an 825 MW coal fired utility boiler (1650 MW for both units). Bruce Mansfield 1 was installed in late 1975. The system consists of six parallel scrubber and absorber trains. Each train (Figure 8-2) consists of a variable-throat venturi scrubber, a wet induced draft fan, and a fixed-throat venturi scrubber (EPA, January 1979).

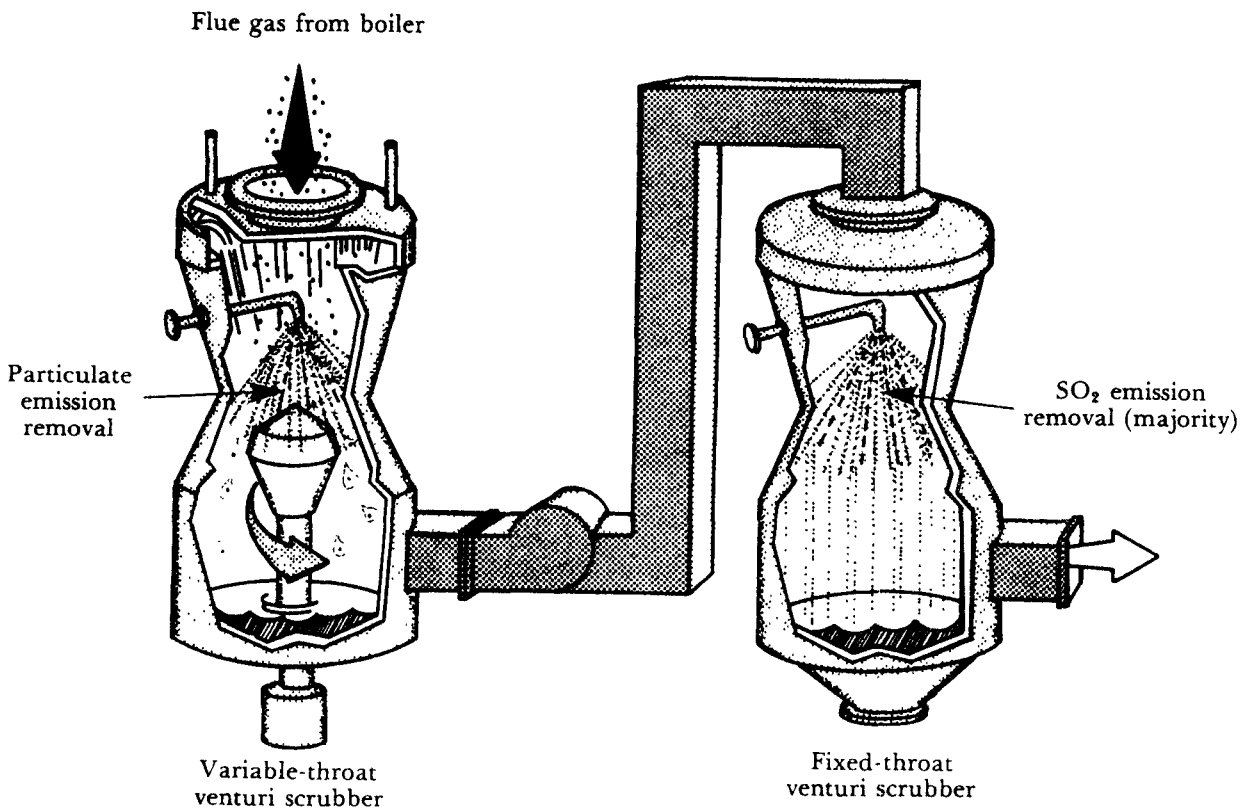


Figure 8-2. Scrubber and absorber train for lime scrubbing system.

Other FGD systems have been installed that consist of an electrostatic precipitator for particulate emission control and a mobile bed absorber for SO_2 emission control (Figure 8-3). This system was installed at Louisville Gas and Electric Company's Cane Run 190 MW power station, Louisville, Kentucky. The system is designed for 99.0% particulate emission control and 85 to 89% SO_2 emission control. The startup of this system was in August, 1976 (EPA, January 1979).

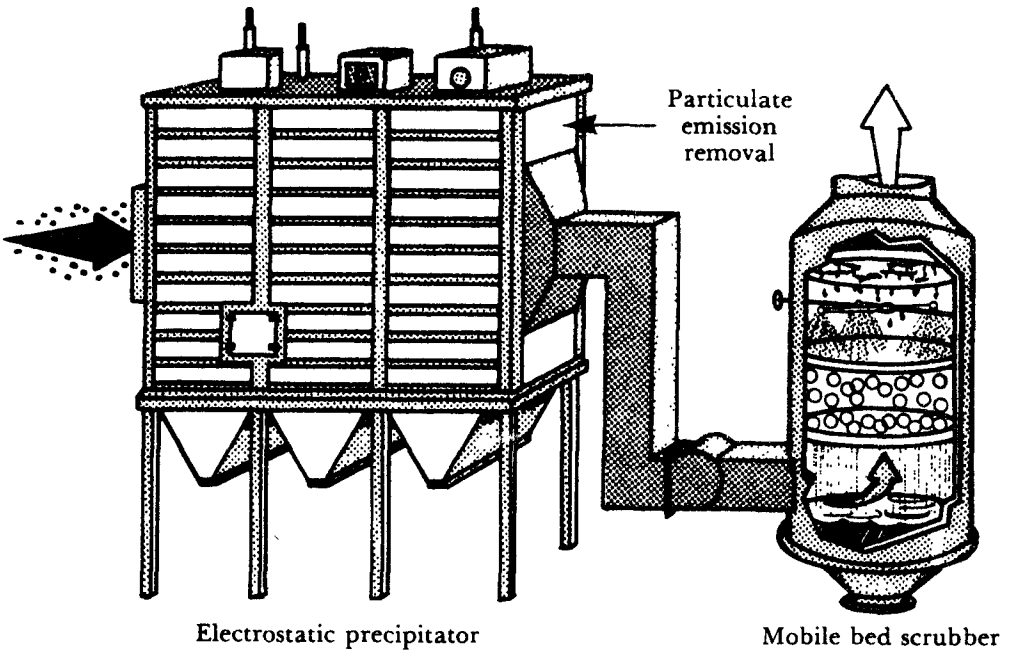


Figure 8-3. Electrostatic precipitator and mobile bed scrubber train for lime scrubbing system.

A listing of selected lime scrubbing systems installed in the U.S. is given in Table 8-2. This listing contains the unit rating for the power plant, the FGD system, emission control efficiency, startup dates and sludge disposal method. Additional installations can be found in *Electric Utility Generating Units—Flue Gas Desulfurization Capabilities as of October 1978* (EPA, January 1979).

Table 8-2. Selected lime scrubbing FGD systems.

Station	Size (MW)	Initial startup date	New or retrofit	Particulate emission control	SO_2 emission control	Coal sulfur content (%)	SO_2 emission removal efficiency (%)	Sludge disposal
Kentucky Utilities; Green River 1, 2 and 3	64	8/75	Retrofit	—	—	3.7	80	—
Pennsylvania Power Co.; Bruce Mansfield 1 and 2	825 825	12/75 (unit 1) 10/77 (unit 2)	New	Variable-throat venturi scrubber	Fixed-throat venturi scrubber	2.4	92-95	Stabilized sludge in an off-site dammed resevoir
Columbus and Southern Ohio Electric Co.; Conesville 5 and 6	411 375	1/77 6/78	New	Electrostatic precipitator	Turbulent contact absorber (mobile bed)	4.2 to 5.1	89.6	Stabilized sludge disposed in an outside landfill
Louisville Gas and Electric; Cane Run 4	178	8/76	Retrofit	Electrostatic precipitator	Mobile bed absorbers	3.8	86-89	Stabilized sludge in an on-site pond
Louisville Gas and Electric; Cane Run 5	183	12/77	Retrofit	Electrostatic precipitator	Spray towers	3.8	85	Stabilized sludge in an on-site pond

Sludge produced from a lime scrubbing FGD system must be discarded properly. It can be stored in a settling pond or dewatered to take up less space in a landfill. Clarifiers, centrifuges and vacuum filters accomplish dewatering. Occasionally chemicals are used to make the sludge into a cement-like material. The Bruce Mansfield plant uses a material called Calcilox (developed by Dravo Corporation). The cement-like sludge is permanently stored in an earthen dammed reservoir 7 miles from the plant site (EPA, January 1979).

Operating Experience

Early lime FGD systems were plagued with a number of operational and maintenance problems. One problem was that spray nozzles and mist eliminators became clogged with hardened scrubbing slurry. Scale buildup ($CaSO_4$) on the spray nozzles limited scrubber operation time. New spray nozzle design and careful control of the recirculated slurry have reduced internal scrubber scaling (EPA Technology Transfer, Second Progress Report for Lime/Limestone Scrubbing). Mist eliminator clogging was a major problem at the Bruce Mansfield plant. Scaling on the mist eliminators caused excessive pressure drops. A wash system was devised to reduce excessive scale buildup. Other problems with occasional mist

carry-over caused scale buildup on flue linings and caused acid rainout in areas near the stack. These problems have been eliminated by the careful design of mist eliminators and by monitoring pressure drop across them (EPA, January 1979).

Another problem concerned stack gas reheaters. Stack gas is reheated to avoid condensation and corrosion to ductwork and the stack; and to enhance plume rise and pollutant dispersion. Reheat is accomplished by using steam coils in the stack or by using hot air supplied by auxiliary oil heaters in the stack. Some reheater failures were caused by acid attack to reheater components. Other reheaters vibrated too much causing structural deterioration. These problems have been eliminated by changes in reheater design (EPA, January 1979).

Corrosion of scrubber internals, fans and ductwork, and stack linings have been reduced by using special materials such as rubber or plastic coated steel and by carefully controlling slurry pH with monitors. Additional operation and maintenance problems and solutions are found in *Proceedings: Symposium on Flue Gas Desulfurization, Volume I and II* (EPA, March 1978, July 1979, and 1981).

Limestone Scrubbing

Limestone scrubbing uses an alkaline slurry from *limestone* ($CaCO_3$) in an absorber to react with SO_2 in the flue gas. Calcium sulfite ($CaSO_3$) and calcium sulfate ($CaSO_4$) salts are formed in the reaction and are removed as sludge. Two major differences between lime and limestone scrubbing are their uses of feed preparation equipment and their liquid-to-gas ratios. Even with these differences, these processes are so similar that an FGD system can be set up to use either lime or limestone to absorb SO_2 gas.

Limestone scrubbing's process chemistry is also very similar to lime scrubbing's. Limestone ($CaCO_3$) is slaked with water to form aqueous $CaCO_3$ and is sprayed in the absorber. Sulfite and sulfate ions are produced as SO_2 gas contacts the water. These ions combine with calcium ions to produce calcium sulfite and calcium sulfate sludge. The basic reactions are:

$$SO_2 + CaCO_3 + H_2O + O_2 \rightarrow CaSO_3 + H_2O + CO_2 + O_2$$

$$SO_2 + CaCO_3 + H_2O + \tfrac{1}{2} O_2 \rightarrow CaSO_4 + H_2O + CO_2$$

System Description

The equipment necessary for SO_2 absorption is the same as for lime scrubbing except in the slurry preparation. The limestone feed (rock) is reduced in size by crushing it in a ball mill. Limestone is sent to a size classifier. Pieces larger than 200 mesh are sent back to the ball mill for recrushing. Limestone is slaked with water in a slurry supply tank. Limestone is generally a little cheaper than lime, making it very popular for use in FGD systems.

Limestone FGD systems vary, depending on the FGD vendor and individual plant installations. The Kansas City Power and Light La Cygne power station uses a venturi scrubber for particulate emission control and a sieve tray absorption tower for SO_2 emission control. The 820 MW plant uses eight scrubbing modules; one venturi scrubber and one absorber per module as shown in Figure 8-4. The Northern States Power Co. Sherburne Station No. 1 and 2 uses a scrubbing system consisting of 12 scrubber and absorber modules. Other systems use electrostatic precipitators for particulate emission control and absorbers for SO_2 emission control. A listing of selected limestone scrubbing U.S. installations is given in Table 8-3.

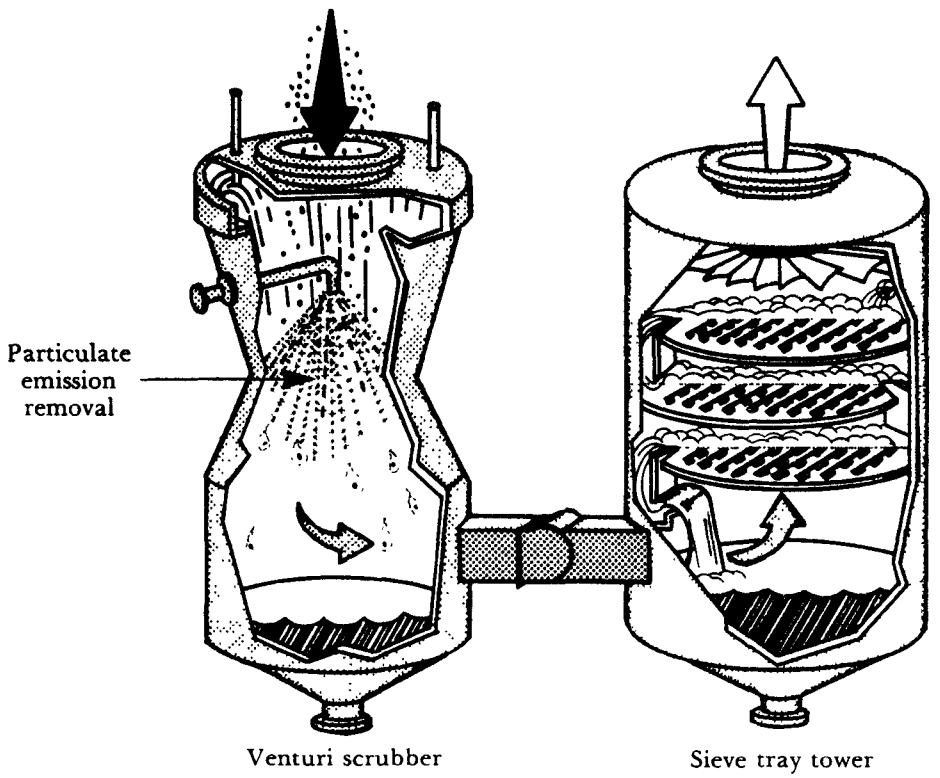


Figure 8-4. Limestone scrubbing system.

Table 8-3. Selected limestone scrubbing FGD systems.

Station	Size (MW)	Initial startup date	New or retrofit	Particulate emission control	SO_2 emission control	Coal sulfur content (%)	SO_2 emission removal efficiency (%)	Sludge disposal
Kansas City Power and Light Co.; La Cygne 1	820	2/73	New	Venturi scubber	Sieve tray absorber	5.0	80	Settling pond
Northern States Power Co.; Sherburne 1 and 2	680 680	3/76 1/77	New	Rod deck venturi scrubber	Spray tower	0.8	50-55	Settling pond
Arizona Public Service Co.; Cholla 1	119	10/73	Retrofit	Cyclone and flooded disc venturi scrubber	Spray tower	0.4-1.0	50-60	Unstabilized sludge in an on-site pond
Arizona Public Service Co.; Cholla 2	250	6/78	New	Flooded disc scrubber	Packed tower	0.4-1.0	75	Unstabilized sludge in an on-site pond
Central Illinois Light; Duck Creek 1	400	8/78	New	Electrostatic precipitator	Rod deck venturi scrubber	2.5-3.0	92	Unstabilized sludge in an on-site pond
Tennessee Valley Authority; Widows Creek	550	5/77	Retrofit	Variable-throat venturi scrubber	Multigrid absorption tower	3.7	85-94	Unstabilized sludge in an on-site pond
Texas Utilities; Martin Lake 1 and 2	750 (each)	10/77	New	Electrostatic precipitator	Packed tower	1.0 0.9	74 71	Stabilized dewatered sludge in an on-site landfill
Kansas Power and Light; Lawrence 4	125	1/76	Retrofit	Rod deck venturi scrubber	Spray tower	0.55	96-98	Unstabilized sludge in an on-site pond
Kansas Power and Light; Lawrence 5	420	4/78	New	Rod deck venturi scrubber	Spray tower	0.55	65	Unstabilized sludge in an on-site pond

Operating Experience

Early limestone FGD systems had scrubber operating problems similar to lime scrubbing systems. Plugged and clogged nozzles, scrubber internals, and mist eliminators resulted from inefficient SO_2 absorption by limestone in the scrubber. Increased absorption efficiency is achievable at high pH values since more alkali is available to dissolve the SO_2 gas. However, scaling problems will occur if the scrubber is operated at very high pH values (>9.0). The pH levels must be maintained by carefully controlling lime or limestone feed and mixing with water to prevent inefficient system operation. If the pH gets too low (<5.0) the removal efficiencies are low; and at high pH levels (>9.0), scaling of scrubber components will occur.

As can be seen from Tables 8-2 and 8-3 the SO_2 removal efficiencies for various lime and limestone FGD installations range from 50 to 92%. These FGD systems were designed to meet existing air pollution regulations. Lime and limestone FGD

systems are capable of removing SO_2 with efficiencies in excess of 90% (EPA, March 1978; EPA 600/7-78-032a). The addition of small amounts of soluble *magnesium* (< 1% by weight) to the scrubber liquor can greatly increase SO_2 removal efficiencies to as high as 99% (EPA, March 1978; EPA 600/7-78-032a). Magnesium is added in the form of magnesium oxide, magnesium sulfate or dolomitic lime (used in lime scrubbing systems). Magnesium compounds are more soluble than calcium compounds and react rapidly with SO_2.

EPA is currently working on a program that uses an additive of *adipic acid* to limestone FGD systems. Adipic acid can increase SO_2 removal efficiencies from 85% to as high as 97% (EPA, August 1980). Adipic acid is a crystalline powder derived from petroleum. EPA experiments have shown that when a limestone slurry reacts with SO_2 in the scrubber, the slurry becomes very acidic. This acidity limits the amount of SO_2 that can be absorbed. Adding adipic acid to the slurry slightly increases the slurry's initial acidity, but prevents it from becoming highly acidic during the absorption of SO_2. The net result is an improvement in the scrubbing efficiency.

EPA research has shown that adipic acid can reduce the total limestone consumption by as much as 15%, thus reducing operating costs. Adipic acid is nontoxic (it is used as a food additive) and does not degrade the calcium sulfite and calcium sulfate sludge. Adipic acid is not currently being used in a commercial FGD installation. Full scale tests at an electrical generating station are in the planning stage (EPA, August 1980).

Another scrubber operating problem occurring in lime and limestone FGD systems is that calcium sulfite is formed as part of the sludge. Calcium sulfite settles and filters poorly and can be removed from the scrubber slurry only in a semiliquid or paste-like form. This semiliquid waste must then be stored in lined ponds. The EPA's IERL group at RTP, NC is developing a way to solve this problem through a process called *forced oxidation*.

In forced oxidation, air is blown into the scrubber slurry tank that contains primarily calcium sulfite and water. The air oxidizes the calcium sulfite to calcium sulfate.

$$CaSO_3 + H_2O + \tfrac{1}{2} O_2 \rightarrow CaSO_4 + H_2O$$

Calcium sulfate formed by this reaction grows to a larger crystal size than calcium sulfite. As a result, the calcium sulfate is easily filtered, forming a drier and more stable material which can be disposed of in a landfill. This material ($CaSO_4$) can also be used for cement, gypsum wallboard, or as a fertilizer additive.

Forced oxidation can also help control scaling problems to scrubber internals. This process helps control scale by removing calcium sulfite from the slurry in the form of calcium sulfate which is more easily filtered. This will prevent calcium sulfites and calcium sulfates from being recirculated in the absorber.

Double Alkali Scrubbing

Double or dual alkali scrubbing is a nonregenerable FGD process that uses two alkali sprays to remove SO_2 from combustion exhaust gas. A sodium solution absorbs SO_2 followed by absorbent regeneration. The spent absorbent is reacted with a lime or limestone alkali slurry. Calcium sulfites and sulfates are precipitated and discarded as sludge. The regenerated sodium scrubbing solution is returned to the absorber loop. The double alkali process has reduced plugging and scaling problems in the absorber since sodium scrubbing compounds are very soluble. FGD systems are capable of 95% SO_2 reduction.

Particulate matter is removed prior to scrubbing in the absorber by an electrostatic precipitator or a venturi scrubber. This is done to prevent fly ash erosion to the absorber internals and to prevent any appreciable oxidation to the sodium solution in the absorber due to catalytic elements in the fly ash (EPA, March 1978). A typical dual alkali system is shown in Figure 8-5. SO_2 removal in the absorber occurs by reacting SO_2 gas with a circulating sodium alkaline solution.

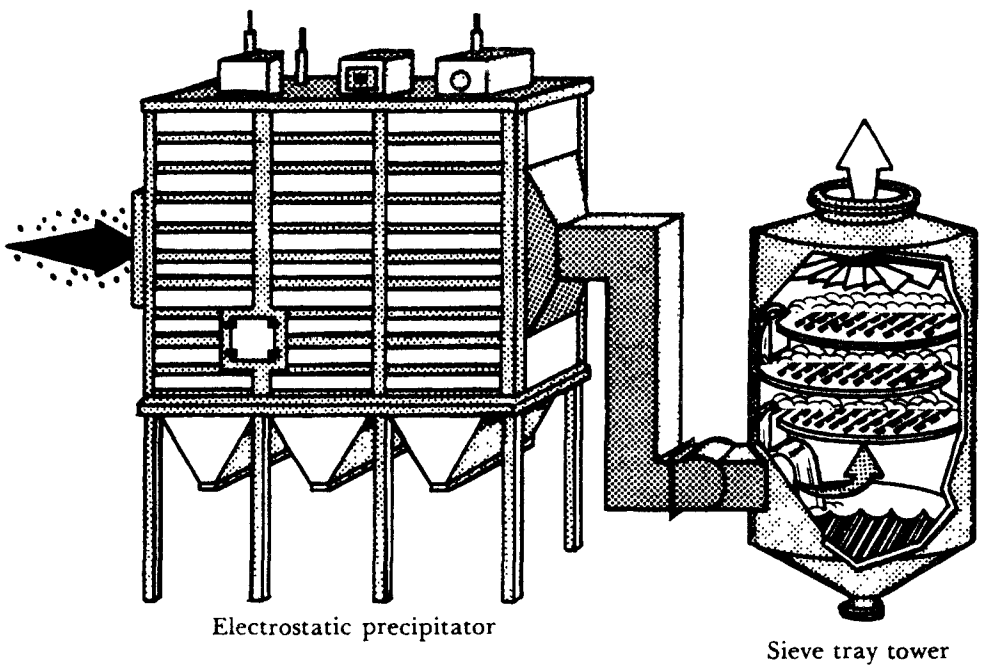


Figure 8-5. Dual alkali scrubbing system.

The sodium solution is usually a mixture of sodium carbonate (soda ash, Na_2CO_3), sodium sulfite (Na_2SO_3), and sodium hydroxide (caustic, NaOH). During the reaction the sodium solution is consumed to form sodium bisulfite ($NaHSO_3$), sodium sulfate (Na_2SO_4), and sodium bicarbonate ($NaHCO_3$) salts. SO_2 absorption can be represented by the following simplified reactions:

$$2Na_2CO_3 + SO_2 + H_2O \rightarrow Na_2SO_3 + 2NaHCO_3$$

$$NaHCO_3 + SO_2 \rightarrow NaHSO_3 + CO_2\uparrow$$

$$2NaOH + SO_2 \rightarrow Na_2SO_3 + H_2O$$

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3$$

$$2NaOH + SO_3 \rightarrow Na_2SO_4 + H_2O$$

$$Na_2SO_3 + O_2 \rightarrow Na_2SO_4$$

After reaction in the absorber, spent scrubbing liquor is bled to a reactor tank for regeneration. Sodium bisulfite and sodium sulfate are inactive salts and do not absorb any SO_2. Actually, it is the hydroxide ion (OH), sulfite ion (SO_3) and carbonate ion (CO_3) that absorb SO_2 gas. Sodium bisulfite and sodium sulfate are reacted with lime or limestone to produce a calcium **sludge** and a **regenerated sodium solution**:

$$2NaHSO_3 + \underset{\text{(lime)}}{Ca(OH)_2} \rightarrow Na_2SO_3 + \underset{\text{(sludge)}}{CaSO_3 \cdot \tfrac{1}{2}H_2O\downarrow} + 3/2H_2O$$

$$Na_2SO_4 + \underset{\text{(lime)}}{Ca(OH)_2} \rightarrow 2NaOH + \underset{\text{(sludge)}}{CaSO_4\downarrow}$$

From the reactor, the slurry is pumped either to a clarifier or thickener where precipitated solids (sludge) are separated from the scrubbing liquor. These solids are dewatered by a vacuum filter and occasionally stabilized with a chemical or a lime and fly ash mixture. Unstabilized sludge is discarded in a settling pond; stabilized sludge in a proper landfill. Some sodium sulfate is unreacted (lost) in the regeneration step. Additional sodium is added to the regenerated solution in the form of soda ash or caustic soda. This regenerated absorbent is now ready to be used again.

System Description

Currently, over 12 dual alkali processes are installed in the U.S. as FGD systems on utility and industrial boilers. Louisville Gas and Electric Co. installed this process on a 300 MW boiler at the Cane Run power station. Particulate matter is removed by an electrostatic precipitator before entering the absorber. The absorber is a tray tower that is designed to remove 90 to 95% of the SO_2 emissions generated from burning 3.5 to 4.0% sulfur coal. A dual alkali system using three absorbers was installed on three industrial boilers (total 165 MW) at the Arco Polymers plant in Monaca, Pennsylvania. This system removed 86 to 93% of the SO_2 emissions generated from burning coal with up to 3% sulfur content. This system was brought on line on August 16, 1980 and after a shakedown period has been operating in compliance with State regulations of 0.75 lb/10^6 Btu (Reiners et al., 1981). A listing of some selected dual alkali installations is presented in Table 8-4.

Table 8-4. Selected dual alkali FGD systems.

Station or plant	Size (MW)	Initial startup date	Particulate emission control	SO_2 emission control	Coal sulfur content (%)	SO_2 removal efficiency (%)
Louisville Gas and Electric; Cane Run 6, Louisville, KY	300	2/79	Electrostatic precipitator	Tray tower	3.5-4.0	90-95
Southern Indiana Gas and Electric Co.; West Franklin, Ind.	500	4/79	Electrostatic precipitator	Two stage disc absorber	4 5	85
Arco Polymers; Monaca, PA	165	8/80	Electrostatic precipitator	Four stage disc and doughnut baffle absorber	3.0	86-93
Caterpillar tractor; Joliet, IL	34	1974	Mechanical collector	2 Zurn dustraxtor scrubbers	3.2	90
General Motors; Parma, OH	64	1974	--	Tray tower	2.5	90
Central Illinois Public Service; Newton, IL	575	1979	Spray tower	Mobile bed absorber	4.0	95

Operating Experience

As can be seen from Table 8-5, dual alkali scrubbing systems have been operating for a number of years. Typical problems with pH control, mist eliminator liquid carry-over, raw materials handling, and regenerated scrubber liquor pumping operations have occurred. These problems have been rectified and installed systems have been operating reliably for a number of years. The major benefit of the dual alkali process is that by using a sodium solution as the absorbent, many scaling problems that have plagued lime and limestone scrubbers can be eliminated. However, double alkali scrubbing is more expensive than lime and limestone scrubbing.

Regenerable FGD Processes

Regenerable FGD processes are wet scrubbing processes that remove SO_2 from the flue gas and generate some usable product. Regenerated products include elemental sulfur, sulfuric acid, or in the case of lime and limestone scrubbing, gypsum wallboard. Regenerable processes used on industrial and utility boilers include *Wellman-Lord, magnesium oxide*, and *citrate*. Approximately 4% of all operating and committed FGD systems in the U.S. use the Wellman-Lord process; 2% use the magnesium oxide process; and less than 1% use the citrate process (JAPCA, 1981). The major advantage of a regenerable process as opposed to a nonregenerable process is the avoidance of a sludge disposal problem.

Wellman-Lord

The Wellman-Lord process is a regenerable FGD process used to reduce SO_2 emissions from utility and industrial boilers and produce a usable product. This process is sometimes referred to as the Wellman-Lord/Allied Chemical process; Allied Chemical referring to the regeneration step.

SO_2 is absorbed by aqueous sodium sulfite solution forming a solution of sodium bisulfite. The solution is sent to an evaporator-crystallizer where sodium sulfite and concentrated SO_2 gas are produced. Sodium sulfite is recycled to the absorber and concentrated SO_2 is converted to a saleable product; usually sulfuric acid or elemental sulfur. Flue gas from the boiler is pretreated prior to entering the absorber. Particulate matter is removed by an electrostatic precipitator, fabric filter, or a wet scrubber. A small venturi scrubber removes any additional particulate matter, chlorides and sulfur trioxide. Flue gas is then sent to the absorber where SO_2 gas is absorbed according to the following reaction:

$$SO_2 + Na_2SO_3 + H_2O \rightarrow 2NaHSO_3$$

Some oxidation occurs in the absorber forming sodium sulfate which is unreactive with SO_2 gas:

$$Na_2SO_3 + \tfrac{1}{2}O_2 \rightarrow Na_2SO_4$$

The formation of sodium sulfate depletes the supply of sodium sulfite available for scrubbing. This can be made up by adding sodium carbonate to the scrubbing slurry to combine with sodium bisulfite according to the following chemical reaction:

$$Na_2CO_3 + 2NaHSO_3 \rightarrow 2Na_2SO_3 + CO_2\uparrow + H_2O$$

System Description

The Wellman-Lord process was installed at the Northern Indiana Public Service Company's Dean H. Mitchell plant on a 115 MW boiler. The process equipment includes: an electrostatic precipitator for removing particulate matter; a venturi scrubber for cooling flue gas and for removing SO_3 and chlorides; an SO_2 absorber; an evaporator-crystallizer for regenerating the absorbent; and the Allied Chemical process for reducing concentrated SO_2 gas into elemental sulfur or sulfuric acid (Figure 8-6). The absorber is a tray tower with three stages. SO_2 gas is scrubbed with a sodium sulfite solution at each stage. A mist eliminator removes entrained liquid droplets from gas exiting the absorber. There is a direct-fired natural gas reheat system in the absorber stack to reheat cleaned gas for good dispersion of the steam plume.

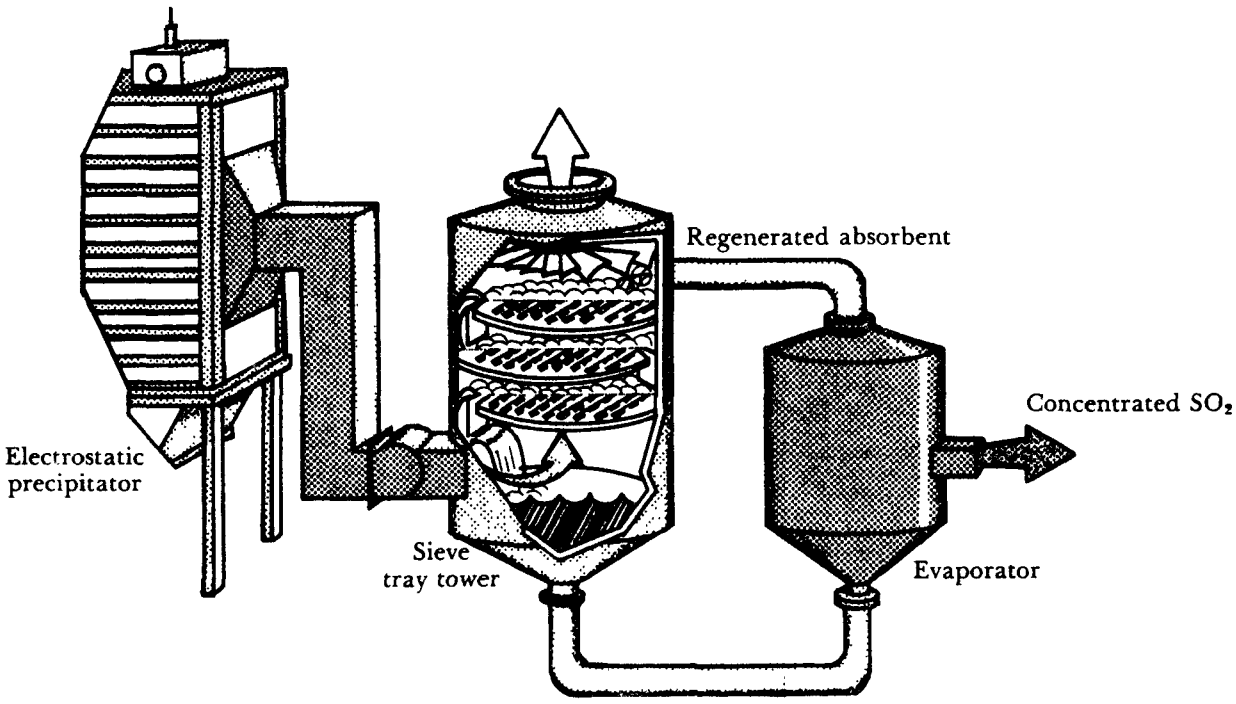


Figure 8-6. Wellman-Lord system.

The solution (sodium bisulfite), collected at the bottom of the absorber, overflows into an absorber surge tank. This solution is pumped through a filter to remove any collected particulate matter. A small side-stream is sent to a purge treatment tank where sodium sulfate is removed. The solution is then pumped to the evaporator for regeneration of the sodium sulfite solution.

The evaporator is a forced-circulation vacuum evaporator. Solution is recirculated in the evaporator where low pressure steam evaporates water from the sodium bisulfite solution. When sufficient water is removed, sodium sulfite crystals form and precipitate. Concentrated SO_2 gas (95% by volume) is removed by the steam. The sodium sulfite crystals form a slurry that is withdrawn continuously and sent to a dissolving tank, where condensate from the evaporator is used to dissolve the sodium sulfite crystals into a solution. This solution is pumped back into the top stage of the absorber (EPA, Technology Transfer Capsule Report; EPA 625/2-77-011).

The following reaction takes place in the evaporator:

$$2NaHSO_3 \rightarrow Na_2SO_3 + H_2O + \underset{\text{(concentrated)}}{SO_2\uparrow}$$

The water vapor is removed from the evaporator's overhead SO_2/H_2O vapors by water-cooled condensers. The SO_2 is compressed by a liquid ring compressor and sent to the Allied Chemical SO_2 reduction plant.

Operating Experience

The Wellman-Lord process has been installed on two 350 MW coal fired boilers at the Public Service of New Mexico San Juan power plant. This system was supplied by Davy Powergas and is similar to the system supplied on the Mitchell power plant in Indiana. The San Juan plant uses a five stage tray tower instead of the three stage tray tower of the Mitchell plant. The Wellman-Lord process has also been used to control SO_2 emissions at Claus tail gas plants, sulfuric acid plants, and on industrial boilers. A listing of selected installations is given in Table 8-5. SO_2 removal efficiency ranges from 85 to 90%, with a high of 98% on units installed in Japan (EPA, March 1978; EPA 600/7-78-032b).

Table 8-5. Wellman-Lord installations in the United States.

Company and location	Feed gas origin	Gas volume treated		Initial startup date
		(m^3/s)	(scfm)	
Units on stream:				
Olin Corporation Paulsboro, NJ	Sulfuric acid plant	20.1	43,000	1970
Standard Oil of California El Segundo, CA	Claus plant	13.4	28,000	1972
Allied Chemical Corporation Calumet, IL	Sulfuric acid plant	13.4	28,000	1973
Olin Corporation Curtis Bay, MD	Sulfuric acid plant	34.8	74,000	1973
Standard Oil of California Richmond, CA	Claus plant	13.4	28,000	1975
Standard Oil of California El Segundo, CA	Claus plant	13.4	28,000	1975
Standard Oil of California Richmond, CA	Claus plant	13.4	28,000	1976
Northern Indiana Public Service Company Gary, IN	115-MW coal fired with 80% load factor and recovery capacity	105.0	223,000	1977
Public Service Company of New Mexico Waterflow, NM	375-MW coal fired boiler system, San Juan Station No. 1	840.0	1,780,000	1978
Public Service Company of New Mexico Waterflow, NM	375-MW coal fired boiler system, San Juan Station No. 2			
Units in design or construction:*				
Getty Oil Company Delaware City, DE	60-MW mixed fuel boiler system, Delaware City No. 1	235.0	520,000	1980
Getty Oil Company Delaware City, DE	60-MW mixed fuel boiler system, Delaware City No. 2			
Getty Oil Company Delaware City, DE	60-MW mixed fuel boiler system, Delaware City No 3			
Public Service Company of New Mexico Waterflow, NM	550-MW coal fired boiler system, San Juan Station No. 3	1121 0	2,400,000	1981
Public Service Company of New Mexico Waterflow, NM	550-MW coal fired boiler system, San Juan Station No 4			

*As of February, 1979

Source: EPA, February 1979.

Magnesium Oxide Process

Magnesium oxide scrubbing is a regenerable FGD process used to remove SO_2 from combustion exhaust gas. Magnesium oxide (MgO) slurry absorbs SO_2 and forms magnesium sulfite. Magnesium sulfite solids are separated by centrifugation and dried to remove moisture. The mixture is calcined to regenerate magnesium oxide and produce concentrated SO_2 gas for production of sulfuric acid or elemental sulfur.

Particulate matter is removed from boiler exhaust by a precipitator or wet scrubber prior to entering the absorber. Magnesium oxide slurry is sprayed and absorbs SO_2 according to the following simplified reactions:

$$Mg(OH)_2 + 5H_2O + SO_2 \rightarrow MgSO_3 \cdot 6H_2O$$

$$MgSO_3 \cdot 6H_2O + SO_2 \rightarrow Mg(HSO_3)_2 + 5H_2O$$

$$Mg(HSO_3)_2 + MgO \rightarrow 2MgSO_3 + H_2O$$

$$2MgSO_3 + O_2 + 7H_2O \rightarrow 2MgSO_4 \cdot 7H_2O$$

The aqueous slurry used for scrubbing contains the hydrated crystals of MgO, $MgSO_3$ and $MgSO_4$. A continuous side stream of this recycled slurry is sent to a centrifuge where partial dewatering produces a moist cake. The liquor removed from the crystals is returned to the main slurry stream. The moist cake is dried at 350 to 450°F in a direct contact or rotary bed dryer. The dried cake is then sent to a calciner where coke is burned at very high temperatures (1250 to 1340°F) to regenerate magnesium oxide crystals according to the following reactions:

Cake dryer

$$MgSO_3 \cdot 6H_2O \xrightarrow[\text{heat}]{} MgSO_3 + 6H_2O\uparrow$$

$$MgSO_4 \cdot 7H_2O \xrightarrow[\text{heat}]{} MgSO_4 + 7H_2O\uparrow$$

MgO regeneration in calciner

$$MgSO_3 \xrightarrow[\text{heat}]{} MgO + \underset{\text{(concentrated)}}{SO_2\uparrow}$$

$$C + \tfrac{1}{2}O_2 \xrightarrow[\text{heat}]{} CO\uparrow$$

$$CO + MgSO_4 \xrightarrow[\text{heat}]{} CO_2 + MgO + \underset{\text{(concentrated)}}{SO_2\uparrow}$$

System Description

The magnesium oxide process has been installed on a few power plant boilers burning coal or oil in the U.S. A list of some installations is given in Table 8-6. A test project was installed at Boston Edison's Mystic Station, a 155 MW plant that burns oil. Since oil was burned, a low concentration particulate matter was in the flue gas and did not need to be removed before the gas entered the absorber. Magnesium oxide slurry absorbed SO_2 in a (cocurrent) venturi scrubber. SO_2 scrubbing, slurry dewatering, and magnesium sulfite drying were done at the Mystic station. Magnesium oxide regeneration and sulfuric acid production was done at the Essex Chemical plant in Rumford, Rhode Island. Regenerated magnesium oxide was returned to the Mystic station. The average SO_2 removal was approximately 92% (EPA, March 1978; EPA 600/7-78-032b). This test project was terminated in 1974. This EPA publication gives the major operating problems and solutions of three magnesium oxide scrubbing installations in the United States.

At the Philadelphia Electric Eddystone Station, Unit 1A, the MgO process was installed on a 120 MW coal fired utility boiler burning 2.5% sulfur coal. Most of the particulate matter is removed by an electrostatic precipitator; the rest by a venturi scrubber. Flue gas then enters a venturi rod scrubber where more than 90% of the SO_2 in the flue gas is removed. The MgO is regenerated in the Essex Chemical plant in New Jersey. Operating problems and solutions are given in EPA 600/7-78-058b (EPA, 1978).

Table 8-6. Magnesium oxide installations in the United States.

Station	Size (MW)	Startup date	Fuel	Particulate emission control	SO_2 emission control	SO_2 emission removal efficiency (%)	Comments
Boston Edison; Mystic station	155	9/72	Oil	–	Venturi scrubber	92	Project terminated 1974
Potomic Electric and Power; Dickerson No. 3	95	9/73	Coal (2.0% sulfur)	Electrostatic precipitator	Venturi scrubber	82-93	Project terminated 1975
Philadelphia Electric; Eddystone No. 1A	120	9/75	Coal (2.5% sulfur)	Electrostatic precipitator and venturi scrubber	Venturi rod scrubber	90	–
Philadelphia Electric, Eddystone No. 1B	240	6/80	Coal (2.5% sulfur)	ESP and venturi	Venturi rod scrubber	90	–

Citrate Process

The citrate process, a regenerable FGD process, was developed by the U.S. Bureau of Mines. The citrate process uses sodium citrate and citric acid as buffering agents to attain a higher solubility of the SO_2 in an aqueous absorbent solution. The chemistry of this process is very complex. The absorption of SO_2 is pH dependent, increasing with higher pH. SO_2 forms H_2SO_3 when absorbed by water, resulting in decreasing pH values. This creates a more acidic solution that inhibits additional

absorption of SO_2 gas. By using a buffering agent to prevent a pH drop, a substantially higher amount of SO_2 can be absorbed. One commercial installation exists at this time, on a 60 MW industrial boiler at St. Joe Minerals in Monaca, Pennsylvania. Particulate matter is removed from the flue gas by an electrostatic precipitator. Chlorides and sulfuric acid mist are removed from the flue gas by a small venturi scrubber before entering the packed tower. SO_2 is absorbed by a solution of sodium citrate, citric acid and sodium thiosulfate to produce sodium sulfite. SO_2 removal efficiency is approximately 90% (Madenburg et al., 1979). The solution (containing absorbed SO_2) is reacted with hydrogen sulfide gas (H_2S) in a closed vessel to precipitate elemental sulfur and regenerate citrate solution. The elemental sulfur precipitate is concentrated by air flotation into a sulfur slurry which is separated from the regenerated solution. Sulfur slurry is heated to form liquid sulfur and the solution is decanted. Hydrogen sulfide gas (used in regeneration) is obtained either as a by-product of petroleum refining or produced on site by reacting recovered sulfur with natural gas and steam.

The citrate process has some drawbacks. It is expensive compared to lime and limestone scrubbing. Production of H_2S from methane and sulfur can be a problem due to the shortage of natural gas. Since the size of the absorption, regeneration, and flotation equipment is large, boiler retrofitting is sometimes difficult.

Dry SO_2 Control Systems

One promising new technology for reducing SO_2 emissions from combustion sources is using dry flue gas desulfurization (FGD). In dry FGD, the flue gas containing SO_2 is contacted with an alkaline material to produce a dry waste product for disposal. This technology includes:

- injection of an alkaline slurry in a spray dryer with collection of dry particles in a baghouse or electrostatic precipitator (ESP);
- dry injection of alkaline material into the flue gas stream with collection of dry particles in a baghouse or ESP;
- addition of alkaline material to the fuel prior to combustion.

These technologies are capable of SO_2 emission reduction ranging from 60 to 90% depending on which system is used. Table 8-7 summarizes the key features of each of these technologies. These technologies have been used on boilers burning low sulfur coal (usually less than 2%) and are attractive alternatives to wet scrubbing technology, particularly in the arid western U.S.

Table 8-7. Key features of dry flue gas desulfurization systems.

Process	Sorbents used	SO_2 emission removal efficiency (%)	Particulate emission removal efficiency (%)	Development status
Spray dryer with a baghouse or ESP	Sodium carbonate Lime Limestone	60-90	99+	Three utility boilers (400-500 MW) to be started up 1981, 1982, 1983. Two industrial boilers on line.
Dry injection with a baghouse or ESP	Sodium carbonate Sodium bicarbonate Nahcolite	60-90	99+	No commercial installations planned as of 1980.
Combustion of coal/limestone mixture with a low NO_x burner	Limestone pellet Lime	75-80	99+*	EPA currently funding pilot tests on small industrial boiler.

*Note: a baghouse or ESP is used for particulate emission control.

Source: EPA, February 1980.

Spray Dryer with a Baghouse or ESP

The only commercial dry FGD installations in the U.S. at this time use a spray dryer. Alkaline is injected into a spray dryer with dry particle collection in a baghouse or ESP. Spray dryers have been used in the chemical, food processing, and mineral preparation industries over the past 40 years. Spray dryers are vessels where hot flue gases are contacted with a finely atomized wet alkaline spray. The high temperatures of the flue gas, 250 to 400°F, evaporate the moisture from the wet alkaline sprays, leaving a dry powdered product. The dry product is collected in a baghouse or ESP (Figure 8-7).

Flue gas enters the top of the spray dryer and is swirled by a fixed vane ring to cause intimate contact with the slurry spray (Figure 8-8). The slurry is atomized into extremely fine droplets by rotary atomizers. The turbulent mixing of the flue gas with the fine droplets results in rapid SO_2 absorption and evaporation of the moisture. A small portion of the hot flue gas is added to the spray-dryer-discharge duct to maintain the temperature of the gas above the dew point. Reheat prevents condensation and corrosion in the duct. Reheat also prevents bags in the baghouse from becoming plugged or caked with moist particles.

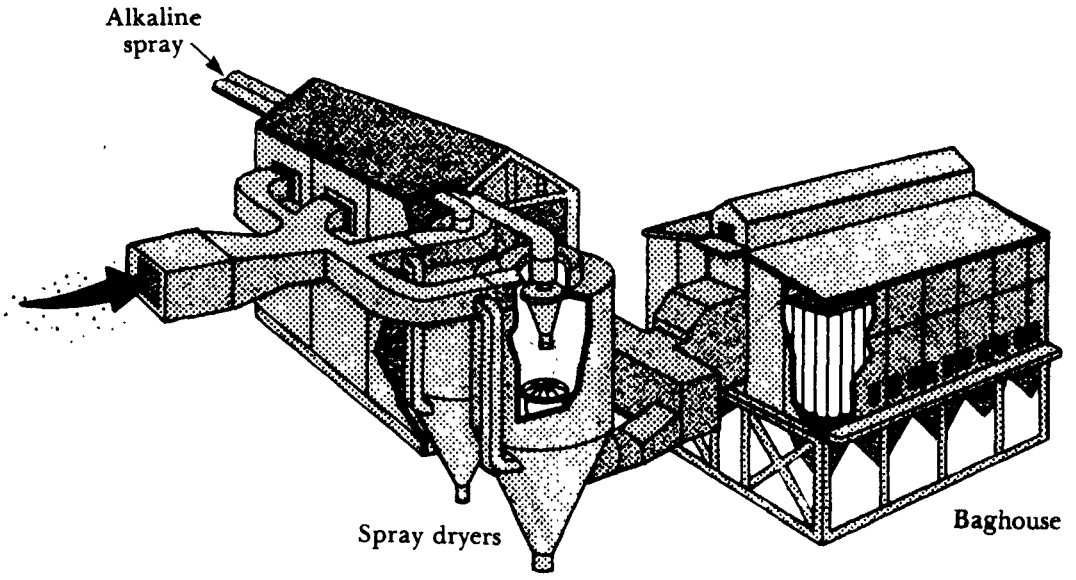


Figure 8-7. Spray dryers with baghouse.

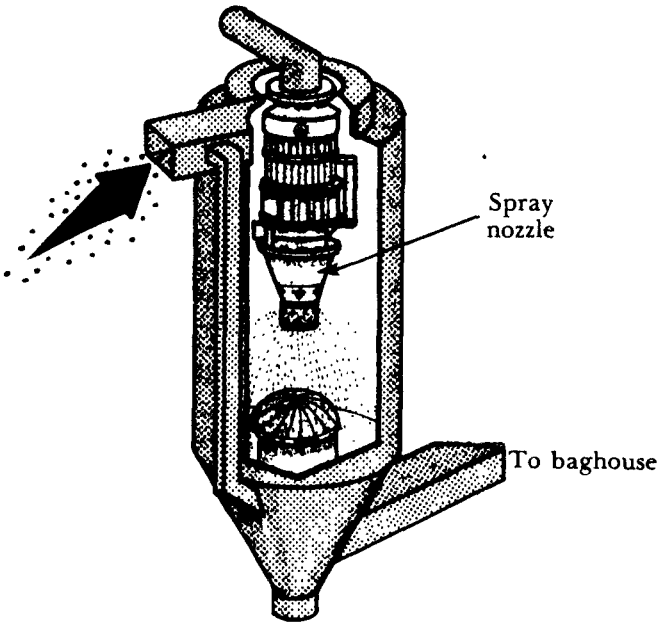


Figure 8-8. Spray dryer.

Sodium carbonate solutions and lime slurries are the most common absorbents used. A sodium carbonate solution will generally achieve a higher level of SO_2 removal than lime slurries (EPA, February 1980). When sodium carbonate is used, SO_2 removal efficiencies are approximately 75 to 90%, lime removal efficiencies are 70 to 85% (EPA, February 1980). However, vendors of dry scrubbing systems claim that their units are capable of achieving 90% SO_2 reduction using a lime slurry in a spray dryer. Lime is very popular for two reasons: lime is less expensive than sodium carbonate; sodium carbonate and SO_2 form sodium sulfite and sodium sulfate which are very soluble causing leaching problems when landfilled.

Some of the evaporated alkaline spray will fall into the bottom of the spray dryer and be recycled. The majority of the spray reacts with SO_2 in the flue gas to form powdered sulfates and sulfites. These particles, along with fly ash in the flue gas, are then collected in a baghouse or electrostatic precipitator (see 413 Student Manual, EPA 450/2-80-066). Baghouses have an advantage because unreacted alkaline material collected on the bags can react with any remaining SO_2 in the flue gas. Some process developers have reported SO_2 removal on bag surfaces on the order of 10% (Kaplan and Felsvang, 1979). However, since bags are sensitive to wetting, a 35 to 50°F margin above the saturation temperature of the flue gas must be maintained (EPA, February 1980). ESPs have the advantage of not being as sensitive to moisture as baghouses. However, SO_2 removal is not quite as efficient using ESPs.

The major differences between wet absorption SO_2 scrubbers and spray dryer systems is in the scrubbing method and the amount of moisture during the scrubbing action. In a wet scrubber, flue gas is **saturated with liquid sprays** (usually alkaline). SO_2 is absorbed by the water and also reacts with the chemical. As previously stated in Chapter 4, **absorption increases as temperature decreases**. Flue gas is cooled and saturated with the scrubbing liquid to remove SO_2. At optimum operating temperatures efficiency is usually $>90\%$.

In a spray dryer, finely atomized alkaline droplets are contacted with flue gas which is at air preheater outlet temperatures (250 to 400°F). The flue gas is humidified to within 50°F of its saturation temperature by the moisture evaporating from the alkaline slurry. Reaction of the SO_2 with the alkaline material proceeds both during and following the drying process, although to what degree is not completely understood. Since the flue gas temperature and humidity are set by air preheater outlet conditions, the amount of moisture that can be evaporated into the flue gas is also set. This means that the amount of alkaline slurry that can be evaporated in the dryer is limited by flue gas conditions (see Appendix I for use of psychrometric chart). Alkaline slurry sprayed into the dryer must be carefully controlled to avoid moisture in the flue gas from condensing in the ducting, particulate emission control equipment, or the stack. SO_2 removal efficiencies are generally $<85\%$ (EPA, February 1980).

A number of spray dryer systems have been planned for or installed on industrial and utility boilers. These are listed in Table 8-8. Spray dryers will become more popular as experience with existing units is further documented. They will be particularly useful in meeting NSPS regulations for utility boilers burning low sulfur coal that require only 70% SO_2 scrubbing.

Table 8-8. Commercial spray dryer FGD systems.

Station or plant	Size (MW)	Installation date	System description	Sorbent	Coal sulfur content (%)	SO_x emission removal efficiency (%)
Otter Tail Power Company; Coyote Station No. 1, Beulah, ND	410	6/81	Rockwell/Wheelabrator-Frye system; four spray towers in parallel with 3 atomizers in each; reverse air–shaker baghouse with dacron bags	Soda ash (sodium carbonate)	0.78	70
Basin Electric; Antelope Valley Station No. 1, Beulah, ND	440	4/82	Joy/Niro Atomizer Inc.; five spray modules with single rotary atomizer in each; reverse air baghouse with Teflon coated fiberglass bags	Lime	0.68-1.22	62-78
Basin Electric; Laramie River Station No. 3, Wheatland, WY	500	Spring 1982	Babcock and Wilcox; four spray reactors with 12 "Y-jet" nozzles in each; electrostatic precipitator	Lime	0.54-0.81	85-90
Strathmore Paper Co.; Woronco, MA	14	12/79	Mikropul; spray dryer and pulse jet baghouse	Lime	2-2.5	75
Celanese Corp.; Cumberland, MD	31	2/80	Rockwell/Wheelabrator-Frye; one spray tower followed by a baghouse	Lime	1-2	85

Source: EPA, February 1980.

Dry Injection

In dry injection systems, a dry alkaline material is injected into a flue gas stream. This is accomplished by pneumatically injecting the dry sorbent into a flue gas duct, or by precoating or continuously feeding sorbent onto a fabric filter surface. Most dry injection systems use pneumatic injection of dry alkaline material in the boiler furnace area or in the duct that precedes the ESP or baghouse. Sodium-based sorbents are used more frequently than lime. Many dry injection systems have used nahcolite, a naturally occurring mineral which is 80% sodium bicarbonate found in large reserves in Colorado. Sodium carbonate (soda ash) is also used but is not as reactive as sodium bicarbonate (EPA, February 1980). The major problem of using nahcolite is that it is not presently being mined on a commercial scale. Large investments must be made before it will be mined commercially. Other natural minerals such as raw trona have been tested; trona contains sodium bicarbonate and sodium carbonate.

Dry injection systems have been tested at a number of power stations throughout the U.S. Descriptions of these pilot systems can be found in *Survey of Dry SO_2 Control Systems* (EPA, February 1980). The major problems with dry injection systems are the low sodium utilization in the process and the disposal of leachable sodium-sulfur compounds. EPA reports that only 40 to 60% of the dry alkaline injected material is used at high SO_2 removal conditions (EPA, February 1980).

Other Dry SO_2 Processes

Coal and limestone fuel mixtures have been tested as a method for reducing SO_2 emissions. This technology using limestome pellets in a fluidized bed combustion boiler or limestone injection with a low NO_x burner was discussed earlier in this chapter. Another promising dry process, the Shell UOP process, uses a copper oxide catalyst for SO_2 emission reduction. This process was discussed in detail in Chapter 7 of this manual.

Comparison of FGD Systems

Comparisons between dry and wet FGD systems can be made in four major areas: waste disposal, chemical reagent requirements, SO_2 removal efficiencies, and economics. A comparison of the FGD systems covered in this manual will follow.

Lime, limestone, double alkali, and dry FGD systems produce a sludge that must be disposed of properly. New installations must meet all solid waste regulations including stringent RCRA regulations. The regenerable FGD processes—including Wellman-Lord, magnesium oxide, and citrate—generate a usable product (sulfur or sulfuric acid) that has a commercial value. The dry FGD systems produce a dry waste product that can be discarded using conventional fly ash removal systems instead of sludge removal systems. However, sodium-based dry FGD systems are undesirable because they produce waste products that easily leach from conventional landfills.

Lime and limestone are cheaper than sodium based absorbents and are readily available. In dry scrubbing systems, a higher stoichiometric ratio of lime is necessary to achieve SO_2 reduction efficiencies similar to wet lime and limestone scrubbing efficiencies. Consequently, costs for lime or limestone materials will be somewhat higher in dry systems.

Regarding SO_2 removal efficiencies, all of the wet scrubbing processes are capable of at least 90% removal efficiency. Most units are capable of 95% removal efficiency. These systems have been installed and operating for a number of years with good reliability. Dry scrubbing is capable of removing at least 75 to 85% SO_2. Vendors of dry FGD equipment claim 90% capabilities. Dry FGD systems are capable of meeting the NSPS regulation (1978) of 70% scrubbing when SO_2 emissions are less than 0.6 lb/10^6 Btu. This technology is very attractive for utility boilers burning low sulfur western coal. With improvements, dry scrubbing may be used on boilers burning medium sulfur coal (2 to 3%) in the near future.

• *Cost Comparisons*

Cost estimates for FGD systems have been made by various groups including the EPA, Tennessee Valley Authority (TVA), and the Electric Power Research Institute (EPRI). Two articles in the *Journal of the Air Pollution Control Association* (JAPCA, April and May 1981) summarize FGD systems on an environmental and cost basis. Cost estimates of selected FGD systems are given in Table 8-9. This table was constructed using cost estimate data from a paper presented at the FGD symposium in Houston, Texas, October, 1980 (McGlamery et al., 1980). The paper presented cost estimates of FGD systems based on 1979 capital costs and 1980 operating costs. The authors of this paper also worked up a cost estimate of a limestone scrubber based on 1982 capital costs and 1984 operating costs. As can be seen from Table 8-9, the cost of a 500 MW power plant burning 3.5% sulfur coal, meeting the 1.2 lb $SO_2/10^6$ Btu NSPS requirement, was 98 $/kW for capital cost and 4.02 mills/kWh operating costs.

Table 8-9. Cost estimates of selected FGD systems.*

FGD system	Mid-1979 capital investment ($/kW)	Mid-1980 first year operating cost (mills/kWh)
Limestone	98	4.02
Lime	90	4.25
Double alkali	101	4.19
Wellman-Lord	131-141	5.11-6.03
Magnesium oxide	132	5.08
Citrate	143	6.44

*Cost basis: 500 MW power plant burning 3.5% sulfur coal achieving 90% SO_2 removal efficiency.

Source: McGlamery, 1980.

The costs of a limestone scrubber on a similar power plant with the following **new premises** was also calculated:

- meeting 1978 NSPS requirements
- spare absorber and emergency bypass
- ID booster fans instead of FD fans
- ponding and landfill estimates to meet RCRA requirements
- forced oxidation of sludge
- use of a spray tower instead of a mobile bed scrubber
- changes in economic premise including 1982 capital investment and 1984 operating costs

The cost of a limestone scrubber under this new premise is 194 $/kW capital cost and 9.9 mills/kWh operating costs. These cost estimates are almost **double** those given in Table 8-9. Therefore, the cost data given in Table 8-9 should be used for **comparison** of FGD systems **only**.

Over the past 15 years, a wealth of material has been written and documented concerning FGD control technology. The authors of this manual suggest that the readers turn to the many publications from EPA-IERL concerning this subject, particularly the proceedings from the annual FGD symposiums sponsored by the EPA.

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Chapter 9

Industrial Exhaust Systems

Hoods

Control of all air pollutants begins at their point of capture, the hood. The overall control efficiency that can be achieved is a function of the capture efficiency of the hood. For example, if a poorly designed hood captures only 70% of the emissions generated, then the overall efficiency of the system can never be greater than 70% even if the control device is 100% efficient. Any pollutant can be captured effectively with an appropriately sized hood and a fan with enough power to supply the suction. Hoods are designed, however, to achieve the maximum pollutant capture efficiency with a minimum of air flow and power consumption.

Hoods are generally classified according to their shape and the manner in which they capture the contaminant. Hoods are referred to as **enclosure** or **nonenclosure** depending on the manner in which they capture the pollutant. Some common hood shapes are the *slot, plain opening, booth,* and *canopy* (Figure 9-1). Hood entry coefficients will be discussed later in this chapter. Enclosure hoods may totally or only partially enclose the source of emissions. The term "enclosure" is somewhat of a misnomer since some of these hoods can be located close to the point of emissions so that capture occurs before dispersion of the contaminant can take place. For example, in many paint spraying operations a booth is used to capture the emissions generated. Occasionally the painting is done inside the booth or it may be done a few feet in front of the booth. Both of these operations would be an example of an enclosure hood.

Nonenclosure hoods refer to hoods that are usually placed around the perimeter of the contaminant-generating source. The most common nonenclosure hood used to capture gaseous emissions is the *lip* or *rim* exhaust located on open surface tanks.

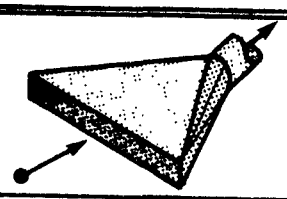
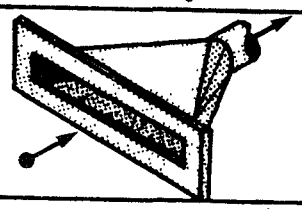
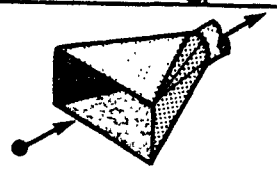
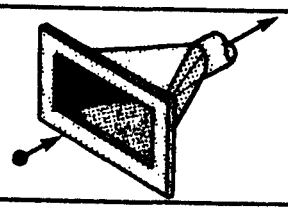
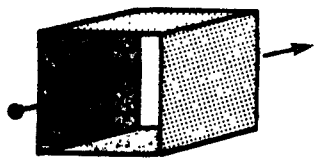
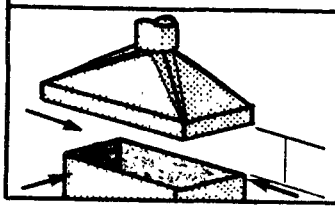
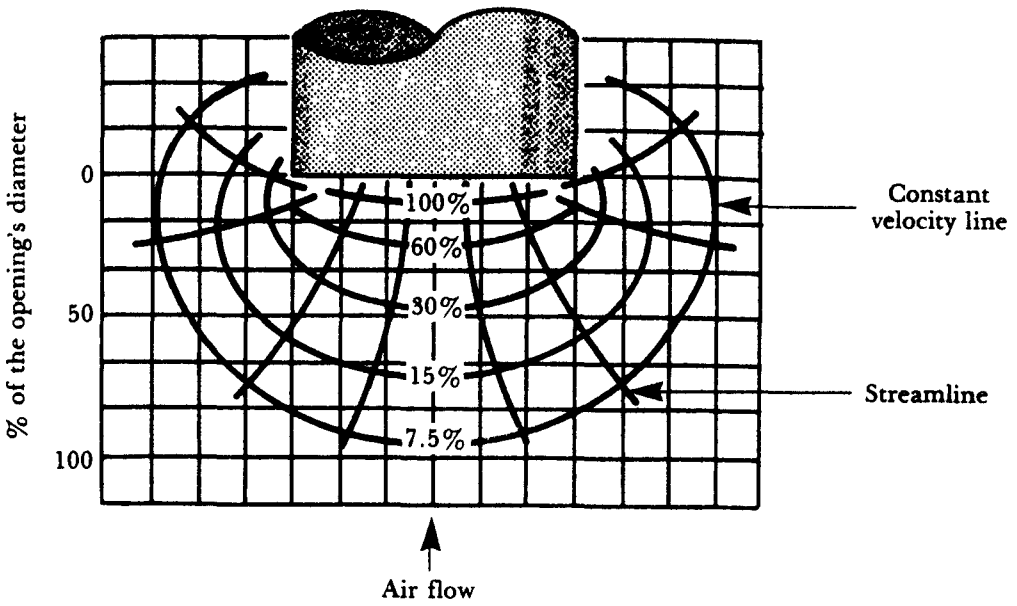
Hood	Description	Entry coefficient (C_e)
	Slot	0.60
	Flanged slot	0.82
	Plain opening	0.72
	Flanged opening	0.82
	Booth	0.82
	Canopy	0.82

Figure 9-1. Typical hoods and their entry coefficients.

Hood Design

The air volume or suction required for a hood depends on: hood shape, hood type, hood size, required capture velocity, distance of the hood from the source of pollutants, and temperature of the contaminant exhaust stream. Due to the numerous variables, equations for predicting the required exhaust volume are based on empirical data for a particular type of hood.

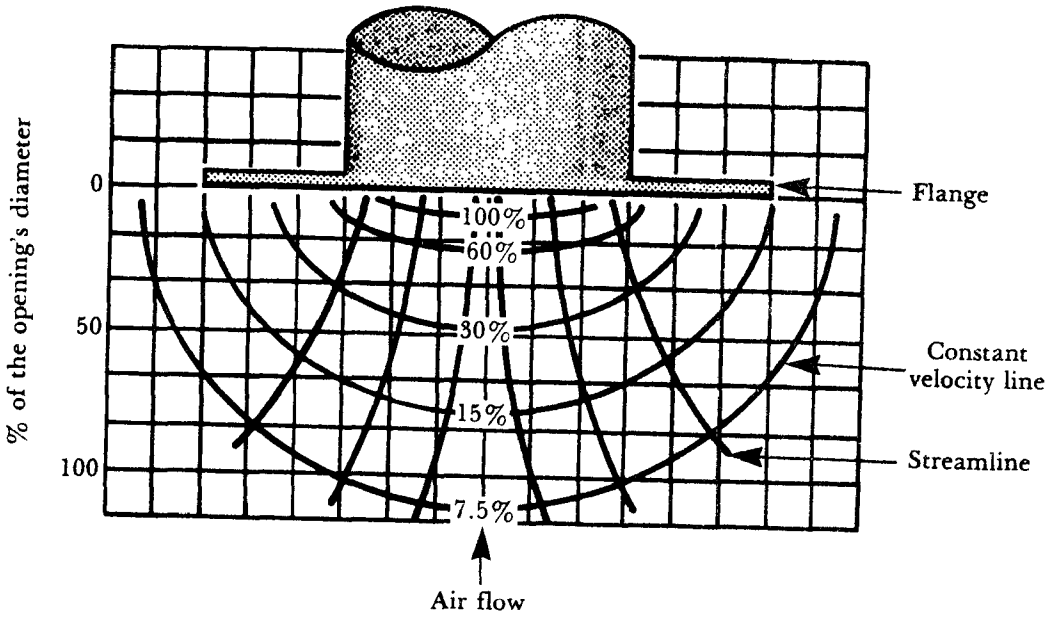
The basis for hood design equations starts with the simplest case—air flowing into a circular hood or duct. From studies done by Dalle Valle and others, the actual flow pattern of air being sucked into the hood is illustrated in Figure 9-2. The curved radial lines in Figure 9-2 are referred to as contour lines and represent lines of constant velocity. The lines perpendicular to the duct are streamlines and represent the direction of flow.



Source: *Industrial Ventilation*, 1976.

Figure 9-2. Flow diagram of air being pulled into a plain opening.

Two important generalizations are illustrated in Figure 9-2. First, a rapid decrease in velocity occurs with increased distance from the hood face. Therefore, the closer the hood is to the source (or by totally enclosing the source), the more efficient and effective the capture of pollutants. Secondly, air is drawn into the openings from all directions, meaning that more air is drawn in than required. Flanges can be used to eliminate air pull from areas where no contaminants are present. Figure 9-3 illustrates the effect adding flanges has on the contour and streamlines. For most cases it is estimated that flanges can reduce air requirements by 25% and need never extend more than 6 inches beyond the hood entrance (*Industrial Ventilation*, 1976).



Source: *Industrial Ventilation*, 1976.

Figure 9-3. Effect of flange on air flow into a plain opening.

The following design equations for hood systems are categorized according to the shape of hood. All equations are for **cold** contaminant exhaust streams unless otherwise stated. A more extensive review of hood types and their design equations is given in *Industrial Ventilation* (1976).

Total Enclosure Hood

For any totally enclosed hood or for emissions generated inside a booth, the air flow into the hood can be computed by:

(Eq. 9-1)

$$Q = vA$$

Where: Q = air flow (into hood face), cfm
v = air velocity required to capture the pollutant, fpm
A = total open area of the hood, ft^2*

The capture velocity is the velocity necessary (at any point) to overcome opposing air currents and to capture the air contaminants. Table 9-1a presents a range of capture velocities for certain processes (*Industrial Ventilation*, 1976).

*The units in this chapter are expressed in English rather than metric, since most published tables and charts are in English units.

Table 9-1a. Range of capture velocities and duct velocities.

Condition of dispersion of contaminant	Examples	Capture velocity (fpm)
Released with practically no velocity into quiet air	Evaporation from tanks; degreasing, etc.	50-100
Released at low velocity into moderately still air	Spray booths; intermittent container filling; low speed conveyor transfers; welding; plating; pickling	100-200
Active generation into zone of rapid air motion	Spray painting in shallow booths; barrel filling; conveyor loading; crushers	200-500
Released at high initial velocity into zone of very rapid air motion	Grinding; abrasive blasting, tumbling	500-2000

In each category above, a range of capture velocity is shown. The proper choice of values depends on several factors:

Lower end of range

1. Room air currents minimal or favorable for capture.
2. Contaminants of low toxicity or of nuisance value only.
3. Intermittent, low production.
4. Large hood—large air mass in motion.

Upper end of range

1. Disturbing room air currents.
2. Contaminants of high toxicity.
3. High production, heavy use.
4. Small hood—local control only.

Source: Permission granted by Committee on Industrial Ventilation, American Conference of Governmental Industrial Hygienists, Box 16153, Lansing, MI 48901. Taken from 14th Edition (1976) of *Industrial Ventilation.*

Table 9-1b. Recommended minimum duct velocities.

Nature of contaminant	Examples	Duct velocity (fpm)
Gases, vapors, smokes, fumes, and very light dusts	All vapors, gases, and smokes; zinc and aluminum oxide fumes; wood, flour, and cotton lint	2000
Medium-density dry dust	Buffing lint; sawdust; grain, rubber, and plastic dust	3000
Average industrial dust	Sandblast and grinding dust, wood shavings, cement dust	4000
Heavy dusts	Lead and foundry shake-out dusts; metal turnings	5000

Source: EPA, 1973.

Free Standing Plain Hood

For a *free standing* or *unbounded* hood, Equation 9-1 must be modified to account for the distance a source is from the hood. Dalle Valle developed Equation 9-2 for air flowing into a plain, circular duct or a rectangular hood that is essentially square.

(Eq. 9-2)

$$Q = v_x(10x^2 + A)$$

Where: Q = air flow, cfm
v_x = air velocity required for capture at some distance (x) from hood opening, fpm
x = distance of contaminant source from the center of the hood face, ft
A = area of hood face, ft^2

Equation 9-2 was developed from Figure 9-2 and it also shows the rapid increase in air volume required for capture (due to decrease in velocity) as the source gets farther from the hood. The volume increases in relation to the distance squared.

Equation 9-2 applies to a free standing or unobstructed hood. For a square or rectangular hood that is bounded on one side by a flat surface (the floor or wall) Equation 9-2 then becomes:

(Eq. 9-3)

$$Q = v_x\left(\frac{10x^2 + 2A}{2}\right)$$

Example 9-1

Determine the total volume of air entering a paint spray booth that is 10 ft wide and 7 ft high. Some objects are painted outside the booth but never at a distance of more than 3 ft. Room air currents are kept low.

Solution:

The paint spray booth would be a plain opening bounded on one side by a flat surface (the floor). Therefore, Equation 9-3 applies.

$$Q = v_x\left(\frac{10x^2 + 2A}{2}\right)$$

The capture velocity can be estimated from Table 9-1a. For paint spray emitted into a draftless area the capture velocity should be 100 fpm.

Substituting into Equation 9-3:

$$Q = 100 \text{ fpm}\left[\frac{(10)(3 \text{ ft})^2 + 2(7 \text{ ft})(10 \text{ ft})}{2}\right]$$

$$= 11{,}500 \text{ cfm}$$

An important point to note is that in order to capture the pollutant the velocity at 3 ft from the hood face must be at least 100 fpm. If painting was done farther away from the hood face, the velocity still must be the same. The volume of air would then greatly increase.

For example, if painting is done at 5 ft instead of 3 ft from the hood, then this raises the required air volume pulled into the system to 19,500 cfm.

A common way of expressing control volume is based on the square feet of hood opening. For example:

$$\frac{Q}{A} = \frac{11{,}500 \text{ cfm}}{(7 \text{ ft})(10 \text{ ft})}$$

$$= \frac{164.3 \text{ cfm of air}}{\text{sq ft of hood opening}}$$

Canopy Hood: Cold Processes

Canopy hoods have three or four open sides and are almost identical to plain free-standing hoods. The major difference is that canopies are used over the top of open tanks. The canopy portion of the hood is normally extended beyond the edge of the tank to minimize any side draft disturbances. The air volume for an open canopy hood capturing emissions from a **cold** process is:

(Eq. 9-4) $$Q = 1.4\ h_p l v$$

Where: Q = air flow of an open canopy hood, cfm
h_p = canopy perimeter excluding closed sides, ft
l = vertical distance between canopy and tank hood, ft
v = control velocity, fpm

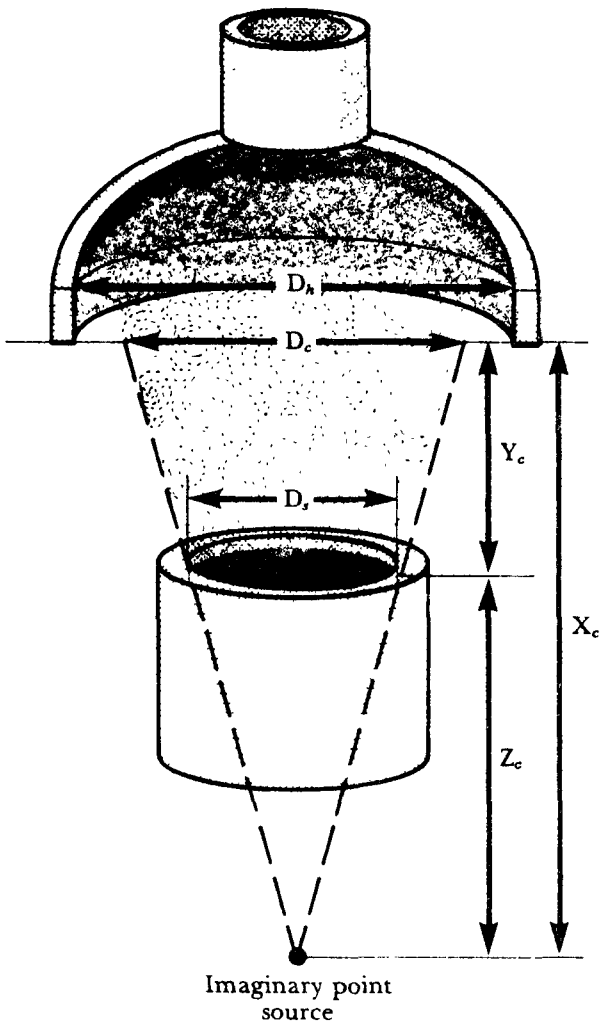
Canopy Hood: Hot Processes

The design of hoods to capture gaseous emissions from a hot process is markedly different from that for a cold process. Expansion of the hot gases due to the thermal effect must be taken into account. As the hot gas rises, it mixes turbulently with the surrounding air. The higher the gas column rises, the larger it becomes, and the more it is diluted with ambient air. Equation 9-5 is an empirical relationship developed by Hemeon (1955) for predicting the expansion of the rising plume for *high canopy* hoods (located more than 3 ft above the heat source):

(Eq. 9-5) $$D_c = 0.5 X_c^{0.88}$$

Where: D_c = diameter of the hot column of air at the level of the hood face, ft
X_c = distance from the hypothetical point source to the food face, ft

X_c is defined in Figure 9-4 as the sum of the distance from the hot source to the hood face, Y_c, plus the distance from the top surface of the heat source to a hypothetical vortex of the fume column, Z.



Source: Hemeon, 1955.

Figure 9-4. Dimensions used in designing high canopy hoods controlling emissions from hot sources.

Values of Z_c can be found from:

(Eq. 9-6) $$Z_c = (2D_s)^{1.138}$$

Where: D_s = width of the hot source
Z_c = distance from top of hot source to hypothetical fume vortex, ft

The actual diameter of the hood opening D_h must extend beyond the rising column of hot gases to ensure complete capture. Most designers increase the hood diameter by a factor of 0.8 Y_c (*Industrial Ventilation*, 1976). Therefore, diameter of the canopy is given by:

(Eq. 9-7)

$$D_h = D_c + 0.8Y_c$$

The volume of gas entering the canopy hood includes some ambient air and is given by:

(Eq. 9-8)

$$Q = vA_c + v_r(A - A_c)$$

Where: Q = total volume of air entering the hood, cfm
v = velocity of the rising gases at the hood face, fpm
A_c = area of the rising column of gas at the hood face, ft^2
v_r = required velocity through the remaining area of the hood, $A - A_c$, fpm
A = total area of hood face, ft^2

The value of v_r depends on the draft in the hood area, height of the hood above the source, and degree of capture required. Values of v_r usually range between 100 and 200 fpm, with higher values giving higher pollutant capture efficiencies.

The control velocity for capturing emissions from a flat, horizontal heat source is given by:

(Eq. 9-9)

$$v = \frac{8(A_s)^{0.33}(\Delta T)^{0.42}}{X_c^{0.25}}$$

Where: v = control velocity, fpm
A_s = top surface area of hot source, ft^2
ΔT = temperature difference between the hot source and ambient, °F
X_c = distance from hood opening to imaginary point source, ft

Control velocities for other, irregularly shaped hot sources can be found in Cheremisinoff (1976).

It is important to note that Equations 9-6 through 9-9 are for *high circular canopy hoods* controlling emissions from a flat, horizontal hot source. Equations for low canopy hoods are listed in Table 9-2.

Table 9-2. Control volumes for low canopy hoods.

Shape of heat source	Control volume
Circular	$Q = 4.7(D_h)^{2.33}(\Delta T)^{0.42}$
Rectangular	$Q = 6.2(w \cdot l_h)^{1.33}(\Delta T)^{0.42}$

Where: D_h = diameter of hood, ft
l_h = length of hood, ft
w = width of hood, ft
ΔT = difference between the temperature of the heat source and ambient, °F

Example 9-2

A zinc-melting pot 5 ft in diameter contains molten metal at 800°F. A high canopy hood is used to capture emissions and is located 8 ft above the pot. Assuming ambient air temperature to be 80°F and using v_r as 200 fpm, calculate the hood size and exhaust rate.

Solution:

1. To calculate the hood size, first find the distance from the top of the pot surface down to the hypothetical point source, Z_c, from Equation 9-7.

$$Z_c = (2D_s)^{1.138}$$
$$= (2 \times 5 \text{ ft})^{1.138}$$
$$= 13.7 \text{ ft}$$

The total height X_c is given by:

$$X_c = Z_c + Y_c$$
$$= 13.7 + 8$$
$$= 21.7 \text{ ft}$$

The diameter of the rising column of gases at the hood face is given by Equation 9-5.

$$D_c = 0.5X_c^{0.88}$$
$$= 0.5(21.7)^{0.88}$$
$$= 7.5 \text{ ft}$$

The hood size required, including increase for the rising hot gas is given by Equation 9-7.

$$D_h = D_c + 0.8Y_c$$
$$= 7.5 \text{ ft} + 0.8(8)$$
$$= 7.5 \text{ ft} + 6.4 \text{ ft}$$
$$= 13.9 \text{ ft}$$

2. To calculate the hood exhaust rate, use Equation 9-8.

   To use Equation 9-8, first compute the area of the melting pot and the velocity of rising gas at hood face.

   Area of hot source:

$$A_s = \frac{\pi}{4} D_s^2$$

$$= \frac{\pi}{4}(5)^2$$

$$= 19.6 \text{ ft}$$

   The velocity is computed from Equation 9-9.

$$v = \frac{8(A_s)^{0.33}(\Delta T)^{0.42}}{X_c^{0.25}}$$

$$= \frac{8(19.6)^{0.33}(800 - 80)^{0.42}}{(21.7)^{0.25}}$$

$$= 157 \text{ fpm}$$

   The area of the rising column of gas is:

$$A_c = \frac{\pi}{4} D_c^2$$

$$= \frac{\pi}{4}(7.5)^2$$

$$= 44.2 \text{ ft}^2$$

   The area of the hood face is:

$$A = \frac{\pi}{4} D_h^2$$

$$= \frac{\pi}{4}(13.9)^2$$

$$= 151.7 \text{ ft}^2$$

   Substituting into Equation 9-8 for the total volume of air:

$$Q = vA_c + v_r(A - A_c)$$

$$= 157(44.2) + 200(151.7 - 44.2)$$

$$= 28{,}440 \text{ cfm}$$

Nonenclosure Lip Hood

One of the most common types of nonenclosure hoods for controlling vapors from open surface tanks is the lip hood pictured in Figure 9-5 (these hoods are sometimes called slot hoods). These hoods are rarely used with hot gases. The lips are designed for a velocity of approximately 200 fpm through the lip face at the required ventilation rate (EPA, 1973).

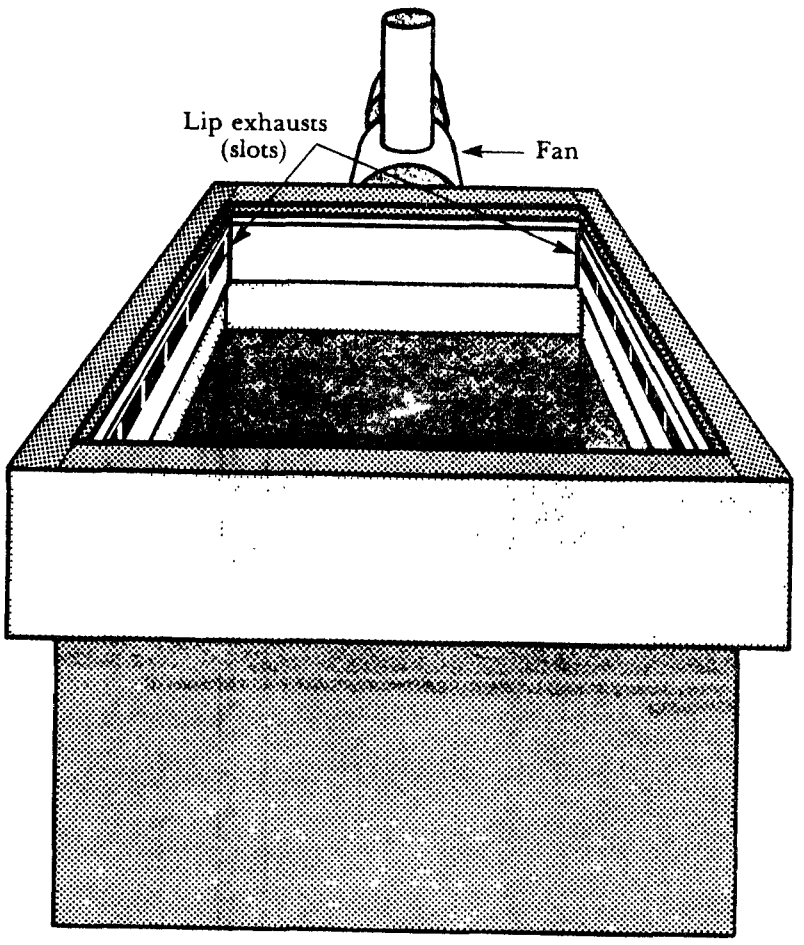
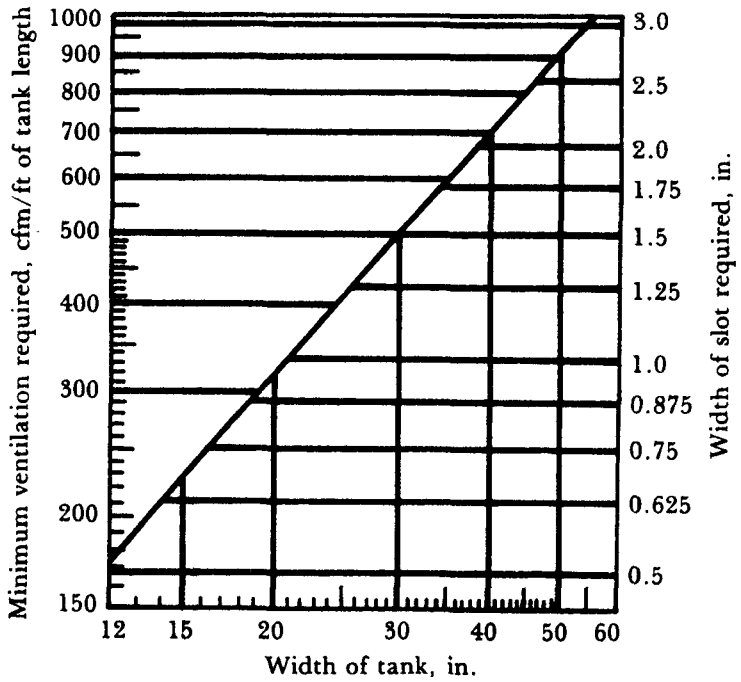


Figure 9-5. Open top tank degreaser with lip exhausts.

For a tank with two parallel lip hoods, the ventilation rate required and the lip width may be estimated from Figure 9-6. If a lip hood is used on only **one** side of a tank and the opposite side of the tank is **bounded** by a vertical wall, Figure 9-6 can be used to estimate the ventilation rate. The procedure, as outlined by the EPA (1973), is to double the tank width and read a ventilation rate from Figure 9-6. The actual ventilation rate is estimated as one-half the value read from the graph. The following example illustrates the use of Figure 9-6.



Source: EPA, 1973.

Figure 9-6. Minimum ventilation rates for lip exhaust hoods.

Example 9-3

A degreasing tank 6 ft long and 4 ft wide is controlled by using parallel lip hoods along each of the 6 ft lengths. Determine the total exhaust rate required and the width of the lips.

Solution:

From Figure 9-6, the ventilation rate can be read as:

860 cfm per ft of tank length

The exhaust volume is:

$$Q = Q_v l_t$$

Where: Q = exhaust volume, cfm
Q_v = minimum ventilation rate per ft of tank length, cfm/ft
l_t = tank length, ft

Then: $$Q = 860 \frac{\text{cfm}}{\text{ft}} \times 6 \text{ ft}$$

$$= 5160 \text{ cfm}$$

Also, from Figure 9-6, the lip width is approximately 2.6 inches.

Hood Entry Coefficients

When air flows into a hood or duct, pressure losses occur which result in a decreased flow rate. These losses are due to turbulence and are dependent on the shape of the hood or duct. The coefficient of entry (C_e) indicates the extent of these losses. For example, in a theoretically perfect hood with no turbulence loss, $C_e = 1.0$. Figure 9-1 lists the entry coefficients for some typical hoods. The coefficient of entry is used to determine the actual flow rate and pressure after the air enters the hood.

Ducts

Air Flow in Ductwork

In describing air flow through a system, a number of pressure terms are used. The three most important are: *static pressure, velocity pressure,* and *total pressure*. Air flowing in a pipe is acted on by static and velocity pressure, which when added together give the total pressure (Figure 9-7).

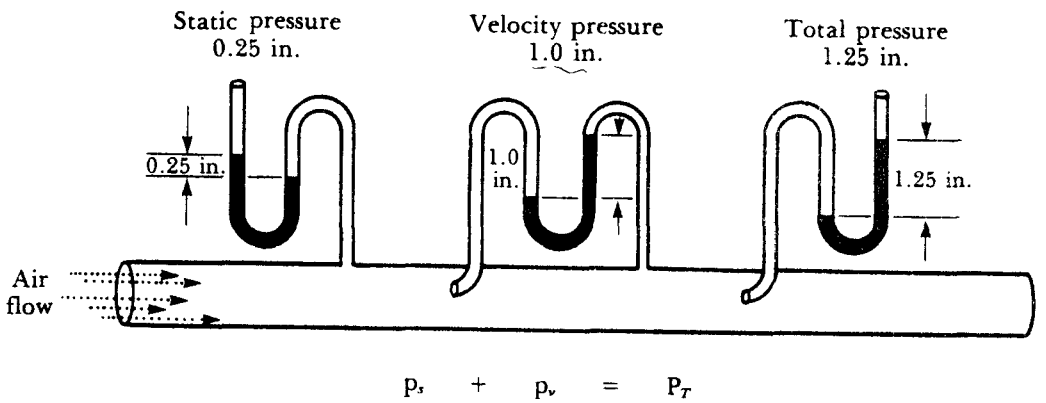


Figure 9-7. Pressure terms used to describe air flow in a duct (measured in inches).

Any gas within a confined enclosure will have a static pressure, whether or not the gas is in motion. Static pressure, p_s, acts in all directions and is independent of the velocity of the gas. Therefore, static pressure is measured at right angles to the direction of air flow to avoid influence from the air velocity. In duct design, the friction losses in a pipe can sometimes be measured by the difference in static pressure (no change in direction or velocity). Therefore, the static pressure is sometimes called the *frictional* or *resistance* pressure. Static pressure can be positive or negative, compared to the **local** atmosphere.

The velocity pressure, p_v, is the pressure created by air traveling at a specific velocity. Velocity pressure acts only in the direction of flow, therefore it is always positive. The velocity pressure is obtained by measuring the total pressure and subtracting out the static pressure. There is a basic relationship between the velocity of

P_v → Airway +
P_s → + or −

a gas and the velocity pressure. For an air stream with a density of 0.075 lbs/ft³ (standard conditions) the air velocity can be computed from the velocity pressure using the following equation (*Industrial Ventilation*, 1976):

(Eq. 9-10)

$$\boxed{v = 4005\sqrt{p_v}}$$

Where: v = air velocity, fpm
p_v = velocity pressure, in. H_2O

$T = 530°R$
$M_s = 29$ lb/lb·mole
$P_s = 29.92$ in. Hg.

Equation 9-10 is derived by applying Bernoulli's equation for air flow in a duct. (The derivation of the equation is given in APTI Course 450 Student Manual, Appendix C.) Equation 9-10 is used extensively in stack sampling, where it is commonly known as the pitot tube equation. The common form of the pitot tube equation is:

(Eq. 9-11)

$$\boxed{v = 85.486 \text{ ft/sec} \sqrt{\frac{T_s \Delta p}{M_s P_s}}}$$ → stack

Where: v = gas velocity, ft/sec
T_s = temperature of the stack, °R
M_s = molecular weight of the stack gas, lb/lb mol
P_s = absolute pressure of the stack, in. Hg
Δp = velocity pressure, in. H_2O

The above equation equals Equation 9-10 when $T_s = 530$°R, $M_s = 29$ lb/lb mol and $P_s = 29.92$ in. Hg. Note that in stack sampling the velocity pressure is symbolized by Δp, while in industrial ventilation it is p_v.

The sum of the static and the velocity pressure at a given point in the duct is the total pressure, P_T (Figure 9-7). Total pressure is measured in the direction of the flow. Therefore, the total pressure is sometimes referred to as the impact or dynamic pressure. The total pressure should not be confused with the absolute pressure. The absolute pressure is the sum of the atmospheric pressure plus the pressure of the system (Chapter 2). The system pressure can be either positive or negative.

Air Flow Resistances or Losses in a Duct

Pressure losses from air flowing in a pipe are due to friction and turbulence. Static or friction pressure losses are caused by air abrading or rubbing against the sides of the duct. Turbulent losses are due to rapid changes in direction or velocity. A change in duct cross sectional area will change the velocity of the gas in the duct. The sum of the friction and turbulent losses over a specific length of pipe is termed the *pressure drop*. The pressure drop of a system is determined by measuring the difference in total pressure at two points in the system.

The friction and dynamic losses can be further subdivided into five categories. These losses are:

Inertial losses: Energy required to accelerate a volume of air from rest. Inertia losses are essentially the same as the velocity pressure.

Orifice losses: Pressure losses at the hood or duct entrance due to turbulence. Orifice losses are dependent on the shape of the opening and are measured by the entry coefficient, C_e.

Straight duct friction losses: Pressure losses due to air rubbing along the sides of the duct. Many charts and tables have been developed that give the friction losses in straight ducts. Figure 9-8 plots the friction losses in inches of water per 100 feet of duct versus air volume with duct diameter and velocity as parameters. If any two of the above quantities are known, the other two can be read from the chart.

Elbow and branch entry losses: Pressure losses due to change in the streamlines of an air stream moving through an elbow or entering a branched connection. These losses are computed as equivalent feet of straight duct that will have the same pressure loss. Equivalent lengths of elbows and entries are given in Table 9-3 (*Industrial Ventilation*, 1976).

Contraction and expansion losses: For gradual *contractions* in the cross sectional area of a duct, the pressure losses are small (Alden, 1970). Abrupt contractions are rare in well-designed exhaust systems except as outlets from chambers. For duct *expansions*, abrupt enlargements result in a greater pressure loss.

Checking the Design of an Exhaust System

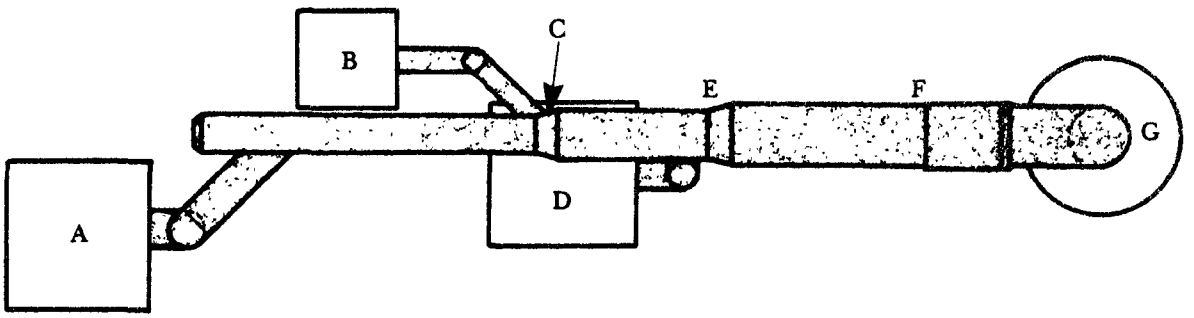
A number of methods and general rules of practice are used in the initial design of an exhaust system. These procedures are covered in detail in *Industrial Ventilation* (1976). This section will concentrate on a procedure to check the flows and pressures of an existing exhaust system. The objectives of checking an exhaust system are (EPA, 1973):

1. to determine the exhaust volume and suction velocity at each point and evaluate the adequacy of contaminant pickup.
2. to determine the system's static and total pressure; and evaluate fan capacity, speed, and horsepower.
3. to determine the total exhaust volume and evaluate the size and performance of the control device.

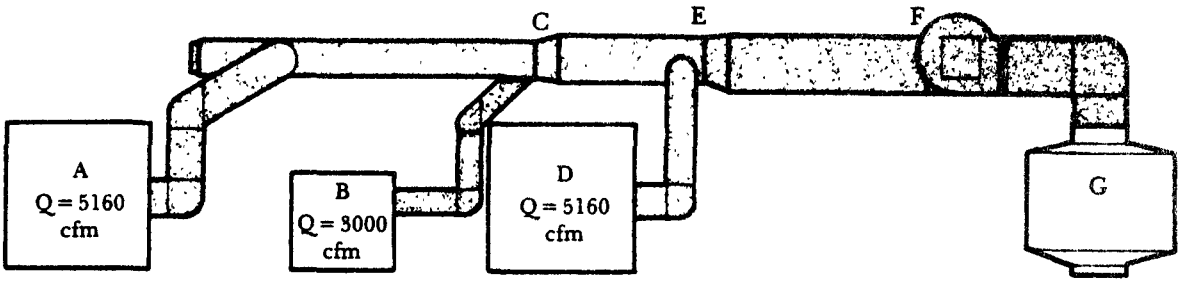
If all calculations are put in tabular form, a systematic and orderly procedure can be followed, whether designing a new system or checking the design of an existing system.

The following illustrative problem is based on a procedure outlined by Alden (1970) for duct design. Flow calculations take place in two steps. The first, Table 9-4a, establishes the minimum pressure in the main section of the duct that will induce the desired flows in each branch. This is done by computing the pressure at each junction of the main at a branch. This pressure includes both hood entry and friction losses in the branch. These values in Table 9-4a are only trial figures and are used as the starting point for the second set of calculations, Table 9-4b. The second set of calculations determines the proper pressures and flow rates.

Top view



Side view



Section	Length, ft	Diameter, in.
AC	20	18
BC	10	18
CE	6	22
DE	8	16
EF	10	26
FG	10	26

Figure 9-9. Layout for Example 9-4.

Solution:

The following notes refer to the column and line numbers in Table 9-4a.

Column 1	Self explanatory.
Column 2	Self explanatory.
Column 3	Required exhaust volume pickup at each hood. These values are given in the problem statement. If they are not supplied, values can be obtained from Table 9-1 or similar tables in other references.
Column 4	Obtained from sketch of plant, Figure 9-9.
Column 5	Determined by dividing the volumetric flow (Column 3) by the area of the duct. Note that all velocities must be greater than 2000 fpm (from Table 9-1).
Column 6	Obtained from the sketch in Figure 9-9.
Column 7	Equivalent length is obtained from Table 9-3 for the given elbow angles.
Column 8	Equivalent length is obtained from Table 9-3 for the given branch entry angles.
Column 9	Sum of Columns 6, 7 and 8.
Column 10	Read from Figure 9-8, the friction chart.
Column 11	Column 9 multiplied by 10 divided by 100. This gives the pressure drop due to friction in each branch.
Column 12	Obtained from Figure 9-1.
Column 13	Negative pressure at the hood required to produce the exhaust volume (Column 3) at this given set of conditions. Column 13 is computed using Equation 9-10.
Column 14	Sum of Columns 11 and 13. This is the minimum negative suction pressure, at the junction of the main, which will produce the proposed flow in the branch.

Table 9-4a. Exhaust system calculations (trial figures).

Col →	1	2	3	4	5	6	7			8		9	10	11	12	13	14
Line # ↓	Branch	Source	Proposed conditions			Straight pipe, l (ft)	Elbow			Branch		Total pipe, l Col 6 + Col 7 + Col 8 (ft)	Friction		Hood entry coefficient, C_e	Tentative pressure	
			Q (cfm)	d (inches)	v in duct (fpm)		No.	Angle (degree)	Eq. l (ft)	Angle (degree)	Eq. l (ft)		(in. H_2O per 100 ft section)	In the branch $\frac{\text{Col 9} \times \text{Col 10}}{100}$ (in. H_2O)		At the hood $\left(\frac{\text{Col 5}}{4005\ \text{Col 12}}\right)^2$ (in. H_2O)	At junction of branch and main Col 11 + Col 13 (in. H_2O)
1	AC	Open tank	5160	18	2920	20	1 1	90 60	47	45	28	95	0 90	0.86	0 60	1.48	2.34
2	BC	Open tank	3000	13	3255	10	1 1	45 90	29	45	19	58	1.50	0 87	0.60	1.85	2.72
3	DE	Open tank	5160	16	3696	8	1	90	24	30	16	48	1.60	0.77	0.60	2 37	3.14

Note. all pressures negative

The following notes refer to the column and line numbers in Table 9-4b.

Column 15	Self explanatory.
Column 16	Self explanatory.
Column 17	Transferred from Column 14 of Table 9-4a.
Column 18 Line 1	Pressure drop across branch BC. Calculation starts at the hood for which the pressure drop from hood to fan inlet is the greatest (usually this is the farthest hood from the fan).
Column 18 Line 2	Pressure drop across section AC which joins branch BC. Since the pressure drop from the junction at C to the atmosphere at A and B must be equal for both branches, the junction pressure must be the same.
Column 19 Line 2	Computes the effect of the hood suction at A due to the increased pressure at junction C. Hood suction increases proportionally to an increase in junction pressure (Column 18/Column 17).
Column 20 Line 2	Gives the effect of the high pressure on the branch volume. Branch volume is related to pressure by:

$$\text{Corrected cfm} = \text{design cfm}\sqrt{\frac{\text{Pressure of branch with larger } \Delta p}{\text{Pressure of branch with smaller } \Delta p}}$$

Column 21 Line 2	Corrected branch volumetric flow rate.
Column 22 Line 2	Expected hood suction.

The next section of columns examine flow through the main:

Column 23 Line 3	Sum of the corrected branch volumes entering the main at C (AC plus BC).
Column 24	Read from the friction chart in Figure 9-8.
Column 25	Taken from the sketch, Figure 9-9.
Column 26	Taken from the sketch of the system, Figure 9-9.
Column 27	Given.
Column 28	Read from Table 9-3.
Column 29	Sum of Columns 26 and 28.
Column 30	Read from Figure 9-8 for the conditions given in Columns 23 through 25.
Column 31	Product of Columns 29 and 30, divided by 100. This number is added to Column 17, line 4 and entered in Column 18, line 4. This is due to the fact that the pressure in Column 17, line 4 is greater than Column 18, line 2.

The total static pressure supplied by the fan is computed by adding lines 6, 7 and 8 of Column 18. The computations proceed in a similar manner, with the negative pressure of the fan inlet found to be 3.28 in. H_2O. The adsorber bed pressure is entered in Column 18, line 8 and the total pressure that must be developed by the fan is 6.76 in. H_2O.

Table 9-4b. Exhaust system calculations.

→	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31
Line # ↓							Expected		In the main							Friction	
	Branch	Section of main	Tentative pressure, Col 14 (in. H_2O)	Expected pressure, Cols 18 + 31 (in. H_2O)	$\frac{\text{Col 18}}{\text{Col 17}}$	$\sqrt{\frac{\text{Col 18}}{\text{Col 17}}}$	Branch volume, Col 3 × Col 20 (cfm)	Hood static pressure, Col 13 × Col 19 (in. H_2O)	Q sum of Col 21 (cfm)	v (fpm)	d (in.)	Straight pipe, l (ft)	Elbows and angle (#/angle)	Straight pipe eq. (ft)	Pipe and elbows, l (ft)	(in. H_2O per 100 ft section)	In section of main (in. H_2O)
1	BC		2 72	2 72	1.0	1.0	3000	1.85									
2	AC		2 34	2.72	1 2	1 1	5676	1.78									
3		CE							8676	3288	22	6	—	—	6	0.90	0.05
4	DE		3 14	3 19	1 01	1.005	5186	2 39									
5		EF							13,862	3762	26	10	—	—	10	0.91	0.09
6	Fan Inlet			3.28													
7		FG		0 48						3762	26	10	1@90°	43	53	0.91	0.48
8	Adsorber			3.00													
Total pressure required by fan = 6.76																	

Note. Lines 1-6, pressures are all negative
Lines 7 and 8, pressures are all positive.
Total pressure is the absolute pressure (regardless of sign).

The above set of calculations determines the system resistance for a *specific* total air volume, Q. Resistances for various air volumes can be computed from the following relationship:

$$\text{New system resistance} = \left(\frac{\text{New system volume}}{\text{Calculated system volume}}\right)^2 \text{Calculated system resistance}$$

By knowing the computed volume and pressure at the fan, the pressure at other volumes can be calculated. These points can then be plotted to give a system curve which describes the system under all flow conditions. The pressure-volume relationship of the fan can be plotted on the same graph. The operating point is the intersection of the two curves. Figure 9-10 illustrates a typical graph of this type. Note that the above relationship is based on the premise that the pressure losses through *each element* of a system vary proportionally to the square of the flow rate. Certain equipment does not exhibit this relationship, especially some air pollution control equipment.

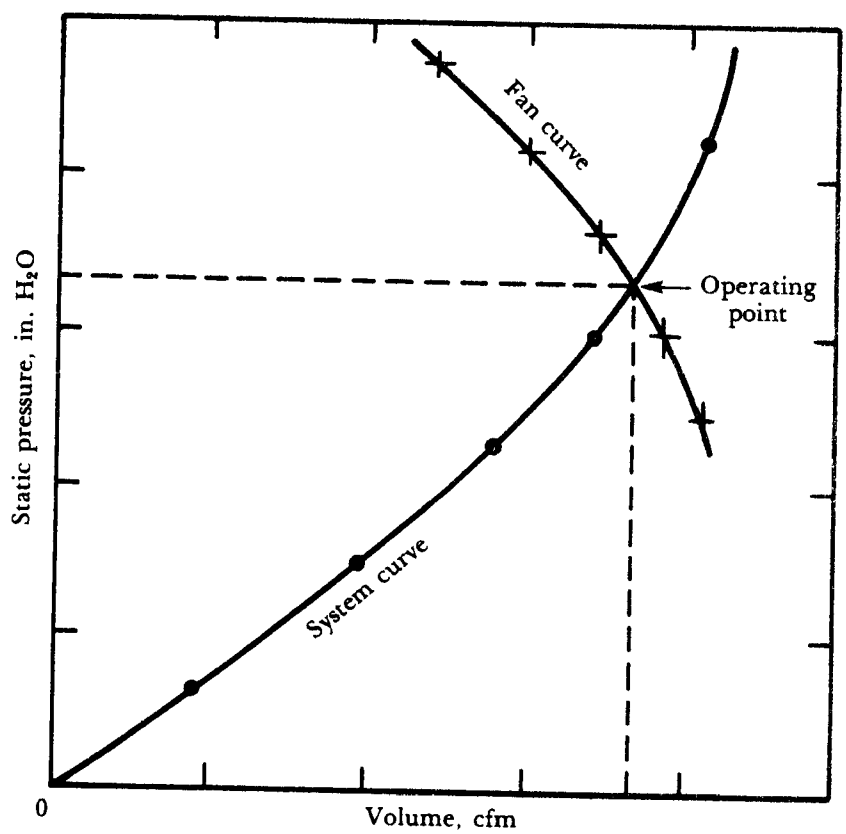


Figure 9-10. Typical point of operation.

Also, in Example 9-4, air was assumed to be at the standard atmospheric conditions of 70°F and 29.92 in. Hg. Under these conditions, the density of air is 0.075 lbs/ft³. Corrections must be made when process conditions vary from these standards. This is especially true for systems handling hot, saturated air streams.

Fans

In controlling air pollution, it is necessary to move the contaminants from their point of generation to the control device. Air and gas moving devices using rotary motion are normally termed *fans*. Fans are classified as *axial* or *centrifugal* depending on the direction of air flow through the impeller.

In axial flow fans, air moves directly forward through the axis of rotation of the fan blades (Figure 9-11). A home window or room fan is an example of an axial flow fan. For industrial use, axial fans are best suited for moving large volumes of clean air with low static pressures.

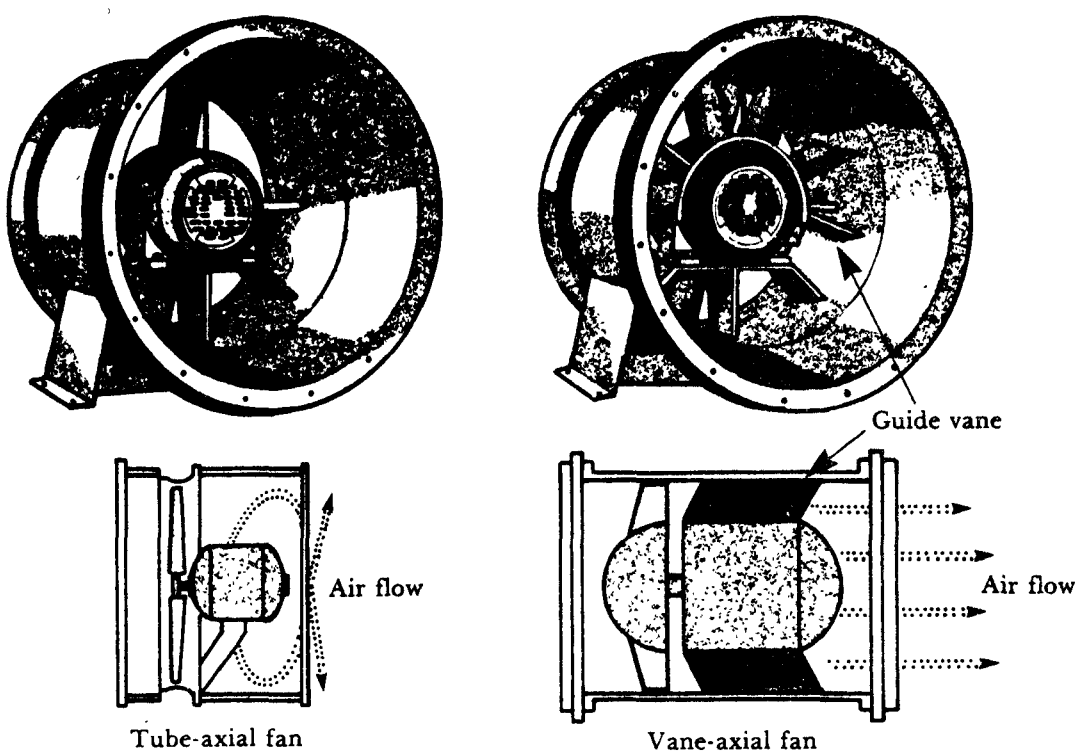


Figure 9-11. Axial fans.

In centrifugal fans, the air is introduced into the center of a revolving wheel or rotor and exits at right angles to the rotation of the blades (Figure 9-12). Centrifugal fans are more widely used than axial fans, since they can handle dirtier air streams with higher system resistances.

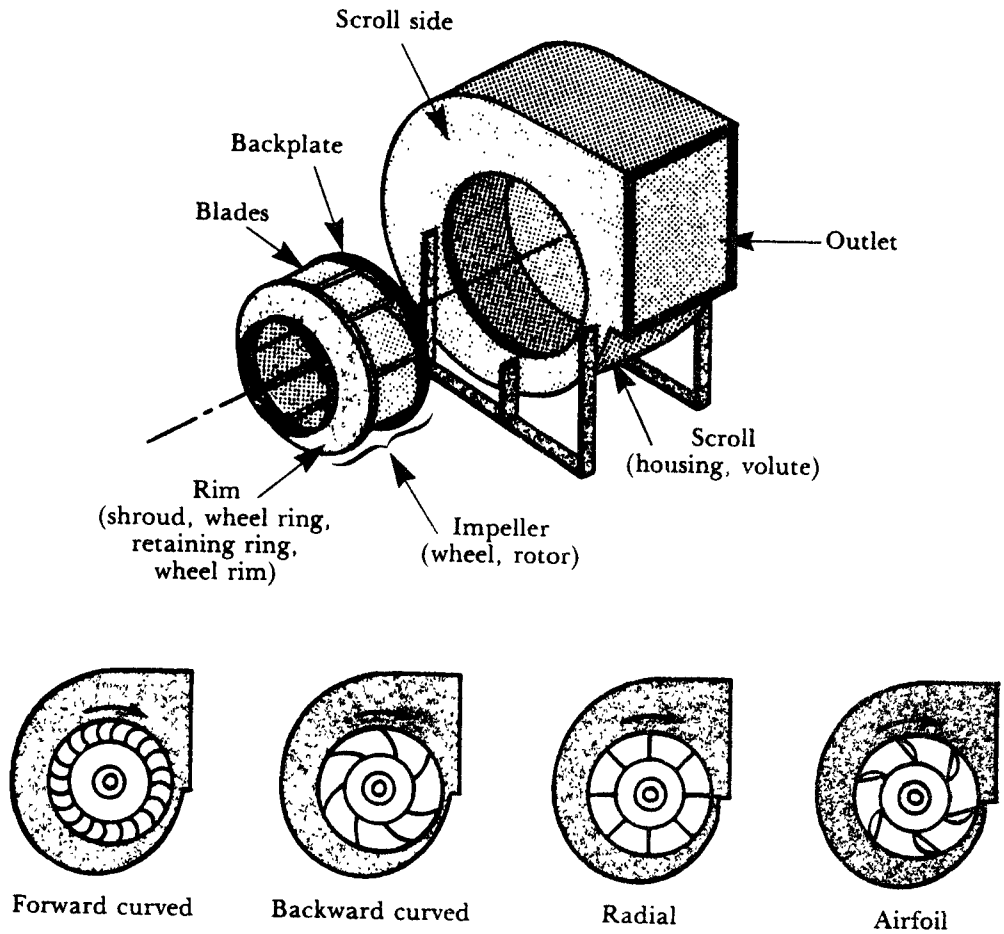


Figure 9-12. Centrifugal fans.

Two other terms used to describe air movers are *blowers* or *compressors*. Distinguishing between air movers and fans is somewhat arbitrary and is based on the static pressure they are capable of producing. The term "fan" is generally used to describe air movers with pressure rises up to about 2 psig (Pollak, 1973). The term "blower" refers to a centrifugal fan or positive displacement compressor operating with static pressures in the 2 to 15 psig range. For pressures higher than 15 psig, the term "compressor" is used. Unlike fans, some blowers and compressors operate by piston movement (positive displacement) rather than rotation.

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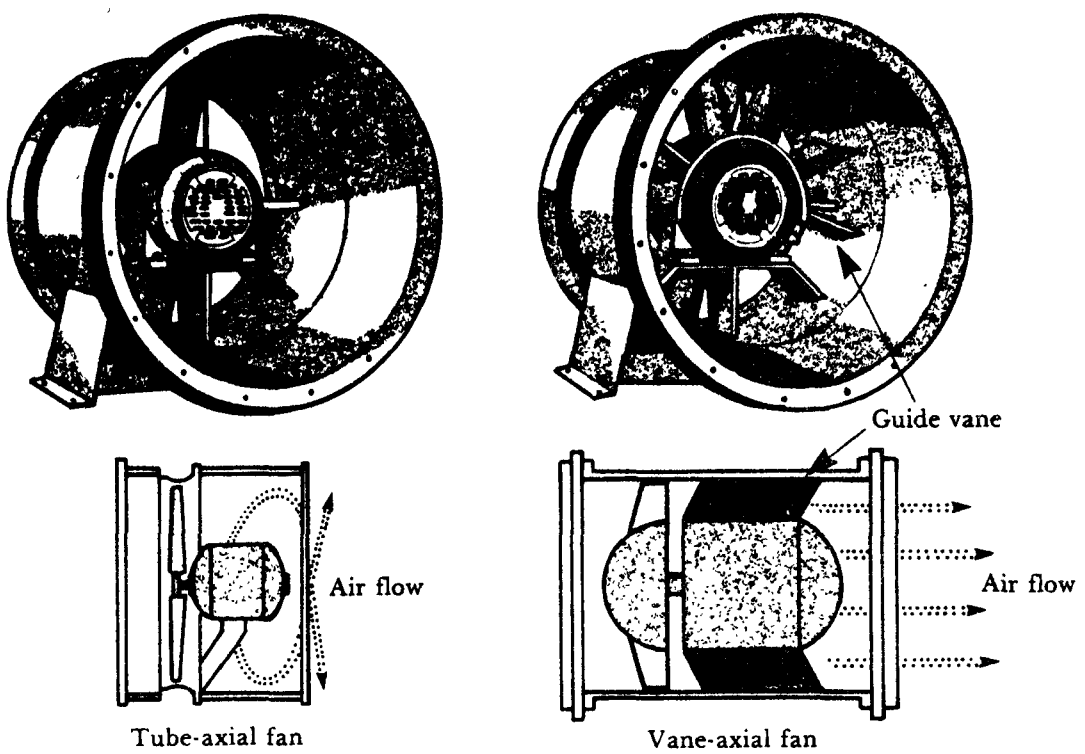

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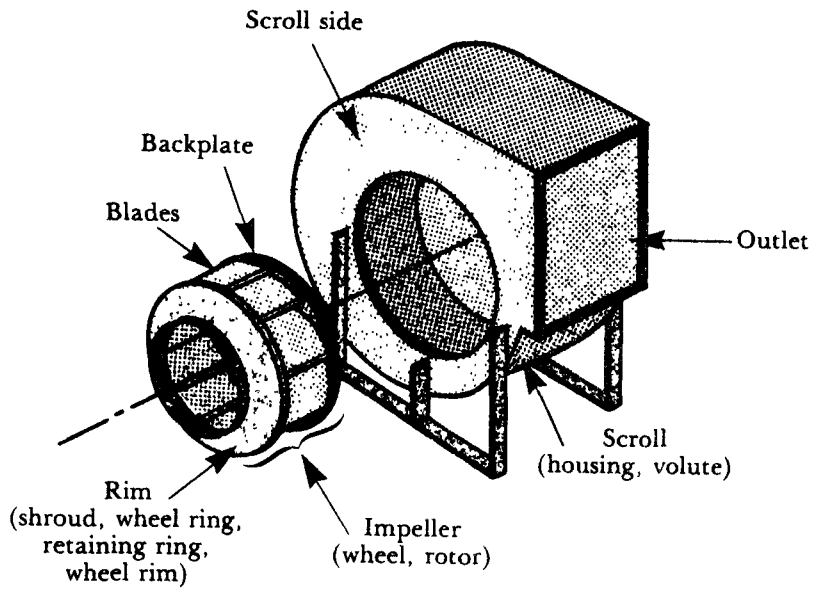

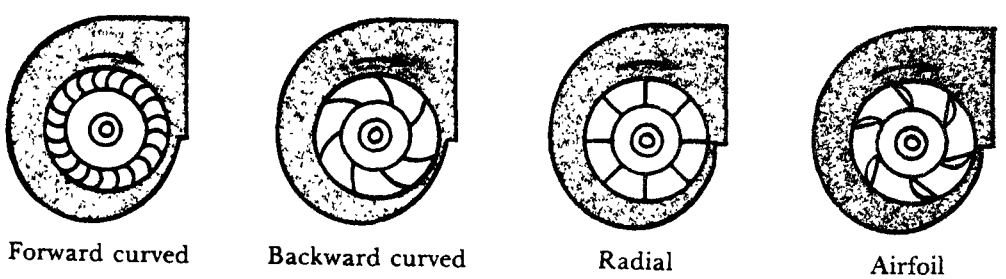

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Axial Fans

Axial fans are classified as tube-axial or vane-axial (see Figure 9-11). Axial fans can be designed to handle a wide range of air volumes, ranging from a few hundred cfm to 500,000 to 600,000 cfm.

Tube-axial fans are the simplest fan type, consisting of a propeller mounted in a tube (Figure 9-11) and therefore, are usually referred to as a *duct* fan. If the propeller blades are mounted outside a tube (roof or room vent) it is referred to as a *propeller* fan.

Tube-axial fans operate at static pressures as high as 4 inches of water and give a maximum mechanical efficiency (output flow power/input brake horsepower) of 75 to 80% (Summerell, 1981).

The *vane-axial* fan is similar to the tube-axial, except it has been aerodynamically designed to operate at higher efficiencies (see Figure 9-11). One difference is that it contains guide vanes that correct the helical motion of the air leaving the blade tip. This reduces the amount of turbulence produced and increases mechanical efficiency to a maximum of 85% (Summerell, 1981). High efficiency vane-axial fans are more efficient than comparable centrifugal fans (Crocker, 1981). They have been used for energy conservation in Europe for a number of years.

Vane-axial fans normally consist of short, stubby blades mounted on a large hub, which also helps to reduce turbulence. Single stage vane-axial fans can operate at pressures as high as 6 to 9 inches of water. Double stage fans are capable of much higher pressures, but are seldom used (Alden, 1970).

Centrifugal Fans

Air entering a centrifugal fan is rotated between the blades and ejected at a 90° angle to the blade. The centrifugal force compresses the air which gives added static pressure to the air. Centrifugal fans are enclosed in a scroll-shaped housing which helps convert kinetic energy to static pressure. Air rotating between the fan blades is compressed in the fan scroll which increases the static pressure. Centrifugal fans are classified by blade configuration as *radial, forward curved, backward curved* or *air foil.*

Radial or Straight Blade

Radial or straight blade fans physically resemble a paddle wheel with long radial blades attached to the rotor and are the simplest design of all centrifugal fans (Figure 9-12). This enables most radial blade fans to be built with great mechanical strength and be easily repaired.

These fans can be used in a variety of applications, especially handling dirty or sticky materials at high pressure. The flat blades tend to resist abrasion better than other designs. They can handle high static pressures, to 60 inches water gage and beyond (Summerell, 1981). However, these fans are the least efficient of the centrifugal type. A typical efficiency is 65%, but an efficiency of 12% is not too uncommon (Perry, 1976).

Forward Curved

Forward curved fans have both the heel and the tip of the blade curve forward in the direction of rotation (Figure 9-12). Blades are smaller and spaced much closer together than in other blade designs. Air leaves the tip of the blades at a greater velocity than the wheel-tip speed. Therefore, these fans are capable of transferring a large volume of gas for a minimum wheel diameter.

Forward curved blade fans are generally not used when dust or sticky materials are present. Contaminants easily stick to the blades and cause operational problems. These fans are used in low pressure (4 inches water gage), high air volume ventilation application. Maximum mechanical efficiency is approximately 70 to 75% (Summerell, 1981).

Backward Curved or Backward Inclined

Backward curved or backward inclined fans have blades inclined in opposition to the direction of rotation. This feature causes air to leave the tip of the blade at a lower velocity than the wheel-tip speed. This improves the mechanical efficiency (75 to 80%) and enables the fan to be used in slightly higher pressures (up to 15 inches water gage) than forward curved fans (Summerell, 1981).

Airfoil

The airfoil is similar to the backward curved fan, except that the blade has been contoured (airfoil) to increase stability and operating efficiency. These fans are more expensive to construct than backward curved fans, but have lower power requirements. They have an operating efficiency of approximately 85% (Perry, 1976). They are susceptible to abrasion and are not usually used for moving dirty gas streams.

In general, centrifugal fans are easier to control, more robust in construction, and less noisy than axial units (Pollak, 1973). Centrifugal fans have a broader operating range at their highest efficiencies. They are better suited for operations with flow variations. Centrifugal fans are more widely used in process operations since they can handle dust and fumes better than axial fans. This is especially true for radial blade centrifugal fans. High efficiency axial fans are beginning to be used on some clean, induced draft applications (i.e.: clean side of pollution control equipment) as an energy conservation method. Axial fans are smaller than centrifugal fans for the same application.

Fan Operation

Fan operation is characterized by the volume of gas flow, the pressure at which this flow is produced, the speed of rotation, the power required, and the efficiency (EPA, 1973). For a specific size fan operating at single speed, **performance curves** can be drawn from experimental data. Figure 9-13 illustrates typical performance curves for a backward curved fan. It is important to note that each type of fan has its own characteristic performance curves.

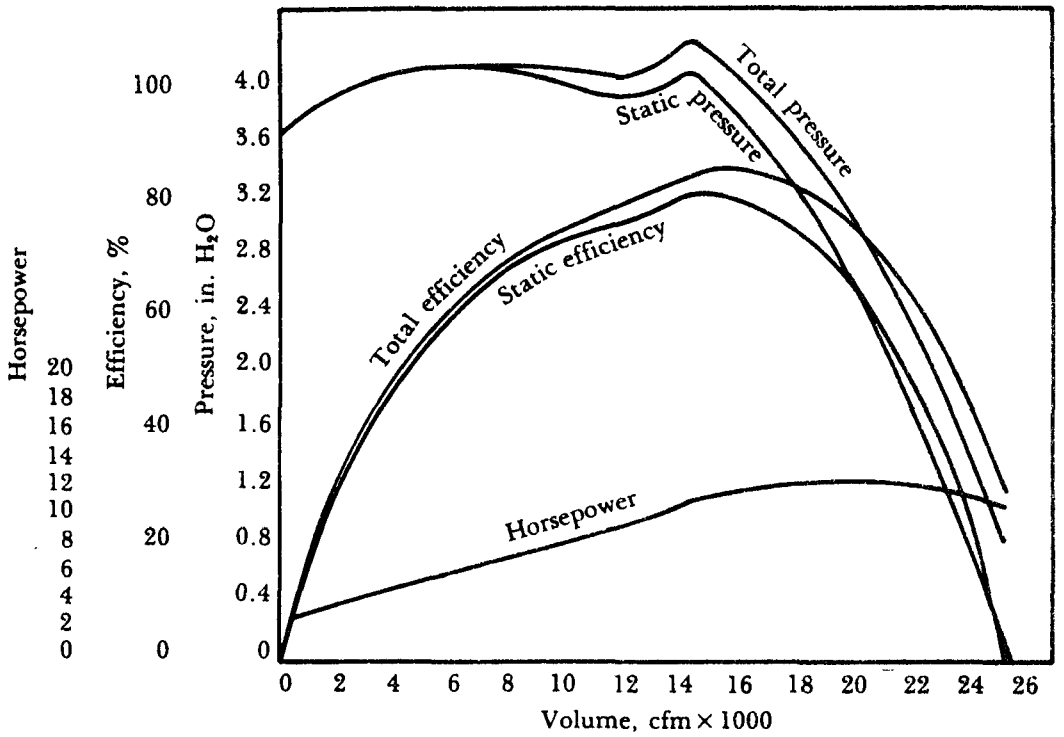
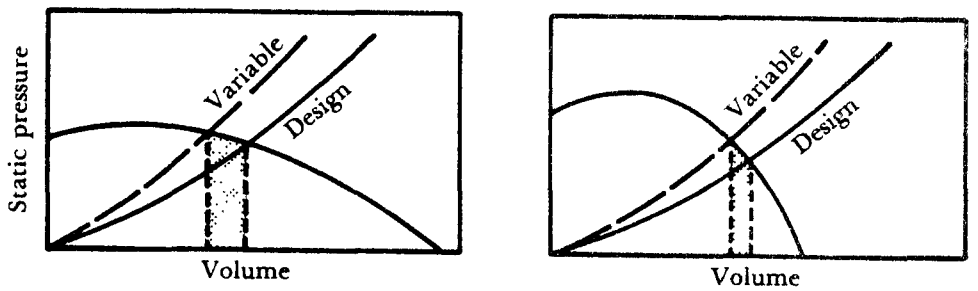


Figure 9-13. Typical performance curve for a backward curved centrifugal fan.

Performance curves can be used to select the proper fan for a particular job. The point of operation (or rating) is determined by the intersection of the static pressure and the system flow resistance lines (see Figure 9-10). Curves like those of Figure 9-10 are used to determine if the operating point is in a desirable location. For example, the system curve should intersect the static pressure line so that operation is near the maximum efficiency range and is in a low power requirement range. Also, the system curve should intersect the pressure curve in an area with a steep slope. This gives good flow control since large pressure changes bring small variations in air volumes (Figure 9-14). If the pressure curves were flat, a small pressure change would be accompanied by a large volume change. These flow considerations are important since system pressure usually is somewhat variable and the operation point is determined from design data which may vary from actual operation.

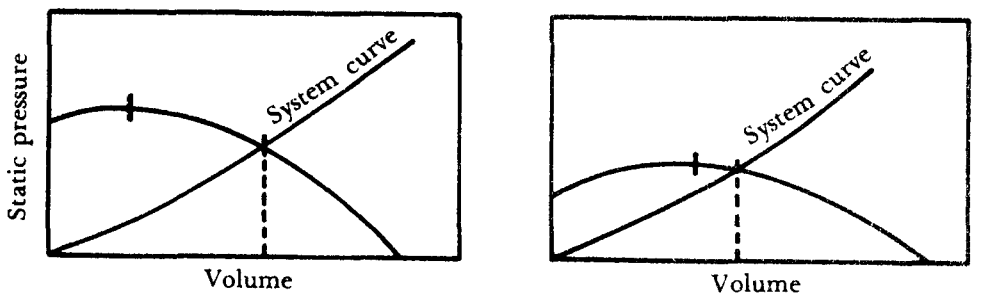
Effect of varying system pressure



Poor selection
Fan with flat pressure curve gives wide volume variation with pressure change.

Good selection
Fan with steep pressure curve gives small volume variation with pressure.

Effect of fan size



Poor selection
Small fan used with system curve crossing fan too far to right of peak.
Excessive horsepower.
Low efficiency.

Good selection
Large fan used with curve crossing fan curve to right of peak.
Low horsepower.
High efficiency.

Source: Permission granted by Committee on Industrial Ventilation, American Conference of Governmental Industrial Hygienists, Box 16153, Lansing, MI 48901. Taken from 14th Edition (1976) of *Industrial Ventilation*.

Figure 9-14. Use of performance curves for fan selection.

To help in the actual selection of fan size and speed, fan manufacturers have put together *multi-rating tables*. These tables give the performance of a particular fan size for its range of capacities. Table 9-5 is a typical fan rating table. The fan size and dimensions are usually listed at the top of the table. Values of static pressure are arranged as columns which contain the fan speed and horsepower required to produce various volume flows. The point of maximum efficiency at each static pressure is usually underlined or printed in a special type.

Table 9-5. Typical fan rating table.

Wheel style: backward—inclined
Wheel diameter: 50½ in.
Maximum fan speed: 1134 rpm
Performances underlined are those at maximum efficiency

cfm	p_s	½ in.		1 in.		1½ in.		2 in.		2½ in.		3 in.	
	OV	rpm	BHP	rpm	BHP	rpm	BHP	rpm	BHP	rpm	BHP	rmp	BHP
5727	1250	216	0.74	278	1.34	330	2.01	378	2.75	421	3.50	460	4.28
6873	1500	236	0.97	291	1.67	339	2.42	384	3.24	424	4.06	462	4.91
8018	1750	250	1.27	305	2.05	352	2.87	393	3.76	430	4.69	467	5.66
9164	2000	271	1.63	320	2.49	366	3.42	405	4.35	441	5.36	475	6.40
10309	2250	293	2.12	338	3.05	381	4.06	419	5.06	453	6.14	486	7.26
11455	2500	315	2.72	356	3.65	396	4.76	432	5.88	468	7.03	499	8.19
12600	2750	337	3.46	377	4.39	413	5.58	448	6.81	482	8.04	514	9.31
13746	3000	360	4.39	399	5.25	430	6.48	465	7.82	496	9.16	527	10.53
14891	3250	382	5.43	421	6.25	451	7.52	481	8.93	512	10.35	542	11.80
16037	3500	405	6.66	442	7.48	473	8.67	501	10.16	529	11.69	557	13.25
17182	3750	429	8.08	463	8.97	496	10.05	521	11.54	547	13.18	574	14.81
18328	4000	451	9.64	486	10.61	517	11.61	543	12.99	566	14.78	591	16.49
19473	4250	474	11.46	510	12.47	539	13.47	565	14.85	587	16.56	610	18.35
20619	4500	497	13.51	532	14.55	560	15.60	587	16.82	610	18.54	630	20.40
21764	4750	520	15.82	556	16.82	584	17.98	608	19.17	632	20.77	652	22.59

Source: Bayler Blower Co.

Note: performances based on standard air density of 0.075 lbs/ft^3.

In specifying the proper size fan, it is desirable to know the horsepower requirement for the system. Two terms are used when describing horsepower: *air horsepower* (AHP) and *brake horsepower* (BHP). Air horsepower is the power output of a fan if the fan were 100% efficient. Since no fan is 100% efficient the more important term is the brake horsepower. Brake horsepower equals the air horsepower times the efficiency of the fan. Knowing the efficiency of the fan, brake horsepower can be calculated from:

(Eq. 9-12)

$$BHP = \frac{Q \times P_T}{6356 \times \mathscr{E}}$$

Where: Q = air flow, cfm
P_T = total pressure of the system, in. H_2O
$\mathscr{E}$ = mechanical efficiency of the fan, percent

In order to select a fan for the exact conditions desired, it is sometimes necessary to interpolate between values presented in the multi-rating tables. Straight line interpolation can be used with negligible error for multi-rating tables based on a **single** fan size. Some multi-rating tables attempt to show ratings for a whole series of geometrically similar (homologous) fans in one table. Unless the desired rating point appears in the table, interpolation is not advised (*Industrial Ventilation*, 1976).

It is also important to note that most multi-rating tables and performance curves are based on air at standard conditions (70°F and 14.7 psia). When air is not at standard conditions, volume, system pressure, and horsepower must be corrected. These corrections are proportional to the change in density at the operating temperature and pressure.

Fan Laws

Fan laws are used to predict how a change in one of the performance variables (size, speed, air density, volume, system pressure, power, efficiency, and sound level) affects the others. Fan laws are based on the premise that, **if two fans are geometrically similar (homologous) their performance curves have the same shape**. Therefore, if the two fans operate at the **same point of rating**, their efficiencies are equal and the ratio of all other variables are interrelated.

The fan laws can be written in as many as ten different ways depending on which variables change. The following relationships are some of the more useful of the fan laws.

For a change in fan speed ($rpm_1 \rightarrow rpm_2$):

- the flow rate (Q) varies directly with the fan speed ratio

$$\frac{Q_1}{Q_2} = \frac{rpm_1}{rpm_2}$$

- the power ($\mathscr{P}$) varies with the cube of the fan speed ratio

$$\left(\frac{\mathscr{P}_1}{\mathscr{P}_2}\right) = \left(\frac{rpm_1}{rpm_2}\right)^3$$

- the static pressure (p_s) varies with the square of the fan speed ratio

$$\left(\frac{p_{s1}}{p_{s2}}\right) = \left(\frac{rpm_1}{rpm_2}\right)^2$$

For a change in fan size (wheel diameter $d_{w1} \rightarrow d_{w2}$):

- the flow rate varies with the cube of the wheel diameter

$$\frac{Q_1}{Q_2} = \left(\frac{d_{w1}}{d_{w2}}\right)^3$$

- the power varies with the fifth power of the wheel diameter

$$\frac{\mathscr{P}_1}{\mathscr{P}_2} = \left(\frac{d_{w1}}{d_{w2}}\right)^5$$

- the static pressure varies with the square of the wheel diameter

$$\frac{p_{s1}}{p_{s2}} = \left(\frac{d_{w1}}{d_{w2}}\right)^2$$

For a change in gas density ($\varrho_1 \rightarrow \varrho_2$)

- the flow rate is constant

$$\frac{\varrho_1}{\varrho_2} = Q$$

- the power varies directly with the density ratio

$$\frac{\mathscr{P}_1}{\mathscr{P}_2} = \frac{\varrho_1}{\varrho_2}$$

- the static pressure varies directly with the density ratio

$$\frac{p_{s1}}{p_{s2}} = \frac{\varrho_1}{\varrho_2}$$

The fan laws can be used to construct performance curves or to determine new operating conditions as process variables change. It is important to note that these fan laws are based on the restriction that the fan at the new conditions will operate at the same point of rating (mechanical efficiency). Therefore, the complete set of operating conditions should be calculated for any change.

Example 9-5

A fan operating at a speed of 1474 rpm delivers 10,200 cfm at 4 in. static pressure and requires 8.85 brake horsepower (BHP). What will be the new operating conditions if the fan is speeded up to 2000 rpm.

Solution:

The new flow is:

$$\frac{Q_2}{Q_1} = \frac{rpm_2}{rpm_1}$$

$$Q_2 = (10{,}200 \text{ cfm}) \left(\frac{2000 \text{ rpm}}{1474 \text{ rpm}}\right)$$

$$= 13{,}840 \text{ cfm}$$

The new static pressure of the system is:

$$\frac{p_{s2}}{p_{s1}} = \left(\frac{rpm_2}{rpm_1}\right)^2$$

$$p_{s2} = 4 \text{ in.} \left(\frac{2000}{1474}\right)^2$$

$$= 7.4 \text{ in.}$$

The horsepower required is:

$$\frac{\mathscr{P}_2}{\mathscr{P}_1} = \left(\frac{\text{rpm}_2}{\text{rpm}_1}\right)^3$$

$$\mathscr{P}_2 = 8.85 \text{ hp} \left(\frac{2000}{1474}\right)^3 = 22.1 \text{ hp}$$

Note the drastic increase in horsepower for increasing the fan speed.

Example 9-6

The exhaust system for a rotary dryer is designed to deliver 12,000 cfm of air at 600°F and 4 in. H_2O static pressure. Fan speed is 630 rpm and requires 13 horsepower. If the dryer has been down for repairs, compute the horsepower required to pull the same amount of air at ambient conditions.

Given: density of air at 70°F = 0.075 lb/ft³
density of air at 600°F = 0.0375 lb/ft³

Solution:

The horsepower is:

$$\frac{\mathscr{P}_2}{\mathscr{P}_1} = \frac{\varrho_2}{\varrho_1}$$

$$\mathscr{P}_2 = 13 \text{ hp} \left(\frac{0.075}{0.0375}\right) = 26 \text{ hp}$$

Note: this would require a motor twice the size as needed for normal operation. An alternative would be to install dampers until the system reaches operating temperature.

Noise

Some fan systems can be objectionably noisy and cause problems in the work area or at neighboring residences. Fan manufacturers can usually supply data on the expected sound level generated by a particular fan. However, it is extremely difficult to determine the maximum noise level of a fan prior to actual operation. The sound power generated by a fan depends on the flow, fan-pressure level, and impeller type and configuration (Pollak, 1973). In general, axial fans are noisier than centrifugal fans.

A number of possible methods can reduce the noise level of a fan.

1. Apply acoustical insulation to the fan housing or ductwork. A housing enclosure around the fan unit could be considered if the noise level is severe.
2. Use a venturi inlet to smooth the flow of air into the fan. This can lower the noise level by reducing turbulence in the inlet of the fan.
3. Install absorption type silencers on the inlet and/or outlet of the exhaust. These silencers are made for insertion in round or rectangular ducts, in standard or special materials, and with special acoustical fills for corrosive atmospheres. These devices do add to the pressure drop in the system.

The above noise reduction techniques are for fans that are designed and operated properly. The point of minimum noise level of a fan is close to the point of maximum efficiency. If a fan is not operating properly, the above solutions may not cure the problem or be the most effective solution. Putting a fan in proper balance can be the most effective solution.

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Appendix A

Common International System of Units (SI)

Quantity (1)	Some common units	Symbol	Equivalent	Symbol
length	kilometer	km		
	meter	m		
	centimeter	cm		
	millimeter	mm		
	micrometer	μm		
area	square kilometer	km^2		
	square hectometer	hm^2	hectare (2)	ha
	square meter	m^2		
	square centimeter	cm^2		
	square millimeter	mm^2		
volume	cubic meter	m^3	liter (3)	L
	cubic decimeter	dm^3		
	cubic centimeter	cm^3	milliliter (3)	mL
speed or velocity	meter per second (12)	m/s		
	kilometer per hour (4)	km/h		
acceleration	meter per second squared	m/s^2		
rotational frequency	revolution per second	r/s		
	revolution per minute (4)	r/min		
mass (5)	megagram	Mg	metric ton	t
	kilogram	kg		
	gram	g		
	milligram	mg		
density	kilogram per cubic meter	kg/m^3	gram per liter	g/L
force	kilonewton	kN		
	newton	N	kilogram meter per second squared	$kg \cdot m/s^2$
movement of force (6)	newton meter	N•m		
pressure (or vacuum)	kilopascal	kPa		
	pascal	Pa	Newton per meter squared	N/m^2
stress	megapascal	MPa		
viscosity (dynamic)	millipascal second (7)	mPa•s		
	pascal second	Pa•s		
viscosity (kinematic)	square millimeter per second (8)	mm^2/s		
energy, work, or quantity of heat	joule (9)	J	Newton meter	N•m
	kilowatt hour (10)	kW•h	kilowatthour	kWh
power, or heat flow rate	kilowatt	kW		
	watt	W		
temperature, or temperature interval	kelvin	K		
	degree Celsius (11)	°C		

Source: The American National Metric Council, 1978.

NOTES

(1) Any measurable property (such as length, area, temperature) is called a *quantity*.

(2) For land or water area only.

(3) To be used only for fluids (both gases and liquids), and for dry ingredients in recipes, or for volumetric capacities. Do not use any prefix with liter except *milli*.

(4) The symbols for minute, hour, and day are min, h, and d, respectively.

(5) Commonly called *weight*.

(6) Torque or bending movement.

(7) 1 mPa•s = 1 cP (centipoise, which is obsolete).

(8) 1 mm^2/s = 1 cSt (centistokes, which is obsolete).

(9) The unit-multiples kilojoule (kJ) and megajoule (MJ) are also commonly used.

(10) To be abandoned eventually. 1 kW•h = 3.6 MJ.

(11) The degree mark ° is always used in °C to avoid confusion with coulomb (C), but never with K for kelvin.

(12) Second is denoted by s in SI units.

Appendix B

Conversion Factors

Length
1 inch = 2.54 cm
1 m = 3.048 ft
1 ft = 0.305 m

Mass
1 lb = 453.6 g
1 kg = 2.2 lb

Pressure
1 atm = 101,325 Pa
= 760 mm Hg (0°C)
= 14.7 psia

Force
1 N = 1 kg•m/s^2

Energy
1 cal = 4.184 J
1 J = 9.48×10^{-4} Btu
1 Btu = 252.2 cal

Kinematic viscosity
1 m^2/s = 10^4 stokes

Power
1 W = 1 J/s
1 hp = 33,479 Btu/hr

Area
1 cm^2 = 0.155 in^2
1 m^2 = 10.764 ft^2

Volume
1 cm^3 = 0.061 in^3
1 m^3 = 35.31 ft^3
1 barrel (oil) = 42 gal
1 ft^3 = 7.48 gal
1 ft^3 = 28.317 liters

Density
1 kg/m^3 = 0.0624 lb/ft^3

Dynamic viscosity
1 Pa•s = 1 N•m/s = 1000 centipoise
1 cp = 0.000672 lb/ft•sec

Volume flow
1 m^3/s = 35.3 ft^3/sec
1 m^3/min = 35.3 ft^3/min
1 scfm = 1.7 m^3/h
1 gpm = 0.227 m^3/h

Velocity
1 m/s = 3.048 ft/sec
1 mi/hr = 0.447 m/s

Geometry
area of circle = πr^2
circumference of circle = $2\ \pi r$
surface area of sphere = $4\ \pi r^2$
volume of sphere = $4/3\ \pi r^3$
area of cylinder = $2\ \pi r h$

Appendix C

Constants and Useful Information

$\pi = 3.14$

Gas constants

$R = 0.0821$ atm liter/g mol•K
$= 83.14 \times 10^6$ g•cm³/s²•g mol•K
$= 8.314$ N•m/g mol•K
$= 0.7302$ atm•ft³/lb mol•°R
$= 1.987$ g•cal/g mol•K or Btu/lb mol•°R

Acceleration of gravity

$g = 32.17$ ft/sec² $= 980.7$ cm/s² $= 9.8$ m/s²

Newton's conversion constant

$g_c = 32.17\ (\mathrm{lb}_{mass})(\mathrm{ft})/(\mathrm{lb}_{force})(\mathrm{sec}^2)$

1 lb mol = 359 ft³ of ideal gas at STP (32°F and 14.7 psia)
1 g mol = 22.4 liters of ideal gas at STP (0°C and 760 mm Hg)

C_p for water ≈ 1 Btu/lb•°R ≈ 1 cal/g °C (at 20°C and 1 atm)
C_p for air ≈ 0.26 Btu/lb•°R ≈ 0.26 cal/g °C

viscosity of water, $\mu = 1$ cp $= 0.01$ g/cm•s (at 20°C and 1 atm)
viscosity of air, $\mu_a = 4.1 \times 10^{-7}$ lb•sec/ft² $= 2 \times 10^{-5}$ N•s/m² (at 20°C and 1 atm)

density of air $= 1.29$ kg/m³ $= 7.49 \times 10^{-2}$ lb/ft³ (at 20°C and 1 atm)
density of water $= 1$ g/cm³ $= 62.4$ lb/ft³ (at 4°C and 1 atm)

1 cubic foot of air weighs 34.11 g

conversion from ppm to g/m³ at STP (273.15 K and 1 atm)

$$\frac{g}{\mathrm{dscm}} = \frac{\mathrm{ppm} \times \mathrm{MW}\left(\dfrac{g}{g\ \mathrm{mol}}\right)}{22.414\ \dfrac{\mathrm{liters}}{g\ \mathrm{mol}} \times 10^{-3}\ \dfrac{\mathrm{m}^3}{10^3\ \mathrm{liters}}\left(\dfrac{293.15\ \mathrm{K}}{273.15\ \mathrm{K}}\right)} \times \frac{1}{1 \times 10^6\ \mathrm{ppm}}$$

Appendix D

Combustion Calculations for Theoretical and Excess Air Requirements

Theoretical or Stoichiometric Air

Consider a generalized fuel with a chemical formula $C_xH_yS_zO_w$ where the indices x, y, z, and w represent the relative number of atoms of carbon, hydrogen, sulfur, and oxygen respectively. Balancing the chemical reaction for the complete oxidation (combustion) of this fuel with oxygen from air gives:

(Eq. D-1)

$$C_xH_yS_zO_w + \left(x + \frac{y}{4} + z - \frac{w}{2}\right)O_2 + \frac{0.79}{0.21}\left(x + \frac{y}{4} + z - \frac{w}{2}\right)N_2 \rightarrow$$

$$\rightarrow xCO_2 + \frac{y}{2}H_2O + zSO_2 + \frac{0.79}{0.21}\left(x + \frac{y}{4} + z - \frac{w}{2}\right)N_2 + Q$$

Where: Q = heat of combustion

The above reaction assumes that:

- air consists of 21% by volume of oxygen with the remaining 79% made up of nitrogen and other inert gases;
- combined oxygen in fuel is available for combustion, thus reducing air requirements;
- fuel contains no combined nitrogen, so no "fuel NO_x" is produced;
- "thermal NO_x" via the nitrogen fixation is small, so that it is neglected in stoichiometric air calculations;
- sulfur in fuel is oxidized to SO_2 with negligible SO_3 formation.

For example, to combust methane (CH_4) Equation D-1 reduces to:

(Eq. D-2)

$$CH_4 + 2O_2 + 7.53N_2 \rightarrow CO_2 + 2H_2O + 7.53N_2 + Q$$

(Eq. D-3) Moles or relative volumes:

$$1 + \underbrace{2 + 7.53}_{\text{total air required}} \rightarrow \underbrace{1 + 2 + 7.53}_{\text{total flue gases}}$$

For every mole or standard cubic foot of CH_4 burned, the reaction requires 9.53 moles or standard cubic feet of air for complete combustion.

For mixtures of gaseous fuels it is easier to compute the amount of air required for each of the constituent compounds, e.g., methane, ethane, ethylene, etc. directly, using the constants from Table 3-1 (Chapter 3), and then adding them to get the total. Further, as the analyses of gaseous fuels are usually available on a volumetric basis, the volume rather than mass of stoichiometric air is of the most interest. Thus, for a unit volume of gaseous fuel, say 1 standard cubic foot, the volume of theoretical air, V_{at}, also in standard cubic feet is:

(Eq. D-4)

$$V_{at} = 2.38(CO + H_2) + 9.53CH_4 + 16.68C_2H_6 + 14.29C_2H_4 + 11.91C_2H_2 + \ldots + 7.15H_2S - 4.76O_2$$

where the molecular symbols now represent the volume fractions of the indicated components, and the numerical coefficients are again found in Table 3-1, but this time from the "mole per mole of combustibles or ft^3/ft^3 combustibles" column. Should the gas mixture contain other combustible substances not already included in Equation D-4, these can be added similarly. Absence of a substance means that its volume fraction is zero and that term will drop out of Equation D-4.

Excess Air

Several concentration corrections have been devised based on the combustion characteristics of fossil fuels. Excess air is defined as that percentage of air added in excess of that required to **just** combust a given amount of fuel. Normally, to achieve efficient fuel combustion, more air is needed than the stoichiometric amount, i.e., one carbon atom to two oxygen molecules.

Depending on the amount of excess air, different concentrations of CO_2 and oxygen in the stack gas will result, as shown in Figure D-1.

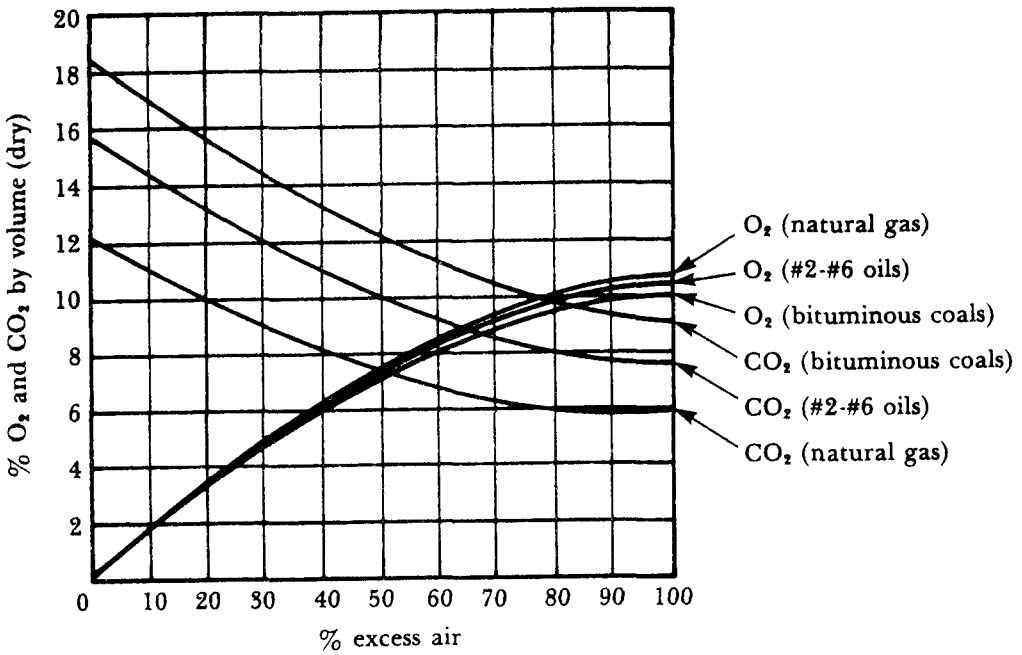


Figure D-1. Excess air concentrations of CO_2 and O_2 in stack.

Since the concentration of the pollutants produced in the source could be reduced by adding more excess air, emission calculations are corrected to a given excess air condition. A value of 50% excess air has been chosen as a reference condition since at one time many boilers operated at this condition. When this correction is made, it also accounts for dilution caused by air leaking in at the preheater or other duct work.

The expression for percent excess air, as given in EPA Method 3, is:

(Eq. D-5)

$$\%EA = \frac{\%O_2 - 0.5\%CO}{0.264\%N_2 - (\%O_2 - 0.5\%CO)} 100$$

The derivation of this expression is given in the APTI Course 450 Student Manual, Appendix D (EPA 450/2-79-006).

Appendix E

Capital Cost Estimations of Gaseous Control Equipment

This appendix contains generalized cost data for air pollution control systems described throughout this manual. These data should be used only as an **estimate** to determine system costs. In some cases, the cost of the control device may represent only a very small portion (<20%) of the total installed cost; in other cases it may represent the total cost. Variations in the total cost can be attributed to a number of variable factors such as cost of auxiliary equipment, new or retrofitted installation, local labor costs, engineering overhead, location and accessibility of plant site, and type of installation (factory or field assembled).

All cost estimating data, except for condenser data, was adapted from an EPA publication, *Capital and Operating Cost of Selected Air Pollution Control Systems* (1978). All equipment costs except condenser data, are based on a reference date of **December 1977** and estimated to be accurate to within ± 20% on a component basis, except where noted. For further information on the basis of these cost estimates the above cited document should be consulted.

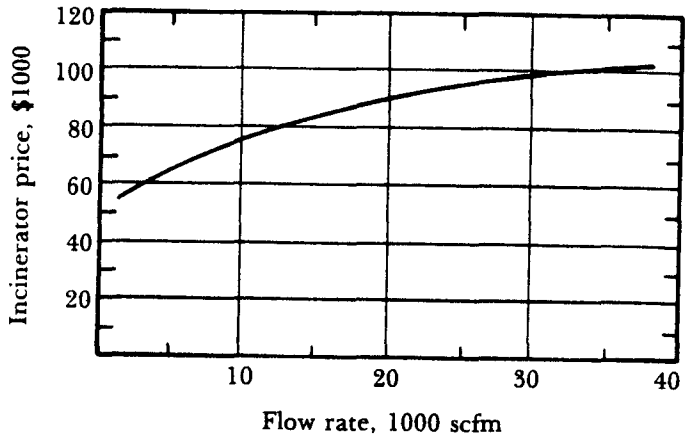
Thermal and Catalytic Incinerators

Prices for thermal incinerators including refractory linings, are contained in Figures E-1 and E-2. Catalytic incinerator prices are found in Figure E-3. Residence times for thermal incinerators are based on 0.5 seconds. From Figure E-1, the price of a thermal incinerator without a heat exchanger for a gas volume of 30,000 standard cubic feet per minute and 0.5 second residence time is $99,000. All gas volumes are measured at standard conditions.

The cost curves for thermal incinerators are based on an operating temperature of 1500°F. The cost of incinerators operating at other temperatures can be determined by adjusting the inlet gas flow rates to account for the temperature difference as follows.

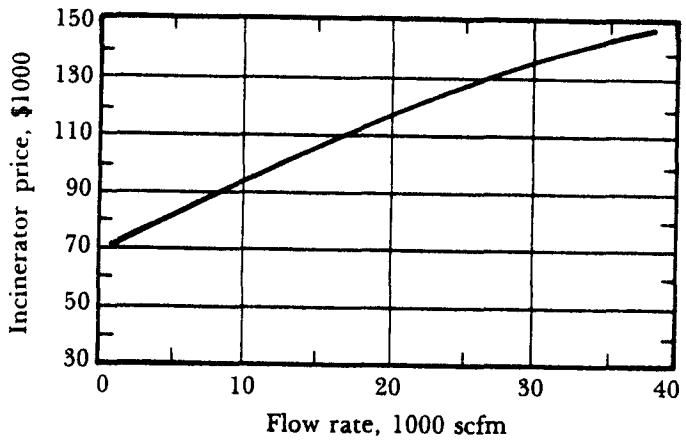
$$\frac{\text{flow rate (scfm) at T}}{\text{flow rate (scfm) at } T_b} = \frac{T + 460}{T_b + 460}$$

Where: T = new temperature, °F
T_b = baseline temperature, 1500°F
(flow rate in scfm)



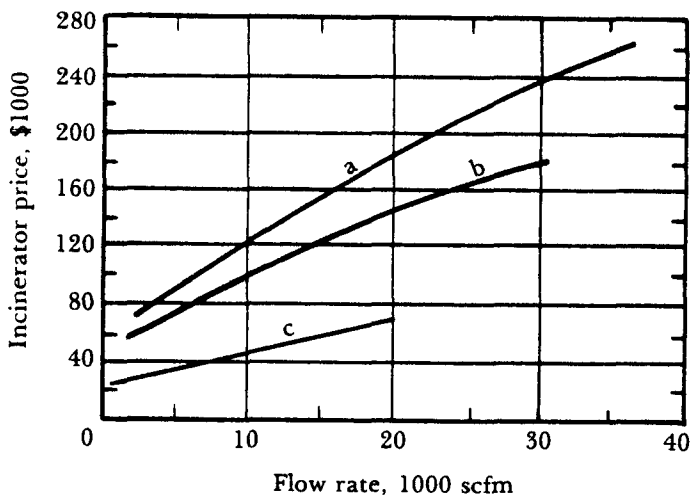
Note:
- Operating temperature of 1500°F.
- Residence time for incineration is 0.5 sec.
- Accuracy of this curve is ±50%.
- Price includes incinerator, fan or blower, controls, and instrumentation.
- Prices will vary as a function of:
  - retention times
  - materials of construction
  - special controls
  - heat content of pollutant
- Inlet concentrations of 0-25% LEL light hydrocarbons.
- Process temperatures of 70-300°F.
- Oil fired burners.

Figure E-1. Prices for thermal incinerators without heat exchangers.



Note:
- Residence time of 0.5 seconds.
- Process temperature is 70°F.
- Operating temperature is 1500°F.
- Curve based on 35% heat recovery.
- Accuracy of this curve is ±50%.
- Price includes incinerator, heat exchanger, fan or blower, damper controls and instrumentation.
- Prices will vary as a function of:
  - retention times
  - materials of construction
  - special controls
  - heat content of pollutant
- Inlet concentration of 0-25% LEL.
- Oil fired burners.

Figure E-2. Prices for thermal incinerators with primary heat exchanger.



Note:
- Accuracies of ±50%.
- Inlet concentration up to 25% LEL, light hydrocarbons except packaged units.
- Operating temperature of 1200°F.
- Price includes incinerator, fan or blower, stub stack, controls and instrumentation.

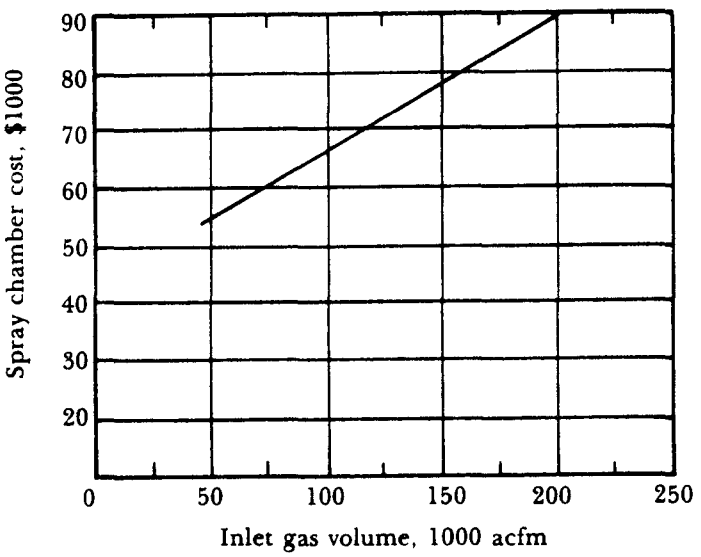
Where: a = custom units with 35% heat recovery using primary heat exchanger
b = custom units without heat exchanger
c = packaged units (for concentration units up to 6% LEL)

Figure E-3. Catalytic incinerator prices.

Absorbers*

Spray Chambers

The cost of a spray chamber is based on the size, volume of chamber, materials of construction, and water flow rate. Figure E-4 illustrates typical equipment costs for a spray chamber.



Note:
- Based on chamber velocity of 600 fpm.
- Length/diameter = 3.
- Effective length = 4 to 4.5.
- Carbon steel construction.
- Does not include refractory.
- Spray chamber cost includes vessel and support rings, platform, ladder, gratings, spray system, and controls.

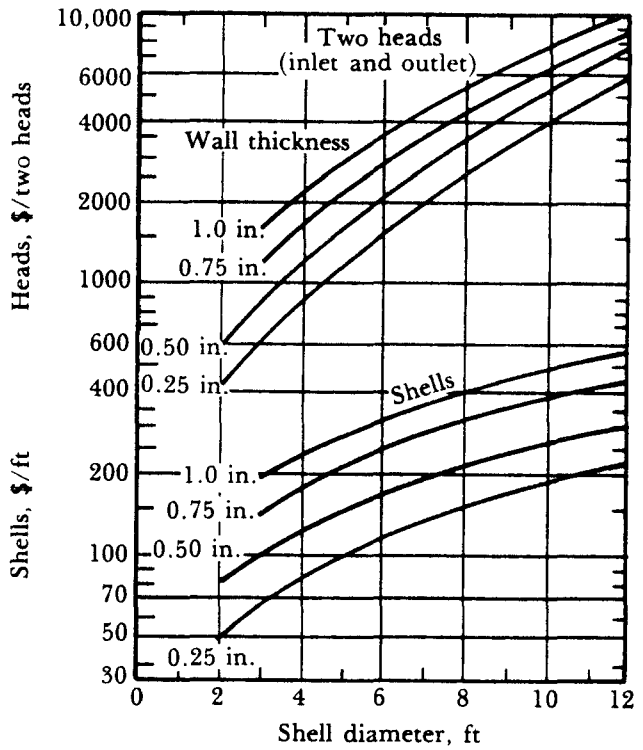
Figure E-4. Spray chamber costs vs. inlet gas volume.

Packed and Plate Towers

The cost of absorption towers is shown in Figures E-5 through E-8. The cost of these towers depends on the size, thickness, and materials of construction of the vessel since these units are basically custom designed for individual processes and applications. Figure E-5 illustrates the fabricated cost of a carbon steel vessel shell in dollars per linear foot plus the cost of two semi-elliptical inlets and outlets. The height of the vessel is determined by multiplying the number of transfer units by the height of the transfer units plus some additional height for vapor-liquid separation at the top of the tower and cleanout at the bottom (typically 2 to 3 ft plus 25% of the tower diameter). The shell thickness is determined from Table E-1 for the expected internal operating pressures and temperatures. In many cases a corrosion allowance of 1/8 to 1/4 inch is also added to the minimum thickness for carbon steel construction. The cost of the fabricated vessel alone, therefore, includes the cost of the shell plus the cost of the inlet and outlet. The fabricated cost of a

*Note: Cost estimating data for venturi scrubbers is given in Appendix D of the Air Pollution Training Institute's Course 413 Manual, *Control of Particulate Emissions.* EPA 450/2-80-066.

skirt which is provided for support of the vessel and flange-type nozzles for shell penetrations must also be added to the vessel cost. Figure E-6 estimates the cost of skirts which include a base plate, anchors, an 18 in. diameter access opening, two reinforced pipe openings and a vent hole. The thickness of the skirt should be approximately equal to the thickness of the shell. Figure E-7 estimates the fabricated cost of nozzles for typical shell penetrations such as those required for the inlet and outlet for the gas and liquid, relief valve connections, pumpout alternates, spray inlets, and manways.



Stainless steel	
Type	Cost adjustment
304	2.0
316	2.3

Figure E-5. Fabricated cost of carbon steel vessel.

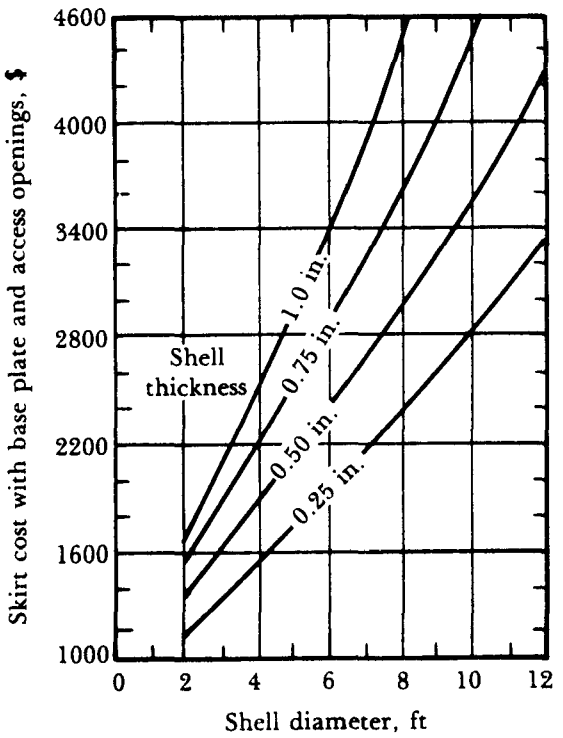
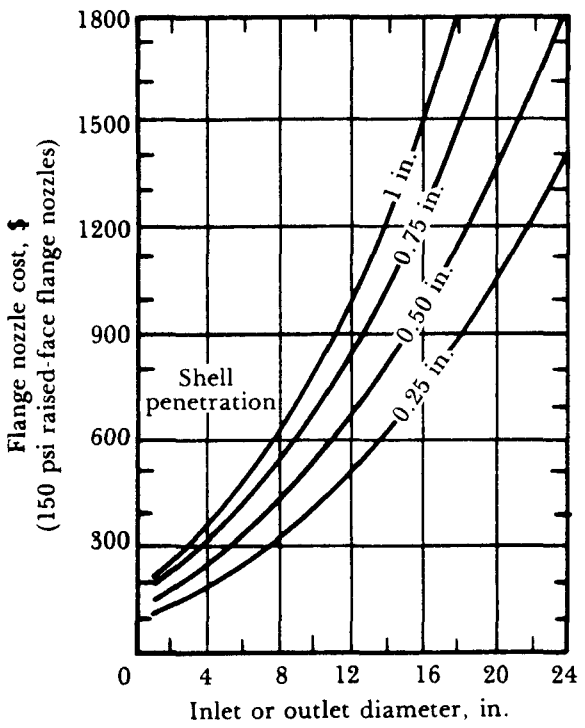


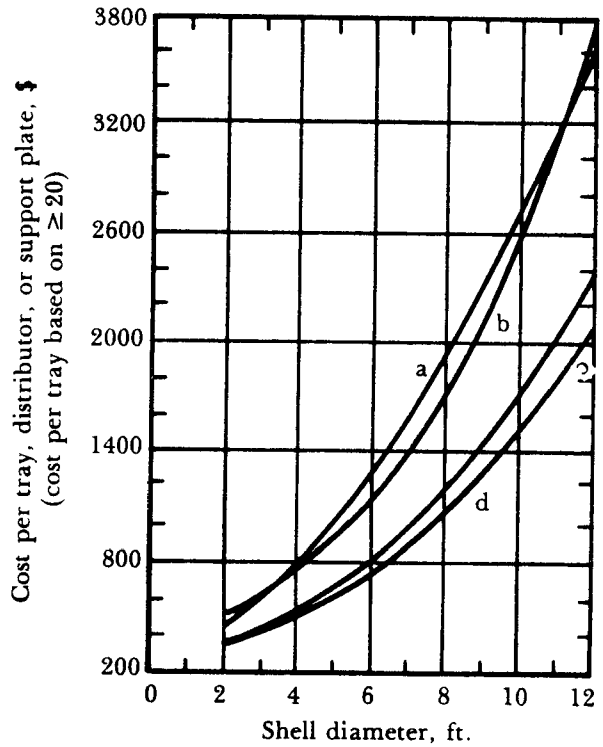
Figure E-6. Skirt and support costs for carbon steel vessel.



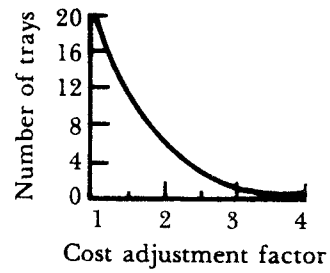
For manways and handholes add 50% for cost of blind flange, gaskets, hinges or davit.

Stainless steel	
Type	Cost adjustment
304	2.0
316	2.3

Figure E-7. Cost of nozzles.



For trays only, cost per tray must be multiplied by cost adjustment factor below.



Stainless steel support plate or distributor	
Item	**Cost factor**
Support plate	1.8
Distributor	1.3

Where: a = carbon steel distributor
b = 304 stainless steel tray
c = carbon steel support plate
d = carbon steel tray

Figure E-8. Cost of tray, support plate, or distributor.

Table E-1. Minimum shell thickness at ambient temperature (carbon steel).

Shell diameter* (ft)	2 (in.)	4 (in.)	6 (in.)	8 (in.)	10 (in.)	12 (in.)
Internal pressure:						
Atmospheric	1/8	1/8	1/8	1/8	1/8	1/8
25 in. WG	1/8	1/8	1/8	1/8	3/16	3/16
50 in. WG	1/8	1/8	3/16	3/16	1/4	1/4
100 in. WG	1/8	1/8	3/16	1/4	5/16	5/16
10 psig	1/8	1/8	1/4	3/8	7/16	1/2
100 psig	1/8	1/4	3/8	1/2	3/4	3/4
200 psig	1/4	1/2	3/4	1	1 1/4	1 1/2

*For corrosion allowance add minimum of 1/8 inch. Thicknesses are for ambient temperatures and internal temperatures up to 600°F. Thickness correction factors for higher temperatures are: 1.04 for 700°F; 1.14 for 750°F; 1.35 for 800°F.

The cost of internal tower equipment such as support plates, trays, and distributors is shown in Figure E-8. The indicated prices apply to both tray towers and packed towers and represent the installed cost per item. The cost of trays is based on quantities of 20 or more. For quantities less than 20, the unit tray cost must be multiplied by the cost adjustment. The cost per cubic foot of internal packing is shown in Table E-2.

Table E-2. Cost of tower packing.

Type and material	1 in. ($/ft³)	1 1/2 in. ($/ft³)	2 in. ($/ft³)	3 in. ($/ft³)	3 1/2 in. ($/ft³)
Pall rings:					
Carbon steel	19	15	13	—	—
304 stainless steel	54	43	35	—	—
Polypropylene	14	—	8	—	—
Intalox saddles:					
Polypropylene	15	—	10	5	—
Porcelain	12	—	9	8	—
Raschig rings:					
Carbon steel	17	14	9	—	—
Porcelain	8	7	6	—	—

The summation of these costs represent the fabricator's shop costs. Added to this must be the fabricator's cost of engineering, administration costs, and profit as determined from Table E-3.

Table E-3. Additional costs for fabricator's engineering, purchasing, administration and profit.

Total cost of fabricated vessel	Cost factor
Less than $5,000	0.25
$5,000 to $10,000	0.23
$10,000 to $20,000	0.20
$20,000 to $30,000	0.19
$30,000 to $50,000	0.18
$50,000 to $80,000	0.17
Over $80,000	0.16

As an example, the design calculations for a carbon steel packed tower indicate that the tower diameter should be approximately 2 ft with a tower height of approximately 15 ft. The inlet gas flow rate is 520 cfm and the liquid flow rate is 3 gpm. The tower packing is to be 1 in. carbon steel Raschig rings and the absorber is to be operated at ambient conditions of temperature and pressure.

Under these conditions, a 0.25 in. plate thickness for the shell may be selected which would provide adequate allowance for corrosion problems. A suitable inlet gas velocity for this absorber would be approximately 1800 fpm which would require a gas inlet and outlet of approximately 8 in. diameter. A suitable liquid

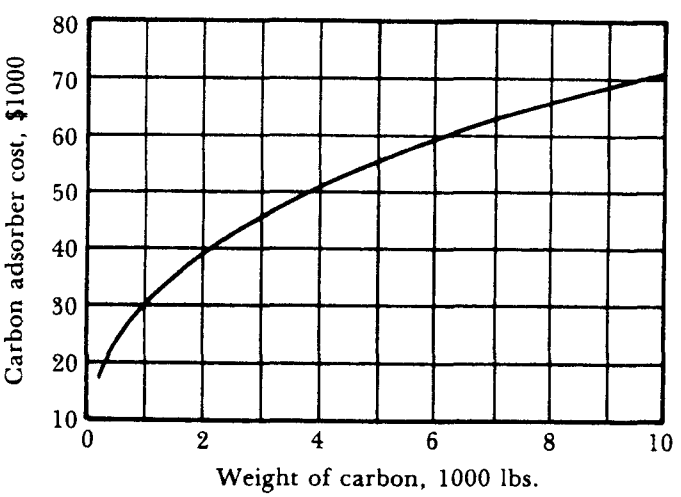
inlet and outlet pipe for 3 gpm would be approximately 1 in. A manway of 18 to 24 in. diameter would normally be provided at the top and bottom for larger diameter towers, however, for this small diameter tower, a 12 in. hand hole can be provided at each end of the tower. A drain in the bottom of the tower should also be provided for draining, flushing and cleanout.

Using Figure E-5 the cost of the basic vessel is $420 for the heads plus $52 per foot of shell or $1200. The cost of the skirt, as determined from Figure E-6, is $1140. The installation of nozzles for the various inlet and outlet diameters results in a cost of $2580 for three liquid nozzles at $120 each, two gas inlet/outlet (at 8 in.) at $330 each, and two hand holes with blind flanges at $780 each (Figure E-7).

The tower internals would consist of a packing support plate and two distributors spaced at 5 ft intervals plus the internal packing material. The cost of the support plate and distributors, as determined from Figure E-8, is estimated to be $1220. The volume of packing required for a 15 ft tower with a 2 ft diameter is approximately 47 ft^3 and the estimated cost of the packing, as determined from Table E-2, is approximately $800. The total fabricated cost of the packed tower is therefore $6940. The fabricator's selling price including engineering, administration costs, and profit, but less taxes and freight, is determined from Table E-3 to be $6940 × 1.23 or $8540.

Adsorber

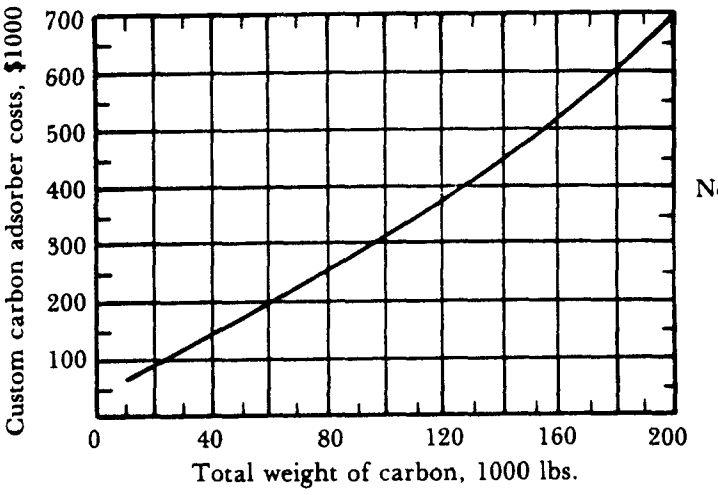
Prices for carbon adsorbers are presented in Figures E-9 and E-10 as a function of total pounds of carbon in the unit. The total or gross number of pounds is determined by the adsorption rate and the regeneration rate of the carbon for the emission being controlled. A carbon adsorber will normally be a dual system with one bed on-line adsorbing while the second bed will be off-line regenerating. A likely estimate of regeneration time for almost all applications would be between 30 minutes to an hour.



Note:
- These prices are for basic systems: including adsorber, carbon, blower or fan, controls and steam regenerator. Special requirements may double the cost.
- These prices are for commercial application. Add 30% for applications involving stringent industrial standards.

Figure E-9. Prices for packaged stationary bed carbon adsorption units with steam regeneration.

Figure E-9 represents packaged units for automatic operation in commercial and industrial applications. Commercial applications cover dry-cleaning and solvent metal cleaning. Industrial applications include lithography and petrochemical processing. Industrial requirements would increase costs 30% over commercial requirements. Industrial requirements would include heavier materials for high steam or vacuum pressure designs and more elaborate controls to assure safety against explosions and prevent hydrocarbon breakthrough. Figure E-10 presents custom units, mostly for industrial applications where the gas flow rate exceeds 10,000 acfm. Table E-4 is provided to estimate annualized control cost requirements for steam, cooling water, maintenance, electricity, and carbon replacement. The carbon weight used to determine the purchase cost can also be used to estimate the carbon replacement cost.



Note:
- Instrumentation and controls included.

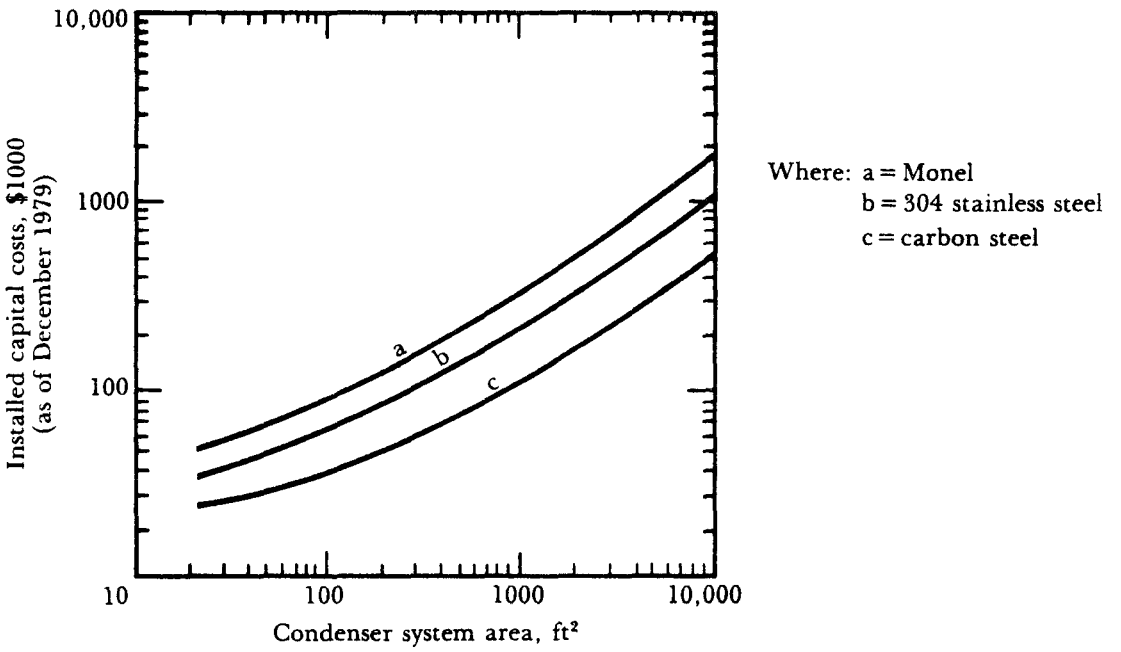
Figure E-10. Prices for custom carbon adsorption units.

Table E-4. Technical assumptions for estimation of direct operating costs.

Item	Assumption
Steam consumption	4 lb per lb pollutant recovered
Cooling water	12 gal per 100 lbs steam
Electricity	5 HP per 1000 acfm
Maintenance	5% of equipment purchase cost
Carbon replacement	Replace original carbon every five years

Condensers

Estimated capital costs are for a complete condenser system including a shell-and-tube condenser, a storage tank, a pump, and the necessary piping are presented in Figure E-11. The shell-and-tube condenser is a fixed-tube type with 1 inch diameter tubes that are 8 feet long. Costs are based on December 1979 and include a 30% contingency allowance. Figure E-11 was taken from an EPA publication as part of a study for controlling emissions from the Organic Chemical Manufacturing Industry (EPA, 1980).



Source: EPA 1980; Organic Chemical Manufacturing.

Figure E-11. Installed capital costs vs. condenser area for a complete condenser section.

References

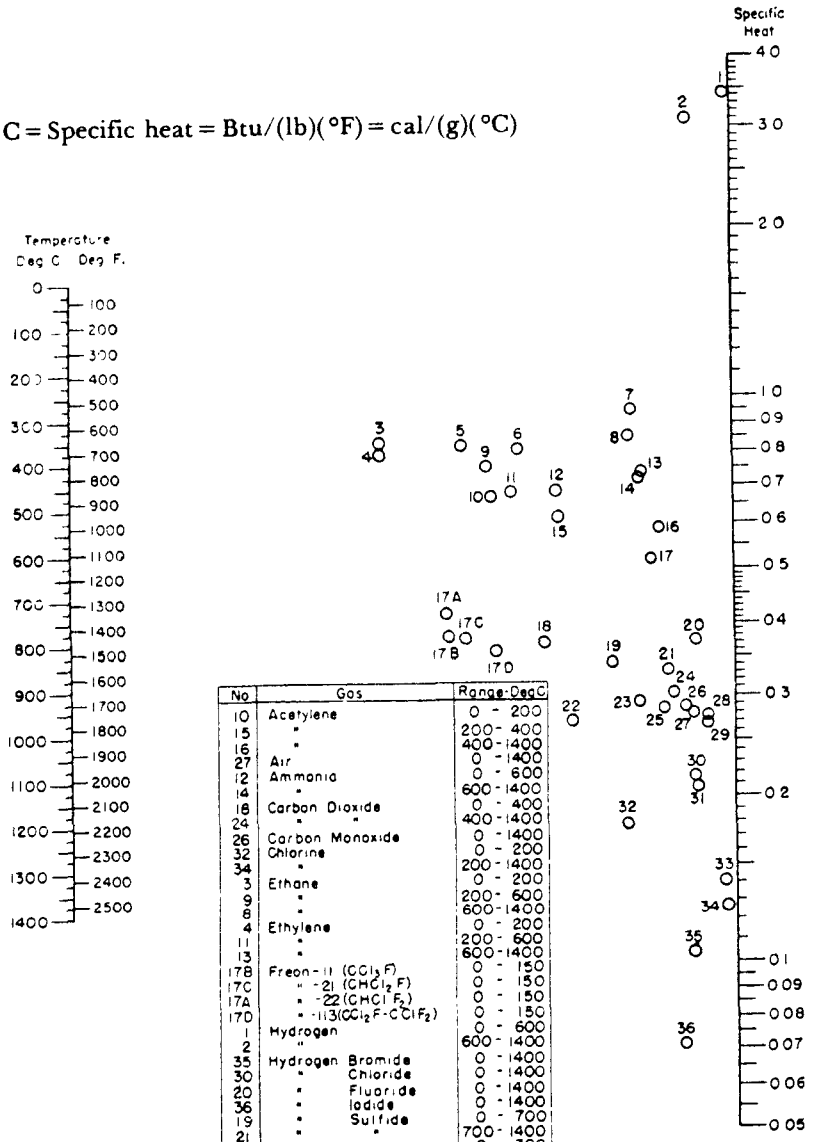
Environmental Protection Agency (EPA). 1978. *Capital and Operating Costs of Selected Air Pollution Control Systems*. EPA 450/5-80-002. Research Triangle Park, NC.

Environmental Protection Agency (EPA). 1980. *Organic Chemical Manufacturing, Vol. 5: Adsorption, Absorption, and Condensation Devices*. EPA 450/3-80-027. Research Triangle Park, NC.

Appendix F

Specific Heats (C_p) of Gases at 1 atm Pressure

C = Specific heat = Btu/(lb)(°F) = cal/(g)(°C)



No	Gas	Range-DegC
10	Acetylene	0 - 200
15	"	200 - 400
16	"	400 - 1400
27	Air	0 - 1400
12	Ammonia	0 - 600
14	"	600 - 1400
18	Carbon Dioxide	0 - 400
24	" "	400 - 1400
26	Carbon Monoxide	0 - 1400
32	Chlorine	0 - 200
34	"	200 - 1400
3	Ethane	0 - 200
9	"	200 - 600
8	"	600 - 1400
4	Ethylene	0 - 200
11	"	200 - 600
13	"	600 - 1400
17B	Freon-11 (CCl_3F)	0 - 150
17C	" -21 ($CHCl_2F$)	0 - 150
17A	" -22 ($CHClF_2$)	0 - 150
17D	" -113 (CCl_2F-$CClF_2$)	0 - 150
1	Hydrogen	0 - 600
2	"	600 - 1400
35	Hydrogen Bromide	0 - 1400
30	" Chloride	0 - 1400
20	" Fluoride	0 - 1400
36	" Iodide	0 - 1400
19	" Sulfide	0 - 700
21	" "	700 - 1400
5	Methane	0 - 300
6	"	300 - 700
7	"	700 - 1400
25	Nitric Oxide	0 - 700
28	" "	700 - 1400
26	Nitrogen	0 - 1400
23	Oxygen	0 - 500
29	"	500 - 1400
33	Sulfur	300 - 1400
22	Sulfur Dioxide	0 - 400
31	" "	400 - 1400
17	Water	0 - 1400

Example: To find the specific heat of air at 300°C, lay a straight edge across paper from 300°C through the circle for 27 and read off specific heat scale as 0.255 cal/g °C.

Source: From *Chemical Engineers' Handbook* by R. Perry and C. Chilton.
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Appendix G

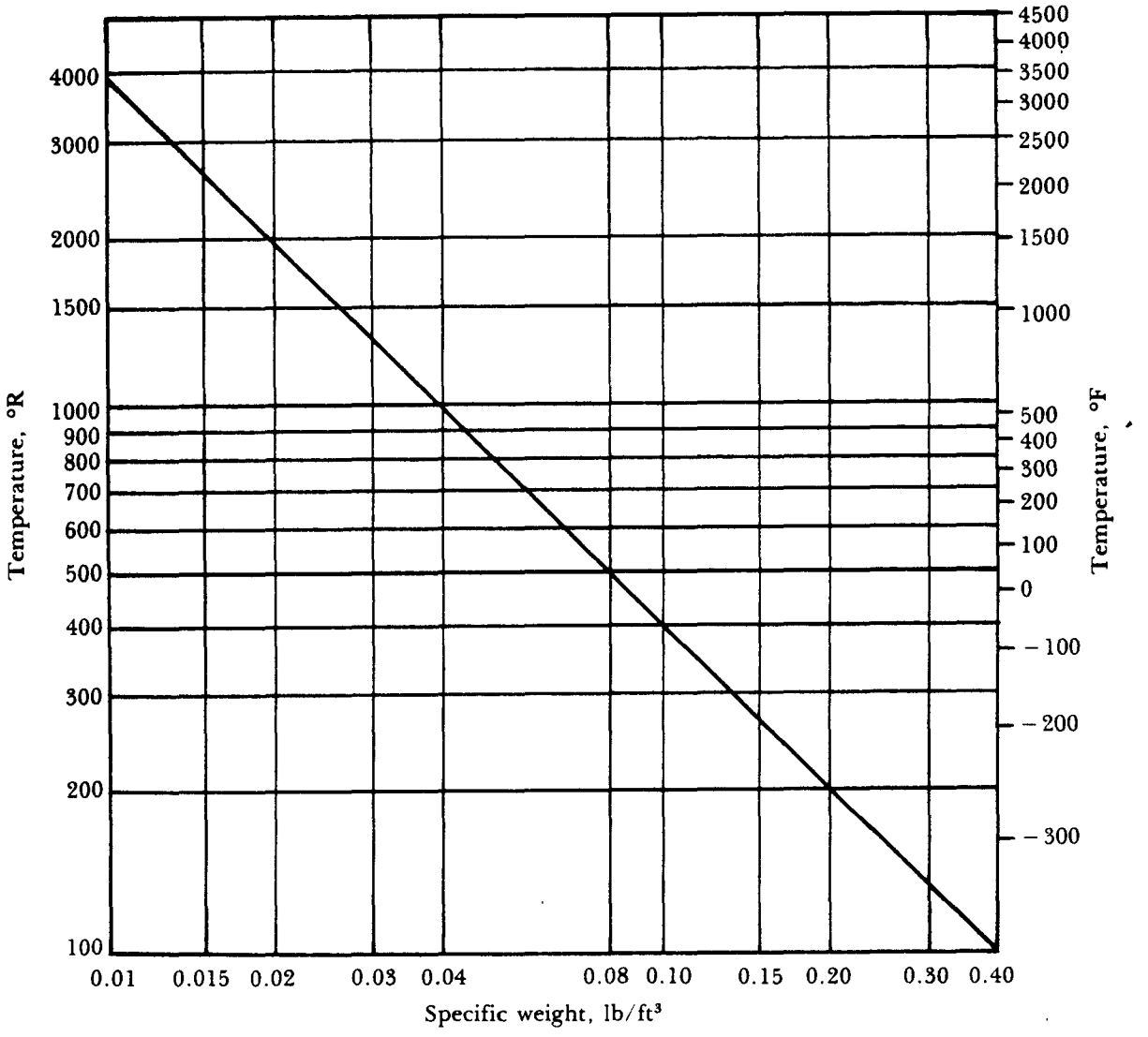
Properties of Liquid Water

Temperature, T (°F)	Viscosity, μ (centipoises)	Density, ϱ (lb/ft^3)
32	1.794	62.42
40	1.546	62.43
50	1.310	62.42
60	1.129	62.37
70	0.982	62.30
80	0.862	62.22
90	0.764	62.11
100	0.682	62.00
120	0.559	61.71
140	0.470	61.38
160	0.401	61.00
180	0.347	60.58
200	0.305	60.13
220	0.270	59.63
240	0.242	59.10
260	0.218	58.51
280	0.199	57.94
300	0.185	57.31

Source: From *Unit Operations of Chemical Engineering* by W. McCabe and J. Smith. ©1967 McGraw-Hill. Used with the permission of McGraw-Hill Book Company.

Appendix H

Specific Weight of Dry Air in lb/ft³ for °F and °R and Absolute Pressure of 29.92 in. Hg.



Appendix I

Use of Psychrometric Chart for Gas Moisture Calculation

Psychrometry is defined as the science dealing with determination of the properties of gas and vapor mixtures. Psychrometric charts are graphical representations of the properties of a gas-vapor mixture. By far, the most important gas-vapor mixture is that of air and water. Many forms of psychrometric charts (or humidity charts) are available for determining the properties of air and water mixtures. Figure I-1 is an example of a psychrometric chart. The following example illustrates the use of this chart in determining the properties of air at different temperatures and degrees of saturation.

Problem Statement

A coal fired utility boiler uses a combination spray and venturi scrubber to reduce particulate matter and SO_2 gas to acceptable levels. Flue gas at a flow rate of 98,750 acfm leaves the economizer section of the boiler at 305°F. The flue gases are ducted into the spray section of the scrubber where the gases are cooled and saturated, and then they flow into the venturi. A wet bulb/dry bulb temperature measure of 110°F/305°F was taken of the gases entering the scrubber.

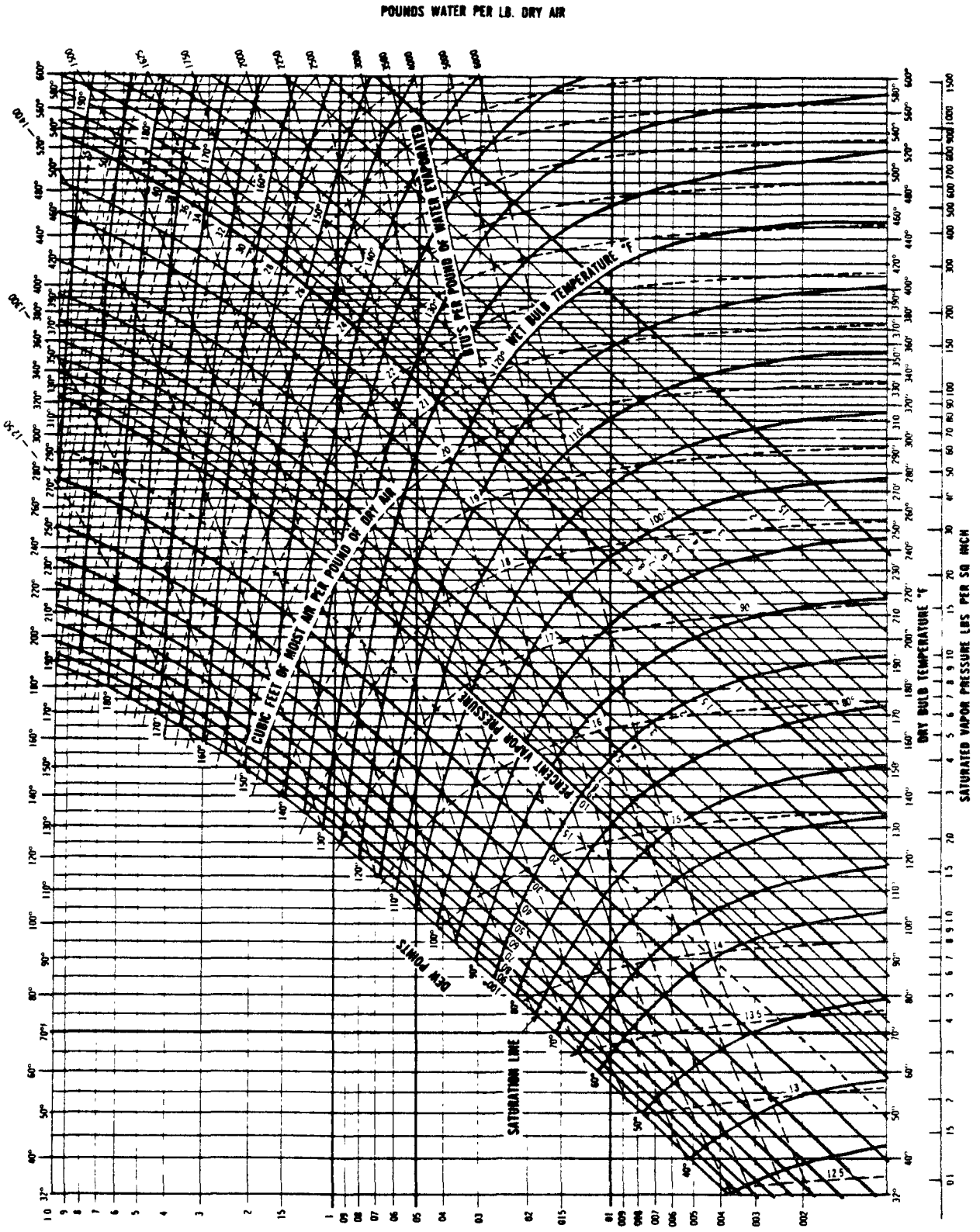
Calculate:
1. saturation temperature and humidity of flue gases entering the spray section
2. saturation humidity
3. evaporation rate
4. saturated volumetric flow
5. percent gas flow reduction as it passes through the scrubber system

Solution:

1. The humidity is read from Figure I-1.

   The humidity is determined by starting at the dry bulb temperature of 305°F, following this line up until it crosses the wet bulb line of 110°F and then reading the humidity as **0.012 lb water/lb dry air**.

   The saturation temperature is the wet bulb temperature, **110°F**. An air stream can be cooled and saturated no lower than its wet bulb temperature.



Source: From *Chemical Engineers' Handbook* by R. Perry and C. Chilton. ©1973 McGraw-Hill. Used with the permission of McGraw-Hill Book Company.

Figure I-1. Psychrometric chart. Properties of air and water-vapor mixtures from 32° to 600° F.

2. The saturation humidity of the flue gases is determined by following the curved wet-bulb temperature line until it hits the saturation line and reading off the humidity at this point, **0.058 lb water/lb dry air.**

3. The evaporation rate of the scrubbing water is the difference between the incoming and saturation humidity:

$$\begin{array}{r} 0.058 \text{ lb water/lb dry air} \\ -\,0.012 \text{ lb water/lb dry air} \\ \hline 0.046 \text{ lb water/lb dry air} \end{array}$$

To determine the water make up rate, first compute the lb of dry air (flue gas) coming into the scrubber. The ft^3 of flue gas can be converted to lb of moist air by reading the volumetric flow from Figure I-1. For a dry bulb/wet bulb reading of 305°F/110°F this is:

$$19.75 \frac{ft^3 \text{ of moist air}}{\text{lb dry air}}$$

The total lb of dry air entering the scrubbing system is:

$$98{,}750 \frac{ft^3}{min} \times \frac{\text{lb dry air}}{19.75\ ft^3 \text{ moist air}} = 5000 \frac{\text{lb dry air}}{min}$$

The water evaporation rate is:

$$5000 \frac{\text{lb dry air}}{min} \times \frac{0.046 \text{ lb water}}{\text{lb dry air}} \times \frac{60 \text{ min}}{hr} = 13{,}800 \frac{\text{lb water}}{hr}$$

4. The saturation volumetric flow rate is read from Figure I-1 where the wet bulb line (110°F) hits the saturation line. These are the curved, dashed lines; the saturation volumetric flow rate is:

$$15.8 \frac{ft^3 \text{ of moist air}}{\text{lb of dry air}}$$

5. To calculate the gas flow reduction first calculate the flow leaving the scrubber system. The volume of gas leaving the scrubber is saturated, therefore:

$$5000^* \frac{\text{lb dry air}}{min} \times \frac{15.8\ ft^3 \text{ of moist air}}{\text{lb dry air}} = 79{,}000 \text{ acfm}$$

$$\%\ \text{reduction} = \frac{98{,}750 - 79{,}000}{98{,}750} = 20.0\%$$

*Note: The volumetric flow of dry air through the system is constant.

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(Please read Instructions on the reverse before completing)

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16. ABSTRACT

The Student Manual is to be used in conducting APTI Course 415 "Control of Gaseous Emissions". This manual supplements the course lecture material, presenting detailed discussion on gaseous emission control equipment. The major topics include: Basic Gas Properties, Combustion, Absorption, Adsorption, Condensation, and Control of NO_x and SO_2 Emissions. This manual will assist the reader in evaluating plans for gaseous emission control systems and in conducting plan reviews.

This manual is intended for use in conjunction with the Instructor's Guide (EPA 450/2-81-004) and the Student Workbook (EPA 450/2-81-006) for APTI Course 415.

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