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Final Report, Vol. I

by
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This manual presents a detailed design review procedure for the complete range of gaseous and particulate control equipment applicable to air pollution sources at army installations. It includes identification of emission sources, estimation of emission rates, step-by-step design review procedure for available control devices, and economic evaluation of various control alternatives. A case study is presented to explain the use of this manual. Appendices to this report are provided in Volume II.

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SUMMARY

This manual is designed to familiarize the engineer with all facets of an air pollution abatement program for the Army installations. Various air pollution emission sources encountered in the Army operations are classified into four major categories as follows:

1. steam generation
2. solid waste incineration
3. ammunitions manufacture
4. evaporative loss sources.

Individual processes within a category are identified and described in detail. Types of pollutants resulting from these processes and methods to control them are studied in depth, and guidelines are offered to select the best control system. Applications of necessary control equipment are presented with typical source operating and emission characteristics. Details of control equipment design and evaluation also include the economic analysis for an accurate assessment and selection.

This manual is broad in scope, yet contains the procedures that make an in-depth individual evaluation possible. Moreover, standardized sheets have been prepared for the Army emission sources and the control equipments. This, combined with the loose-leaf binding, should make the future expansion of this manual very easy and convenient. New emission sources and/or control equipments can be added without any difficulty.

To effectively approach an air pollution problem and use this manual to solve the same, it is important to follow a step-by-step procedure. First, it is necessary to have complete information on the emission source, its characteristics, the amount and nature of particulate and gaseous emissions, and the operating conditions. Study of the regulations then suggests if the emissions are in compliance with the existing laws. If not, the control system is used to control the pollutant that is out of compliance. Depending upon the nature of the pollutant and the regulatory needs, available alternatives are considered and the most economical one is selected. The entire systematic approach is diagrammed in Figure S-1.

Table S-1 summarizes the Army emission sources and applicable control equipment.

FIGURE S-1
SYSTEMATIC APPROACH TO USE OF DESIGN REVIEW PROCEDURES MANUAL

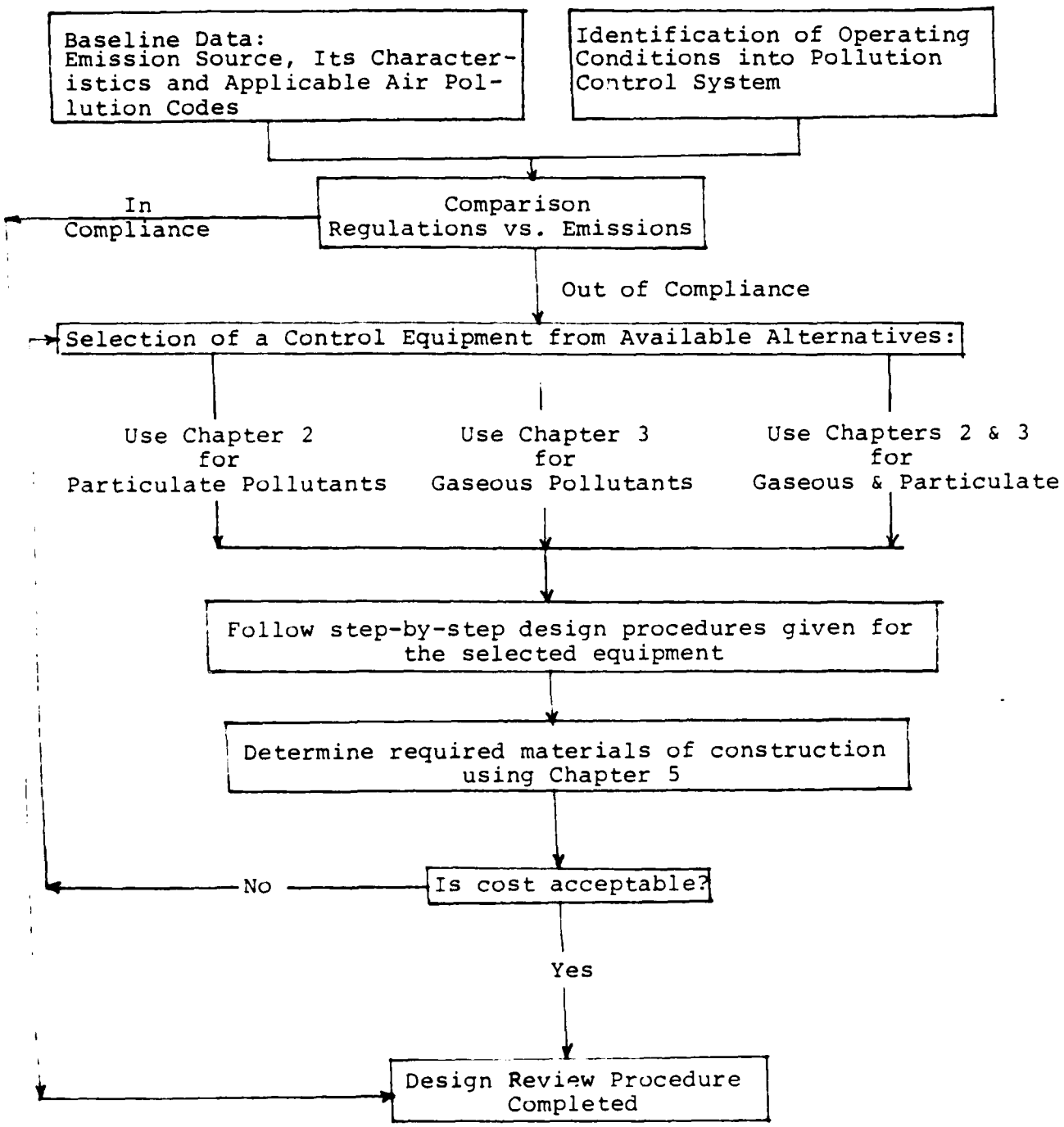


TABLE S-1
AIR POLLUTION CONTROL EQUIPMENT FOR SELECT ARMY APPLICATIONS

Classification	Process	Settling Chamber	Wet Scrubber	Cyclones	Electro-static Precipitator	Incinerator	Fabric Filter	Alternative Methods	Section Reference
Steam Generation	Boilers	x	x	x	x	-	x	Process Control	1.1
Solid Waste Incineration	General Refuse Disposal	-	x	-	x	x	x	---	1.2.1
	Pathological/Medical	x	x	x	x	x	x	---	1.2.2
	Explosives	-	x	x	-	x	x	Flame Arrestor	1.2.3
Ammunition Manufacture	TNT Production	-	x	-	-	x	-	Mist Eliminators, Condensers, Adsorption, Absorption	1.3.1
	Sulfuric Acid Plant	-	x	-	x	-	-	Mist Eliminators Process Change	1.3.2
	Nitric Acid Plant	-	x	-	x	x	-	Mist Eliminators Adsorption, Absorption	1.3.3
Evaporation Loss Sources	Metal Treating and Degreasing	-	x	-	-	x	-	Condensation, Carbon Adsorption	1.4.1
	Metal Coating	-	x	-	-	x	-	Carbon Adsorption	1.4.2
							-	Dry Baffle	
							-	Paint Arrestor Water Wash	
	Liquid Fuel Loading and Unloading	-	-	-	-	-	-	Condensation, Refrigeration, Carbon Adsorption	1.4.3

S.1 EMISSION SOURCES

This section summarizes the air pollution emission sources in the Army operations. Each source is briefly described in a standardized pattern that includes processes causing the pollution, their flow diagrams, the actual pollutants emitted, and their typical emission factors. This provides a clear picture of overall air pollution problems at the Army installations. A blank sheet in a standard format is provided for easy and convenient addition of new emission sources to this section in the future.

Emission Source: Steam Generation

Process: Boilers and Heating Plants

Pollutants: Particulate: Fly Ash
Gaseous: Sulfur Dioxide
Nitrogen Dioxide, Carbon
Monoxide, Hydrocarbons

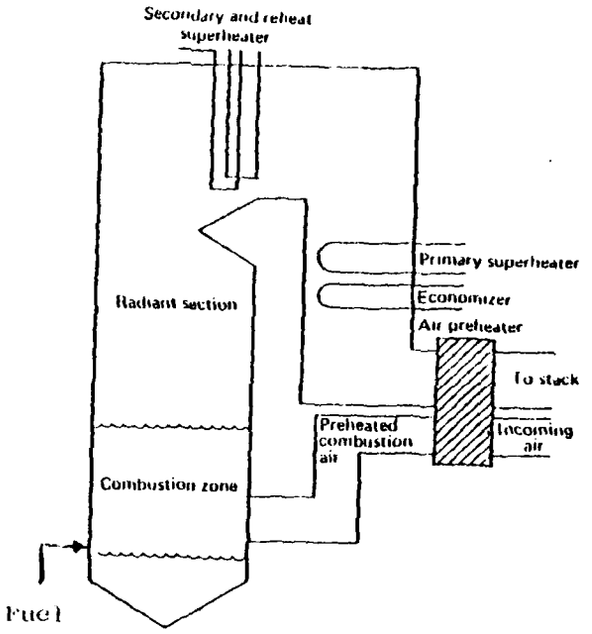
Summary of Process:

As illustrated in the flow sheet, boilers use the heat of combustion of coal, oil, or gas to generate steam. Products of combustion, particulate matter, sulfur dioxide, nitrogen dioxide, carbon monoxide, and hydrocarbons exit with the flue gas and may be discharged to atmosphere through the stack. These emissions are characterized by emission factors as described by Tables 1-1, 1-2, and 1-3. The emission factors for major pollutants for coal and oil are given below:

	Emission Factors	
Pollutant	Coal	Oil
Particulate	(5-16)A*	7-10
Sulfur Dioxide	38S*	2S*
Nitrogen Dioxide	15 to 18	105

The emission factor gives the ambient of pollutant discharged; for coal, lb/ton of coal fired; for oil, lb/1,000 gal of oil fired.

*A is the % wt of fly ash and S is the % wt of sulfur in the fuel. Both sulfur and ash content of fuels may be controlled to reduce emissions. Natrual gas fuel is environmentally clean when combusted and does not require air pollution control.



SCHEMATIC DIAGRAM OF UTILITY STEAM GENERATOR

YORK RESEARCH CORPORATION

Emission Source:	Reference:
STEAM GENERATION	SECTION 1.1

Emission Source: Solid Waste Incineration

Processes: (1) General Refuse Incineration
(2) Pathological Wastes
(3) Explosive Wastes

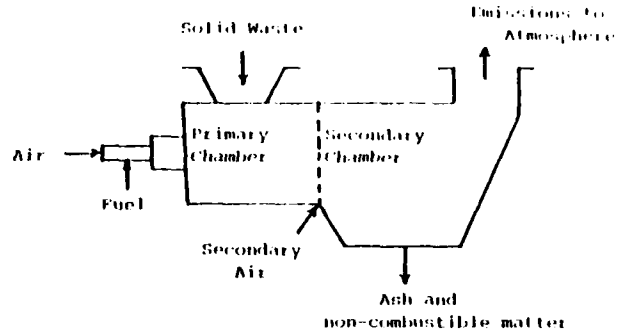
Pollutants: Particulate: Fly Ash
Gaseous: Sulfur Dioxide, Hydrogen Chloride, Hydrocarbons

Summary of Processes:

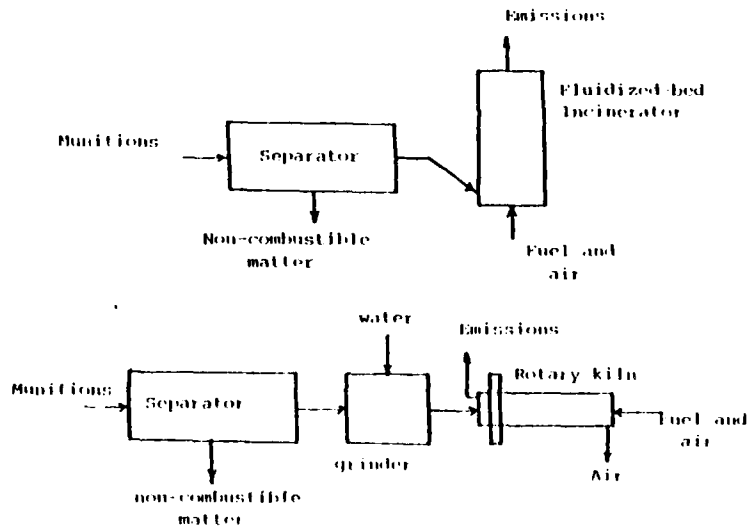
Solid waste incinerators are classified by the type of refuse burned:

(1) General: paper, cardboard, and other combustible matter.
(2) Pathological: hospital waste including organic waste and animal origin.
(3) Explosive: propellants, explosive and pyrotechnics.

General and pathological incinerators are of the multi-chamber type, which includes a primary combustion chamber and a secondary chamber where additional air and secondary fuel are used to complete combustion. Pollutants discharged are characterized by the types of waste matter burned. These are indicated in emission factor Tables 1-18 and 1-20. For explosive wastes, emission factors are not developed and are treated case by case as shown by Tables 1-23 and 1-26. The hazardous nature of disposal requires the incinerator to be constructed for detonations.



(a) TYPICAL GENERAL AND PATHOLOGICAL INCINERATOR



(b) EXPLOSIVE WASTE INCINERATORS

YORK RESEARCH CORPORATION

Emission Source:	Reference:
SOLID WASTE INCINERATION	SECTION 1.2

Emission Source: Ammunitions Manufacture

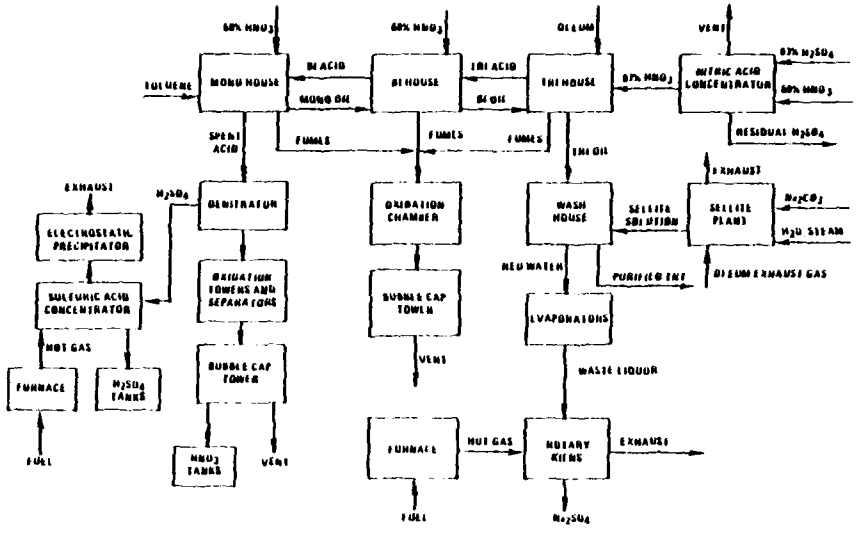
Process: Explosive (TNT) Production

Pollutant: Particulate: Sulfuric Acid and Nitric Acid Mist
Gaseoue: Sulfur Dioxide, Nitrogen Dioxide

Summary of Process:

TNT, trinitrotoluene, is the product of multiple-stage nitration of toluene with a mixture of sulfuric acid and nitric acids. Four processes within the plant are characterized as emission sources. These are the ammonia oxidation plant, AOP; the acid fume recovery system, AFR; direct strong nitric acid process, DSN; and sulfuric acid regeneration, SAR. These emissions are tabulated below:

Source	Emissions, lb/hr	Reference
AOP	NOx, 475-525 for 383 tons/day of 60% HNO_3 produced	Section 1.3.1.3
AFR	NOx, 18.42 for 9,427 lb/hr of 60% HNO_3 produced	Section 1.3.1.4
DSN	NOx, 19.46 for 340 tons/day of 60% HNO_3 produced	Section 1.3.1.5
SAR	NOx 106.6, SO_2 90.7, H_2SO_4 mist 25 lb/hr for 580 tons of 100% H_2SO_4 produced	Section 1.3.1.6



PROCESS FLOW DIAGRAM - TNT PRODUCTION

YORK RESEARCH CORPORATION	
Emission Source: AMMUNITIONS MANUFACTURE	Reference: SECTION 1.3.1

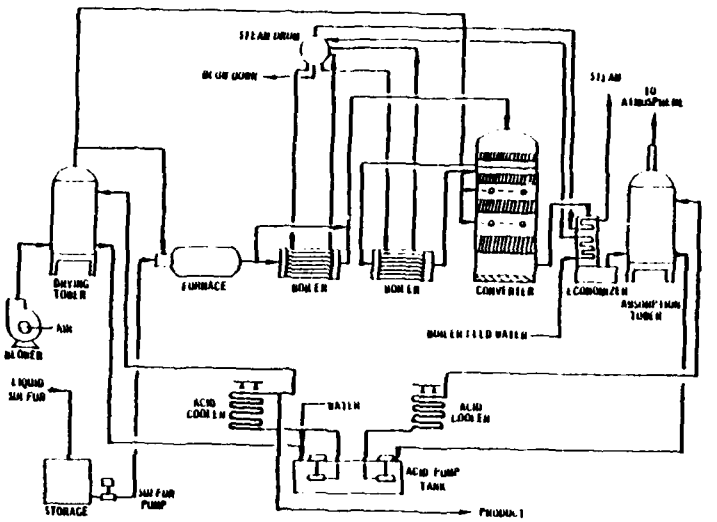
Emission Source: Ammunitions Manufacture

Process: Sulfuric Acid Plant

Pollutants: Particulate: Sulfur Acid Mist
Gaseous: Sulfur Dioxide

Summary of Process:

Sulfur is burned to produce sulfur dioxide, which is converted catalytically to sulfur trioxide. This gas is absorbed in water to produce 98 to 99% sulfur acid. Vapors vented to atmosphere emit acid mist and sulfur dioxide. Table 1-37 shows the emission factors for sulfur dioxide based on the percent of conversion for SO_2 and SO_3. The amount of acid mist vapor emitted is shown in Table 1-38.



TYPICAL SULFURIC ACID PLANT

YORK RESEARCH CORPORATION

Emission Source:	Reference:
AMMUNITIONS MANUFACTURE	SECTION 1.3.2

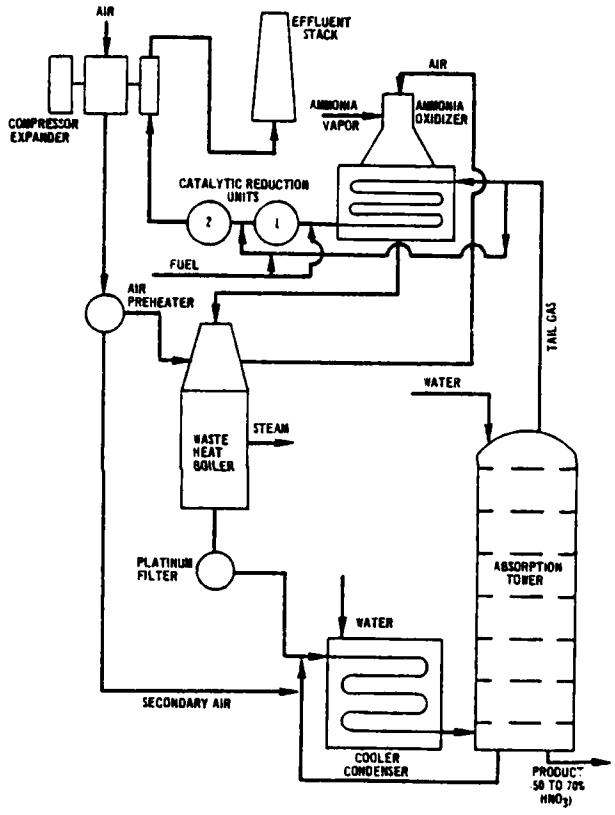
Emission Source: Ammunitions Manufacture

Process: Nitric Acid Plant

Pollutants: Gaseous: Nitrogen Oxides

Summary of Process:

Ammonia is catalytically oxidized to nitric oxide and then to nitrogen dioxide. In this form it is absorbed in water to produce 50-70% nitric acid. Emissions of NOx are shown in Table 1-42.



TYPICAL NITRIC ACID PLANT

YORK RESEARCH CORPORATION

Emission Source:	Reference:
AMMUNITIONS MANUFACTURE	SECTION 1.3.3

Emission Source: Evaporative Loss Sources

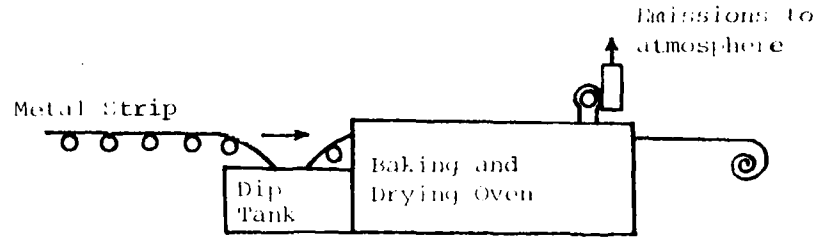
Processes: (1) Metal treating and Degreasing
(2) Metal Coating
(3) Liquid Fuel Loading and Unloading

Pollutants: Gaseous: Solvent Vapors, Coating, Fuel Vapors

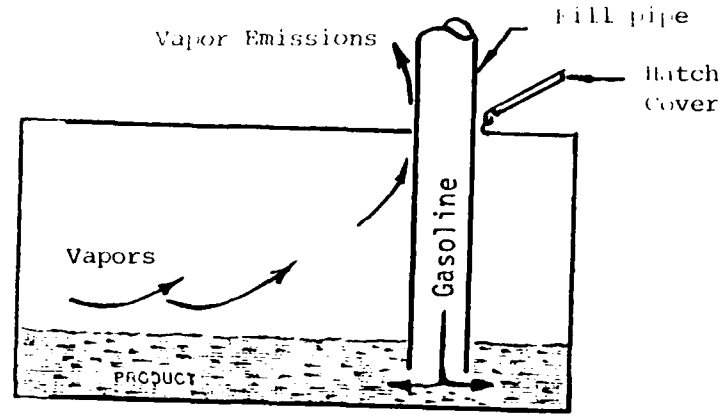
Summary of Processes:

Metals are cleaned by degreasing with hydrocarbons. The solvents utilized may vaporize into the atmosphere. Metal coating operations include spraying, dip-coating, flour coating, baking, and drying. Solvent vapors are emitted into the atmosphere as the coatings are baked and dried. As liquid fuels are transferred from storage, hydrocarbons may evaporate from the fuels.

The emissions are characterized by the types of volatile hydrocarbons processed. The emission of vapor from the degreaser is 1.5 lb/ton of metal cleaned as shown by Table 1-43. For metal coating operations, the amount of hydrocarbon solvents emitted varies according to the type of coating applied from 840 to 1,540 lb/ton of coating as shown in Table 1-45. Hydrocarbon vapors emitted during fuel transfer are about 5 lb/1,000 gallons of fueld as indicated in Table 1-46.



METAL COATING



SUBMERGED TANK TRUCK LOADING

YORK RESEARCH CORPORATION

Emission Source:	Reference:
EVAPORATIVE LOSS SOURCES	SECTION 1.4

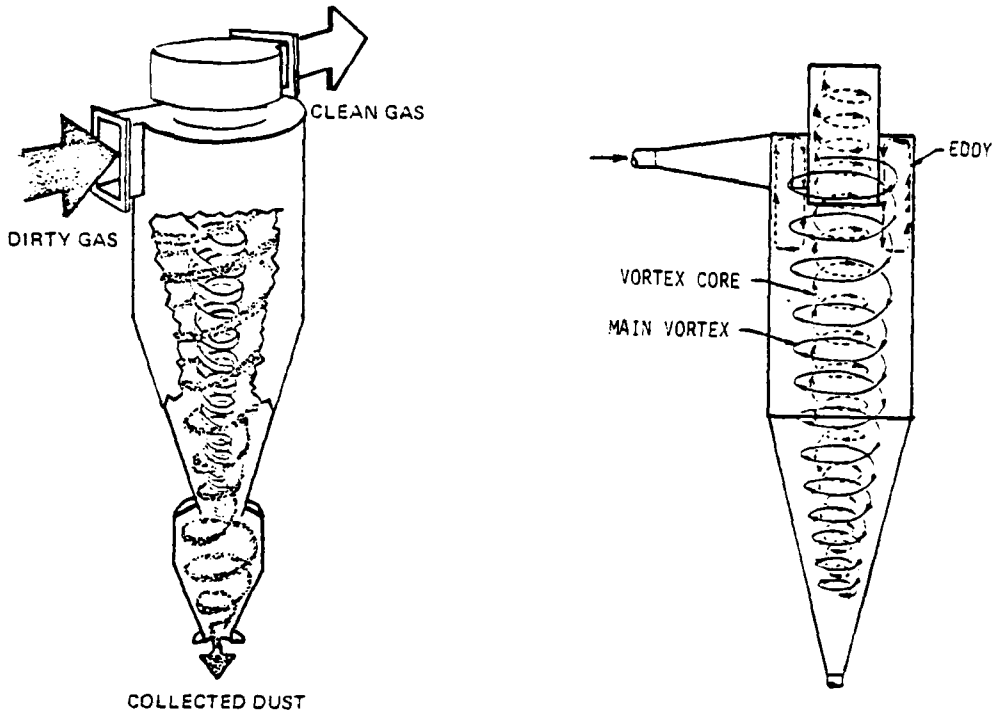
S.2 CONTROL EQUIPMENTS

The following section summarizes the control equipments that are used for air pollution abatement at Army installations and presents them on standardized sheets. Each sheet deals with one type of equipment and includes a picture, short description, typical operating characteristics, advantages, and disadvantages. It provides a very good understanding of the structure and functioning of the equipment. This section also allows the engineer to quickly compare different alternatives for a given air pollution control problem. If necessary, more equipments can be easily added to this section using the standard format provided here.

CONTROL EQUIPMENT: Conventional Cyclones

POLLUTANT TYPE: Particulate

ILLUSTRATION:



Conventional Centrifugal Cyclone

DESCRIPTION:

The cyclone collector is an inertial separator. It consists of a cylinder, a tangential gas inlet, a cone to deliver the collected dust to a central disposal point, and an axial gas outlet. The dirty gas enters tangentially at the top and spirals downward to an outer vortex. Near the bottom of the cone, the gases reverse the direction and begin to move upwards in an inner vortex. The spiraling action of the gases produces sufficient centrifugal force to drive the suspended particulate to the collector wall. These particles then move along the wall towards the dust discharge by the force of gravity and the downward movement of the outer vortex.

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TYPICAL OPERATING CHARACTERISTICS:

1. Conventional cyclones are of medium efficiency (60-80%) and capable of handling high throughput at relatively low pressure losses, typically in the range of 2 to 4 inches of water.
2. Conventional cyclones usually have body diameters ranging from 4 to 12 ft.
3. High-efficiency (80-95%), single cyclone units are generally long and narrow. Body diameters seldom exceed 36 inches and are most often in the range of 12 to 24 inches.
4. High-efficiency, single cyclone units have pressure drops most frequently in the range of 3 to 6 inches of water.

ADVANTAGES:

1. Low cost of construction.
2. Relatively simple equipment with few maintenance problems.
3. Relatively low operating pressure drops (for degree of particulate removal obtained) in the range of approximately 2 to 6 inches water gauge.
4. Temperature and pressure limitations imposed only by the materials of construction used.
5. Dry collection and disposal.
6. Relatively small space requirements.

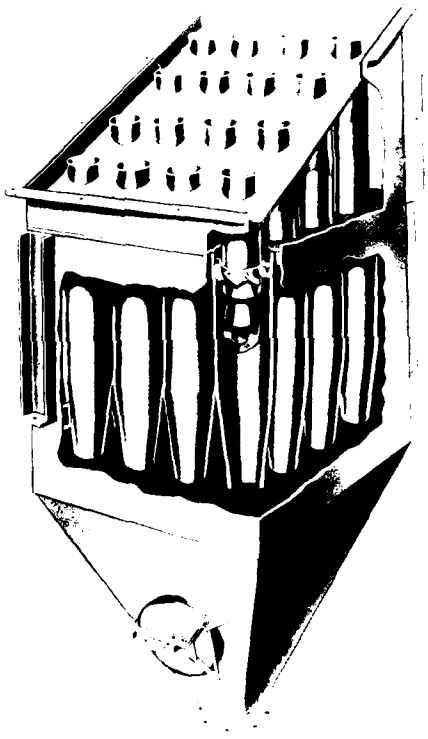
DISADVANTAGES:

1. Relatively low overall particulate collection efficiencies, especially on particulates less than 10 microns in size.
2. Inability to handle tacky materials.

CONTROL EQUIPMENT: Multiple Cyclones

POLLUTANT TYPE: Particulate

ILLUSTRATION:



Multiple Cyclone

DESCRIPTION:

Multiple cyclone collectors usually consist of a number of small diameter cyclones operating in parallel with a common gas inlet and outlet. The flow pattern differs from that of a conventional cyclone; instead of bringing the gas tangentially from the side to initiate the swirling action, the gas is brought in axially at the top of the collection tube and swirled by a stationary vane positioned in the path of the incoming gas.

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Pollutant Type:	Control Equipment:
Particulate	Multiple Cyclones

TYPICAL OPERATING CHARACTERISTICS:

1. Cyclone tube diameters typically range from 6 to 12 inches.
2. Pressure drops are usually in the range of 2 to 6 inches of water.

ADVANTAGES:

1. Collection of smaller particles in the range of 5 to 10 microns with an efficiency as high as 90 percent.

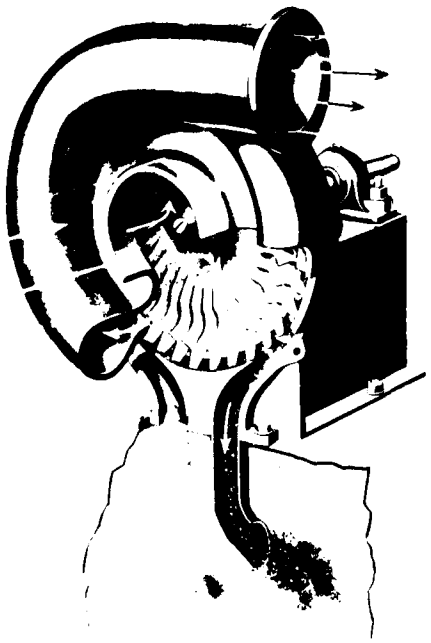
DISADVANTAGES:

1. Plugging and flow equalization problems.

CONTROL EQUIPMENT: Mechanical, Centrifugal Collectors

POLLUTANT TYPE: Particulate

ILLUSTRATION:



DESCRIPTION:

Centrifugal force may also be supplied by a rotating vane. The unit serves both as an exhaust fan and a dust collector. In operationg, the rotating fan blade exerts a large centrifugal force on the particulates, ejecting them from the tip of the blades to a skimmer bypass leading into a dust hopper. Efficiencies of these systems are somewhat higher than those obtained with conventional cyclones.

TYPICAL OPERATING CHARACTERISTICS:

1. Handles gas flows up to 20,000 cfm and temperatures up to 750°F.
2. Pressure drop is about 0.5 inch of water.

YORK RESEARCH CORPORATION

Pollutant Type:	Control Equipment:
Particulate	Mechanical, Centrifugal Collectors

ADVANTAGES:

1. Compact units and small space requirements.
2. Functions both as dust collector and fan.
3. Higher collection efficiencies than conventional cyclones.

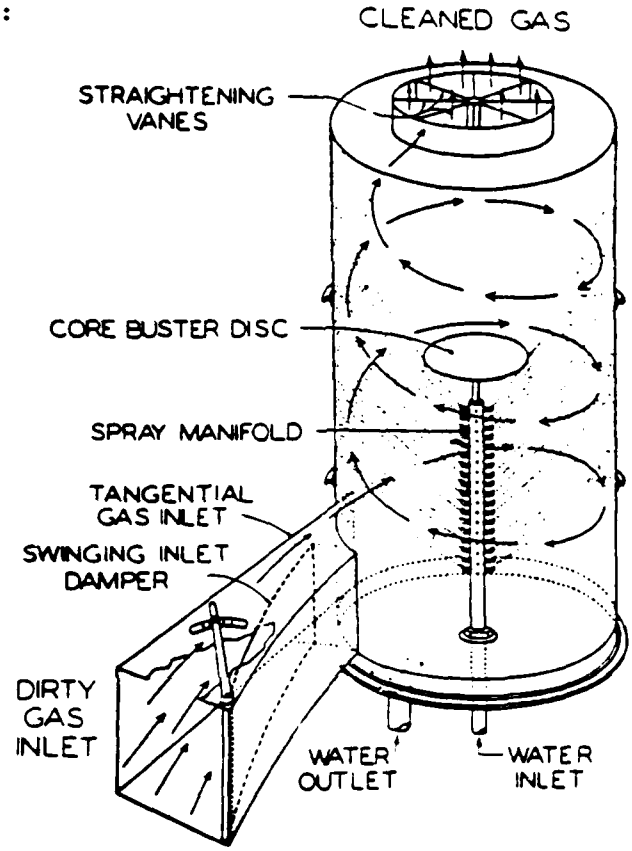
DISADVANTAGES:

1. Cannot handle fibrous, sticky materials.
2. Higher maintenance cost than conventional cyclones.

CONTROL EQUIPMENT: Cyclone Spray Chambers

POLLUTANT TYPE: Particulate

ILLUSTRATION:



Cyclonic Spray Chamber

DESCRIPTION:

There are many other systems capitalizing on centrifugal forces. Many of these systems utilize water to assist in the collection mechanism, and these are discussed in greater detail in Section 2.2. In cyclonic spray chambers, such as in the accompanying figure, the dust-laden gas enters tangentially at the bottom and spirals up through a spray of high-velocity fine water droplets. The dust particles are collected on the fine spray droplets, which are then hurled against the chamber wall by centrifugal action. Other units utilize water to wet and entrap the particles separated from the gas stream by centrifugal action.

TYPICAL OPERATING CHARACTERISTICS:

1. Gas flow ranges from 500 to 25,000 cfm.
2. Gas velocity into cyclone can be up to 200 fps.

3. Pressure loss ranges from 2 to 6 inches of water.
4. Water requirement is 3 to 10 gpm/1,000 cu ft of gas cleaned.

ADVANTAGES:

1. Collection efficiency is 97+% on dust above 1 micron.

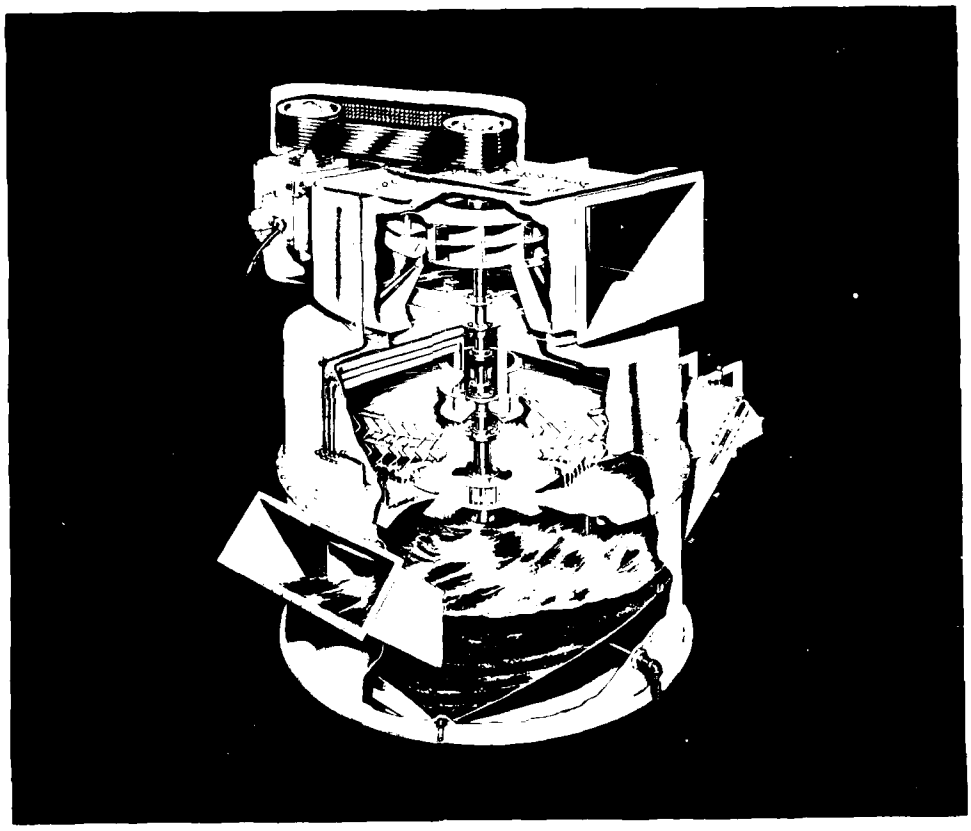
DISADVANTAGES:

1. Pre-cooling of high temperature gases necessary to prevent rapid evaporation of fine droplets.
2. If corrosion-resistant materials are required, costs may be relatively very high.

CONTROL EQUIPMENT: Atomizing-Mechanical Scrubbers

POLLUTANT TYPE: Particulate

ILLUSTRATION:



DESCRIPTION:

In mechanically induced scrubbers, high-velocity sprays are generated at right angles to the direction of gas flow by a partially submerged rotor (see figure). The dirty gas stream passes through the area of the collector that contains the mechanically produced droplets. Scrubbing is achieved by impaction because of both high radial droplet velocity and vertical gas velocity. Liquid atomization occurs at the rotor and the outer wall. Recirculation rates and degree of dispersion vary widely with the different types of rotating elements.

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Pollutant Type:	Control Equipment:
Particulate	Atomizing Mechanical Scrubbers

TYPICAL OPERATING CHARACTERISTICS:

1. Power requirements typically range from 3 to 10 hp/1,000 acfm.
2. Liquid requirements range from 4 to 6 gal/1,000 acfm depending on particle size and desired collection efficiency.

ADVANTAGES:

1. Relatively low liquid and space requirements.
2. High scrubbing efficiency.
3. High dust load capacity.

DISADVANTAGES:

1. Rotor susceptible to erosion from large particles and abrasive dust.
2. High energy scrubbing application usually requires a mist eliminator.

CONTROL EQUIPMENT: Orifice-Type Wet Scrubbers

POLLUTANT TYPE: Particulate

ILLUSTRATION:

DESCRIPTION:

In orifice-type wet scrubbers (sometimes referred to as self-induced spray scrubbers), the gas stream comes into contact with a pool of liquid at the entrance to a constriction (submerged orifice). Liquid is entrained and carried into the restriction where greater liquid-particulate interaction occurs, resulting in high frequency of particulate impaction on the droplets (see figure). Upon leaving the restriction, most of the water droplets (those large enough) are separated by gravity since the gas velocity is reduced from what it was in the restriction. Smaller droplets are subsequently removed by centrifugal force and impingement on baffles located in the upper part of the unit.

TYPICAL OPERATING CHARACTERISTICS:

1. Pressure drop typically ranges from 3 to 10 inches of water.
2. Water requirements range from 1 to 3 gal/1,000 acfm.

ADVANTAGES:

1. Collection efficiencies are often in the range of 90-95% for particulates down to 2 microns.
2. Capable of handling high dust concentrations (no fine clearances to cause plugging).
3. Can handle sticky or linty material.

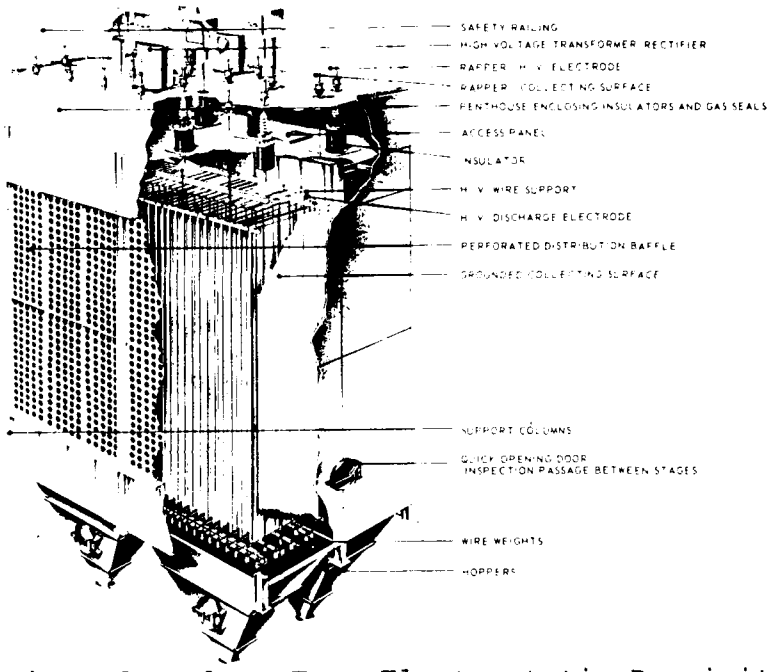
DISADVANTAGES:

1. May have corrosion difficulties.
2. The degree of dispersion of water is not as great as is attained with spray nozzles.

CONTROL EQUIPMENT: Electrostatic Precipitators

POLLUTANT TYPE: Particulate

ILLUSTRATION:



Cutaway View of a Plate-Type Electrostatic Precipitator

DESCRIPTION:

Electrostatic precipitators differ from other gas cleaning equipments in that the force required to separate the particulates from a gas is applied directly to the particles themselves. This force results from the electric charge on the particle in the presence of an electric field. There are different types of precipitators. The plate type is most commonly found in industry. It consists of several plates in parallel with wires of small radius between each pair of plates. High voltage is applied to the wires. Plates are grounded and serve as the other electrode. When gases laden with suspended particulate matter are passed through the high electric field thus created, the ions from the corona emanating from the central electrode region impact with and charge the particles.

The charged particles move towards the plates that act as the collection electrode. When a sufficiently thick layer of dust builds up, it is removed by a sharp impact or rap.

TYPICAL OPERATING CHARACTERISTICS:

1. Gas velocity through the precipitator ranges from 4 to 12 ft/sec.
2. Length to height ratios for the duct vary from 0.5 to 1.5.
3. Secondary voltage from the transformer is in the range of 30 to 100 kv.
4. The optimum spark rate can vary from 0 to 200 sparks/min.
5. Height of the collection electrode ranges from 24 to 40 ft.

ADVANTAGES:

1. Extremely high particulate (coarse and fine) collection efficiencies can be attained (at relatively low energy expenditure).
2. Dry collection and disposal.
3. Low pressure drop (typically less than 0.5 inch of water).
4. Relatively low operating costs.
5. Capable of operation under high pressure (to 150 psi) or vacuum conditions.
6. Capable of operation at high temperature (to 1,300^{o}F).
7. Relatively large gas flow rates can be effectively handled.

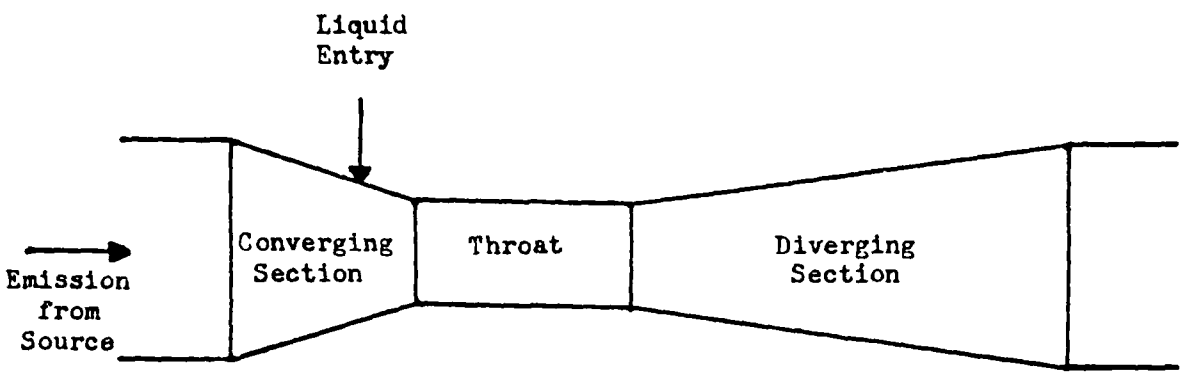
DISADVANTAGES:

1. High capital cost.
2. Very sensitive to fluctuations in gas stream conditions (in particular flows, temperatures, and particulate loadings).
3. Certain particulates are difficult to collect due to extremely high or low resistivity.
4. Relatively large space required for installation.
5. Explosion hazard when treating combustible gases and/or collecting combustible particulates.
6. Ozone is produced by the negatively charged discharge electrode during gas ionization.
7. Relatively sophisticated maintenance personnel required.

CONTROL EQUIPMENT: Venturi Scrubbers

POLLUTANT TYPE: Particulate

ILLUSTRATION:



Venturi Scrubber

DESCRIPTION:

A venturi scrubber is a type of gas-atomized spray scrubber that has a converging section, a throat, and a diverging section in series as shown in the figure. It may have a circular or a rectangular cross section. The gas loaded with the particulates enters the converging section. Usually liquid enters the venturi upstream of the throat through nozzles. It is shattered into very small droplets by high velocity gas. These drops are then accelerated until they reach the gas velocity. Because of the difference between the velocity of gas and that of liquid drops, the gas flows past these drops. The particles suspended in the gas and moving at the gas velocity strike the liquid drops due to various mechanisms, most importantly inertial impaction. They become attached to the drops and are removed from the gas stream. Venturi scrubbers thus provide a small droplet diameter and high relative velocity, the conditions required to achieve high collection efficency for the particulates by impaction.

TYPICAL OPERATING CHARACTERISTICS:

1. Gas velocities through venturi typically range from 12,000 to 24,000 ft/min.
2. Venturi must be followed by a separating section for the elimination of entrained droplets.
3. Water requirements most frequently range from 6 to 10 gal/1,000 acfm of gas treated.
4. High collection efficiencies are achievable with pressure drops ranging from 6 to 70+ inches of water.
5. Collection efficiency is directly related to pressure drop.
6. Variable-throat venturi scrubbers have been introduced to maintain pressure drop with varying gas flows.

ADVANTAGES:

1. Simple construction, hence low initial capital investment.
2. High collection efficiency for even submicron particulates.
3. Can be used for the collection of sticky particulates.

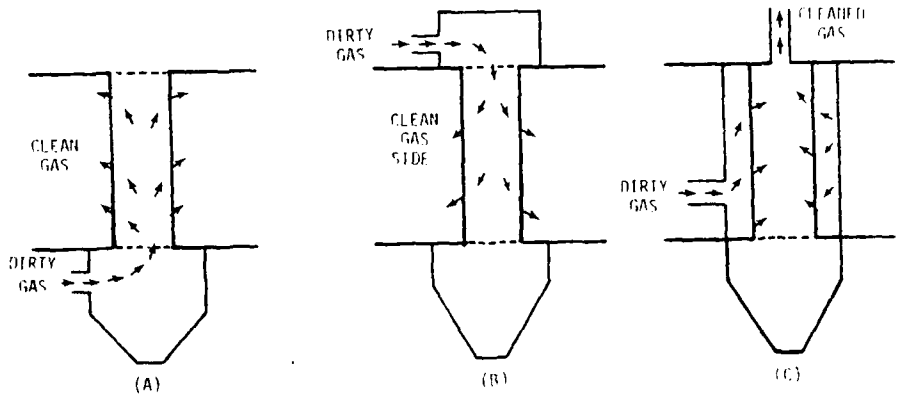
DISADVANTAGES:

1. Relatively high pressure drops, thus high operating cost.
2. Limited applicability for mass-transfer problems.
3. Requires water treatment facilities.

CONTROL EQUIPMENT: Fabric Filters

POLLUTANT TYPE: Particulate

ILLUSTRATION:



Filtering Systems

(A) Bottom Feed (B) Top Feed (C) Exterior Filtration

DESCRIPTION:

In its simplest form, the industrial fabric filter consists of a woven or felted fabric through which dust-laden gases are forced. A combination of factors results in the collection of particles on the fabric fibers. When woven fabrics are used, a dust cake eventually forms; this, in turn, acts predominantly as a sieving mechanism. When felted fabrics are used, this dust cake is minimal or nonexistent. Instead, the primary filtering mechanisms are a combination of inertial forces, impingement, etc., as related to individual particle collection on single fibers. As particles are collected, the pressure drop across the fabric filtering media increases. Due in part to fan limitations, the filter must be cleaned at predetermined intervals. Dust is removed from the fabric by gravity and/or mechanical means. The fabric filters or bags are usually tubular or flat. The structure in which the bags hand is frequently referred to as a baghouse. The number of bags in a baghouse may vary from a couple to several thousand. Quite often when great numbers of bags are involved the baghouse is compartmentalized so that one compartment may be cleaned while others are still in service.

TYPICAL OPERATING CHARACTERISTICS:

1. Surface area of fabric ranges from as low as 10 ft^2 to as high as $1 \times 10^6 ft^2$.
2. Pressure drop in the collector varies from 2 to 4 inches of water. It is 3 to 6 inches of water for ducting.
3. Conveying velocities in the duct work range from 2,000 to 3,000 ft/min.
4. Air-to-cloth ratio is most frequently 10 to 15 cfm/ft^2.

ADVANTAGES:

1. Extremely high collection efficiency on both coarse and fine (submicron) particulates.
2. Relatively insensitive to gas stream fluctuation. Efficiency and pressure drop are relatively unaffected by large changes in inlet dust loadings for continuously cleaned filters.
3. Filter outlet air may be recirculated within the plant in many cases (for energy conservation).
4. Collected material is recovered dry for subsequent processing or disposal.
5. There is no hazard of high voltage, simplifying maintenance and repair and permitting collection of flammable dusts.
6. Relatively simple operation.

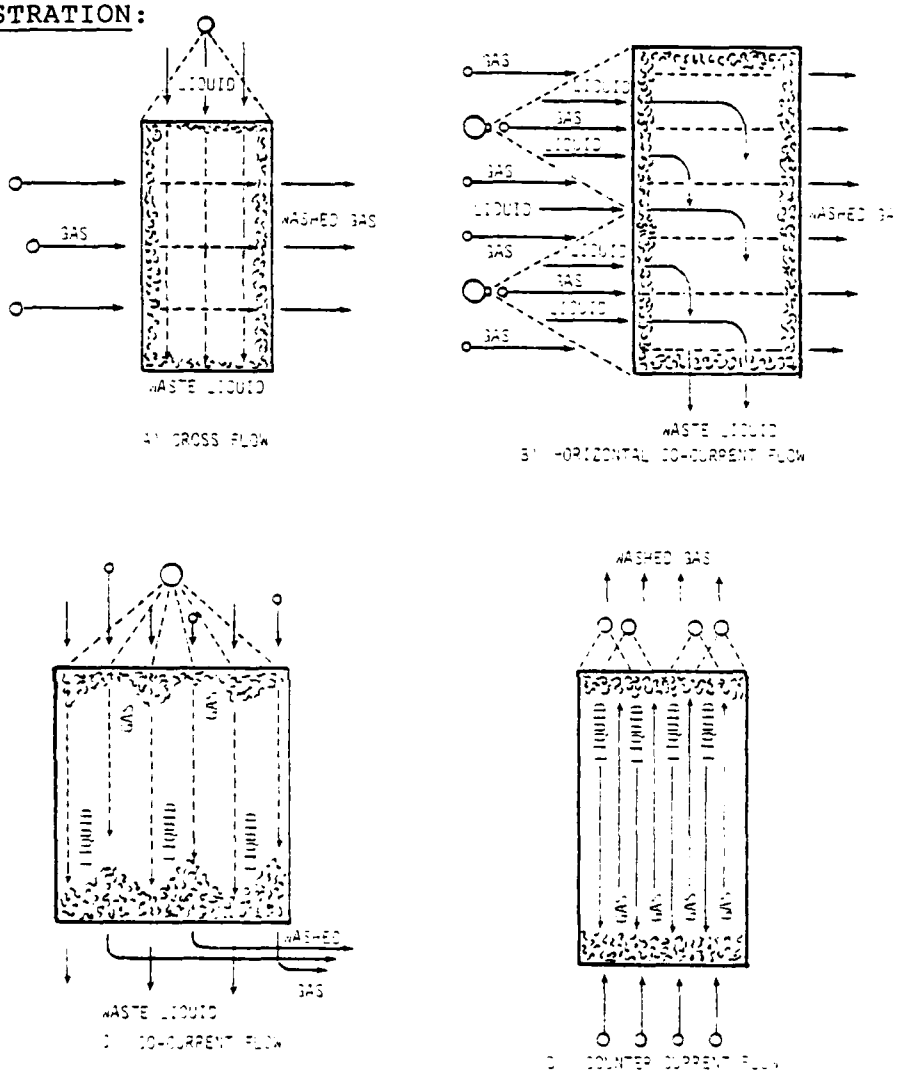
DISADVANTAGES:

1. Temperature much in excess of 500^oF requires special refractory mineral or metallic fabrics that are still in the developmental stage and can be very expensive.
2. Certain dusts may require fabric treatments to reduce dust seeping or, in other cases, assist in the removal of the collected dust.
3. Concentrations of some dusts in the collector ($50g/m^3$) may be a fire or explosion hazard if spark or flame is admitted by accident. Fabrics can burn if rapidly oxidizable dust is being collected.
4. Relatively high maintenance requirements (bag replacement, etc.).
5. Hygroscopic materials, condensation of moisture, or tarry adhesive components may cause crusty caking or plugging of the fabric or require special additives.

CONTROL EQUIPMENT: Packed-Bed Scrubbers

POLLUTANT TYPE: Gaseous

ILLUSTRATION:



Packed-bed Scrubbers

DESCRIPTION:

Packed-bed scrubbers, used for the continuous contact of liquid and gas, are usually vertical columns that have been filled with packing or devices of large surface area. The liquid is distributed over, and trickles down through, the packed bed,

thus exposing a large surface area to contact the gas. The countercurrent packed column (D) is the most common type of unit encountered in gaseous pollutant control. The gas stream (containing the pollutant) moves upwards through the packed bed against an absorbing or reacting liquor (solvent-scrubbing solution) that is injected at the top of the packing. Since the solute concentration in the gas stream decreases as it rises through the column, there is constantly fresher solvent available for contact.

Apart from the countercurrent flow, there are three other ways in which the bed may be operated. These are shown in the illustration (A & B & C). However, the countercurrent scrubbers have been found most efficient.

TYPICAL OPERATING CHARACTERISTICS:

1. The operating gas rates seldom exceed 60-75% of the rate that would cause flooding.
2. Column dimensions specified are usually in readily available "off-the-shelf" sizes (i.e., diameters to the nearest half foot and heights to the nearest foot).
3. Bed depths may vary from 6 inches to several feet depending upon the type of packing and the application.
4. Typical pressure drop through a packed bed is 0.5 inch of water per foot of packing.
5. Gas is generally redistributed every 6 to 10 feet of packed height to minimize channeling.

ADVANTAGES:

1. Relatively low pressure drop.
2. Standardization in fiberglass-reinforced plastic (FRP) construction permits operation in highly corrosive atmospheres.
3. Capable of achieving relatively high mass transfer efficiencies.
4. Increasing the height and/or type of packing can improve mass transfer without purchasing a new piece of equipment.
5. Relatively low capital cost.
6. Relatively small space requirements.

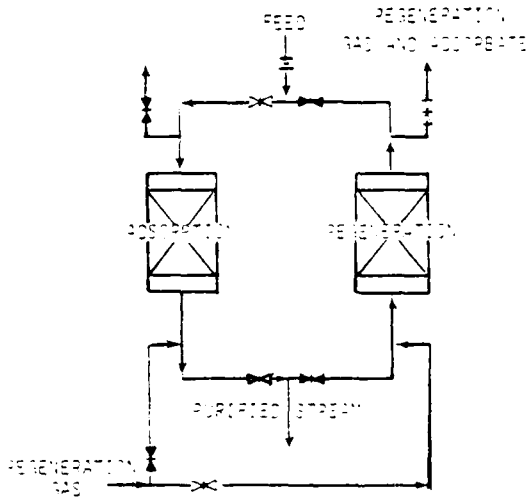
DISADVANTAGES:

1. May create water disposal problem.
2. Particulates deposition may cause plugging of the bed.
3. When FRP construction used, sensitive to temperature.
4. Relatively high maintenance costs.

CONTROL EQUIPMENT: Adsorption System

POLLUTANT TYPE: Gaseous

ILLUSTRATION:



Two-bed Adsorption System

DESCRIPTION:

Two cylindrical vessels filled with an adsorbant in a vertical arrangement are customary for use in a continuous adsorption operation. While one bed is adsorbing vapors from the gas stream, the second bed is being stripped or adsorbate and regenerated.

TYPICAL OPERATING CHARACTERISTICS:

Activated Carbon

1. In general, removal of gaseous vapors by physical adsorption is practical for vapors with molecular weights greater than 45 and boiling points greater than 32°F (an exception to the molecular weight rule is methanol).
2. For average bed conditions, 100 lb carbon can efficiently treat 200 acfm of solvent laden air per hour.
3. Usually regenerate activated carbon with saturated steam at low pressure (5 psig).
4. Steam-to-solvent ratio usually varies between 1 and 4 lb steam per lb solvent desorbed from carbon (when volume % solvent concentration is greater than 0.2). If solvent concentration is much less than 0.2%, the steam requirements can be as high as 30 lb steam per lb solvent.
5. Recommend countercurrent operation (adsorb downward, regenerate upward).

6. For best results, keep operating temperature less than 100°F.
7. Experience has shown that velocities of 60 to 70 fpm at the carbon bed are best.

ADSORPTION ON ACTIVITED CARBON: TYPICAL PARAMETERS

Parameter	Odor Removal	Pollution Control	Solvent Recovery
Concentrations	1 PPMV	1 PPMV-1% vol.	0.1-3% vol.
Bed Depth	1/2 inch	9 inches	24 inches
Regeneration	Reactivate or discard	Regenerate in place	Steam
Adsorption Cycle	1 year	1-8 hours	30-60 min.
Pressure drop	0.25 in.H_2O	10-20 in. H_2O	20-30 in. H_2O
Air Flow	2,000 acfm/ filter	1,000-100,000 acfm/bed	1,000-40,000 acfm/bed

ADVANTAGES:

1. Product recovery may be possible.
2. Excellent control and response to process changes.
3. No chemical disposal problem when pollutant (product) recovered and returned to process.
4. Capability of systems for fully automatic, unattended operation.
5. Capability to remove gaseous or vapor contaminants from process streams to extremely low levels.

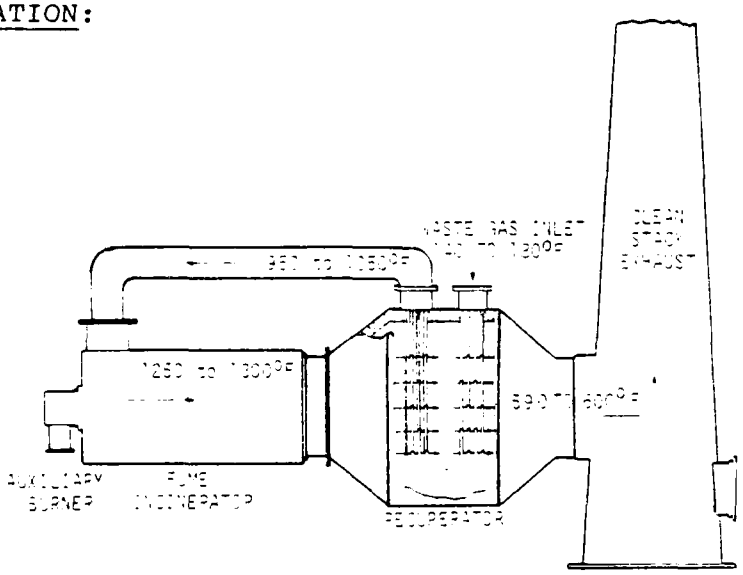
DISADVANTAGES:

1. Product recovery may require an exotic, expensive distillation (or extraction) scheme.
2. Adsorbent progressively deteriorates in capacity as the number of cycles increases.
3. Adsorbent regeneration requires a steam or vacuum source.
4. Relatively high capital cost.
5. Prefiltering of gas stream may be required to remove any particulate capable of plugging the adsorbent bed.
6. Cooling of the gas stream may be required to get to the usual range of operation (less than 100°F).
7. Relatively high steam requirements to desorb high molecular weight hydrocarbons.

CONTROL EQUIPMENT: Thermal Combustion Reactor

POLLUTANT TYPE: Gaseous

ILLUSTRATION:



Thermal Combustion with Energy (Heat) Recovery

DESCRIPTION:

Thermal incinerators or afterburners can be used over a fairly wide range of organic vapor concentrations. Reactions are conducted at elevated temperatures in order to insure high chemical reaction rates for the organics. To achieve this temperature, it is necessary to preheat the feed stream with auxiliary energy. The burner may utilize air in the process waste stream as the combustion air for the auxiliary fuel, or may use a separate source of outside air for this purpose.

TYPICAL OPERATING CHARACTERISTICS:

1. Unit requires operating temperatures in the 1,200° to 1,500°F range for combustion of most pollutants.
2. Residence times between 0.1 and 1 second are typical.
3. Equipment length-to-diameter ratio of 2.0 to 3.0 is usually specified.
4. The average gas velocity at the reactor outlet can range from 10 to 50 ft/sec.

ADVANTAGES:

1. Simplicity of construction.
2. Removal of submicron particles.

3. Capability for virtual complete destruction of organic contaminants.
4. Small space requirements.
5. Low maintenance costs.
6. Capability of steam generation or heat recovery in other forms.

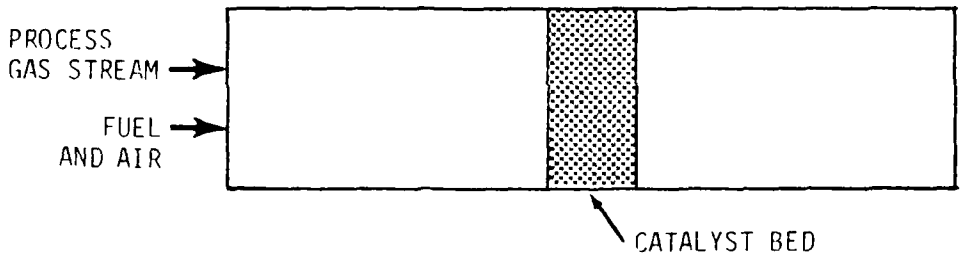
DISADVANTAGES:

1. High operating costs.
2. Potential for flashback and subsequent explosion hazard.

CONTROL EQUIPMENT: Catalytic Combustion Reactor

POLLUTANT TYPE: Gaseous

ILLUSTRATION:



Catalytic Combustion Reactor

DESCRIPTION:

Catalytic reactors are an alternative to thermal incinerators. For simple reactions, the effect of the presence of a catalyst is to:

1. increase the rate of reaction.
2. permit the reaction to occur at a lower temperature.
3. permit the reaction to occur at a more favorable pressure.
4. reduce the reactor volume.
5. increase the yield of a reactant(s) relative to the other(s).

Metals in the platinum family are recognized for their ability to promote combustion at relatively low temperatures. Other catalysts include various oxides of copper, chromium, vanadium, nickel, and cobalt.

TYPICAL OPERATING CHARACTERISTICS:

1. The gas stream is usually delivered to the reactor by a fan at a velocity in the 10 to 30 ft/sec range.
2. Reactor temperature is usually in the 650°to 800°F range.
3. A reactor length-to-diameter ratio less than 0.5 is usually specified.
4. The smaller the size of the catalyst used, the more efficiently it operates; however, the pressure drop through the reactor will increase.
5. Typically 0.5 to 2 cu/ft catalyst/1,000 scf waste stream for 85-95% conversion of hydrocarbons in the waste stream.

ADVANTAGES:

1. Low fuel requirements.
2. Little or no insulation requirements.
3. Reduced flashback problems and explosion hazards.

DISADVANTAGES:

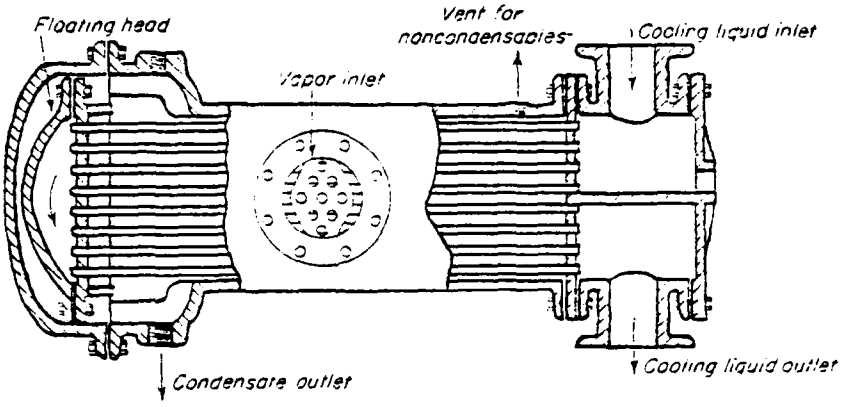
1. High initial cost.
2. Catalyst poisoning.
3. Particulates must first be removed.
4. Catalyst regeneration problems.

CONTROL EQUIPMENT: Condensers

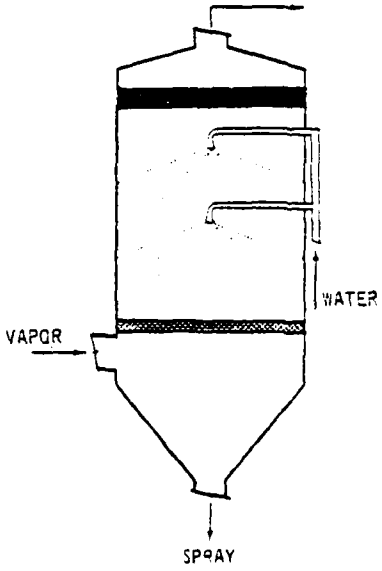
POLLUTANT TYPE: Gaseous

ILLUSTRATION:

(a)



(b)



(a) Surface Condenser (shell and tube type)

(b) Contact Condenser (spray type)

DESCRIPTION:

Condensers can be used to collect the organic emissions to the atmosphere by lowering the temperature of the gaseous stream. There are two basic types of condensers (heat exchangers) used for control: contact and surface. In contact condensers, the gaseous stream is brought into direct

contact with a cooling medium so that the vapors condense and mix with the coolant. However, the most commonly used unit is the surface condenser. Here, the vapor and the cooling medium are separated by the exchanger wall. There are various types of surface condensers including the shell and tube, fin-fan, finned hairpin, finned tube section and tubular.

TYPICAL OPERATING CHARACTERISTICS:

1. The overall heat transfer coefficient is usually low on the order of 10 to 25 Btu/hr-ft^2-$^\circ$F for organic vapor/noncondensable gas systems.

ADVANTAGES:

1. Pure product recovery (in the case of indirect-contact condensers).
2. Water used as the coolant in an indirect-contact condenser (i.e., shell and tube heat exchanger) does not contact the contaminated gas stream and can be reused after cooling.

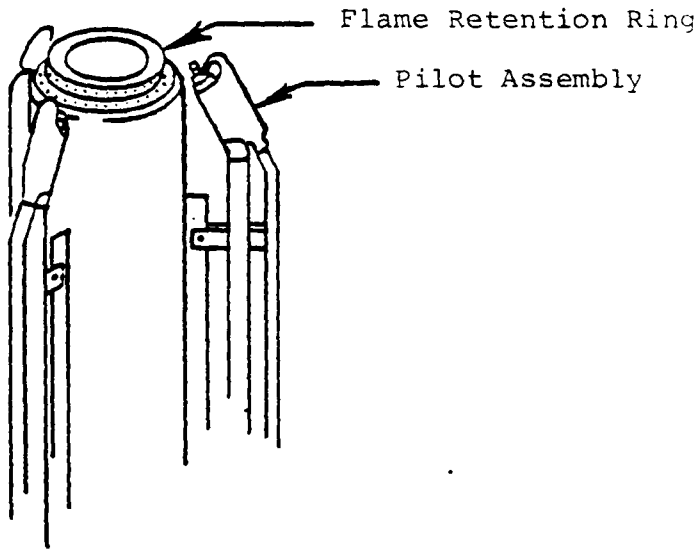
DISADVANTAGES:

1. Relatively low removal efficiency for gaseous contaminants (at concentraions typical of pollution control applications).
2. Coolant requirements may be extremely expensive.

CONTROL EQUIPMENT: Flares

POLLUTANT TYPE: Gaseous

ILLUSTRATION:



Utility Field Flare Tip

DESCRIPTION:

The flare system is used primarily as a safe method for disposing or escess waste gases. All process plants that handle hydrocarbons, hydrogen, ammonia, hydrogen cyanide, or other toxic or dangerous gases are subject to emergency conditions that occasionally require immediate release of large volumes of such gases for protection of plant and personnel. Flares are used for the purpose. In operation, the gas containing the organics is continuously fed to and discharged from a stack, with the combustion occurring near the top of the stack and characterized by a flame at the end of the stack.

TYPICAL OPERATING CHARACTERISTICS:

1. Flares generally operate in the 2,000° to 3,000°F range.
2. Used where concentration of pollutant is high. When the concentration is above the LEL but below the UEL.
3. Operating stack velocities are in excess of the flame propagation rate, and frequently exceed 200 ft/sec.

4. In general, a waste gas, with a heating value greater than 200 Btu/cu ft can be flared successfully. It is usually not feasible to flare a gas with a heating value below 100 Btu/cu ft.
5. It is a good practice to size flare stack on the basis of an exit velocity equal to 20% of the sonic velocity.

ADVANTAGES:

1. Economical and safe disposal of large volumes of excess waste gases.
2. Low capital and operating costs.

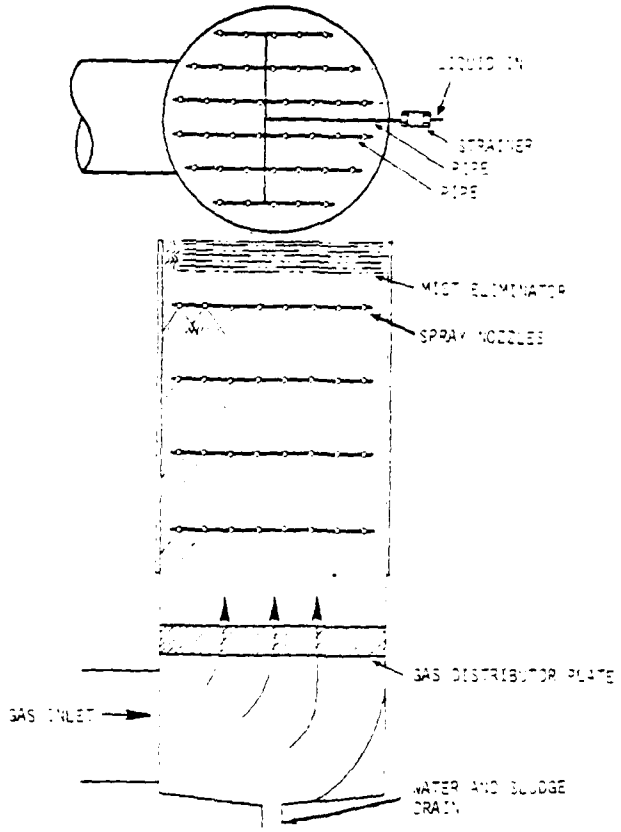
DISADVANTAGES:

1. Can be used economically only when the concentration at the combustibles is within the flammable range.
2. Inability to show compliance with air pollution codes by sampling the gas stream after combustion.
3. Diameter of flair stack depends upon the expected emergency gas flow rate.
4. Stack gas velocity usually limited to about 5,000 ft/sec to prevent flame extinction by blowout.

CONTROL EQUIPMENT: Spray Chambers

POLLUTANT TYPE: Gaseous and Particulate

ILLUSTRATION:



Gravity Spray Tower

DESCRIPTION:

The simplest type of scrubber is a chamber in which spray nozzles are placed (see figure). The gas stream velocity decreases as it enters the chamber, and the wetted particles settle and are collected at the bottom of the chamber. Its efficiency as a dust collector is low except for coarse dust. Efficiency can be improved by baffle plates upon which particles can be impinged.

TYPICAL OPERATING CHARACTERISTICS:

1. Low pressure drop, typically 1 to 2 inches of water exclusive of mist eliminator and distribution plate.
2. Liquid requirements ranging from 3 to 20 gal/1,000 acfm of gas treated.

3. Can handle large gas volumes; 800 to 2,500 lb/hr-ft^2 is typical.
4. Often used as precoolers to reduce gas stream temperatures.
5. Efficiencies often exceed 70% for particles larger than 10 μ.
6. Improved efficiencies are possible by adding high-pressure sprays ranging from 100 to 400 psig.

ADVANTAGES:

1. Can handle relatively high dust concentrations without the fear of chokage.
2. Spray generators need not have fine jets. This allows the recirculation of dirty water resulting in water saving and perhaps a simplification of effluent treatment.

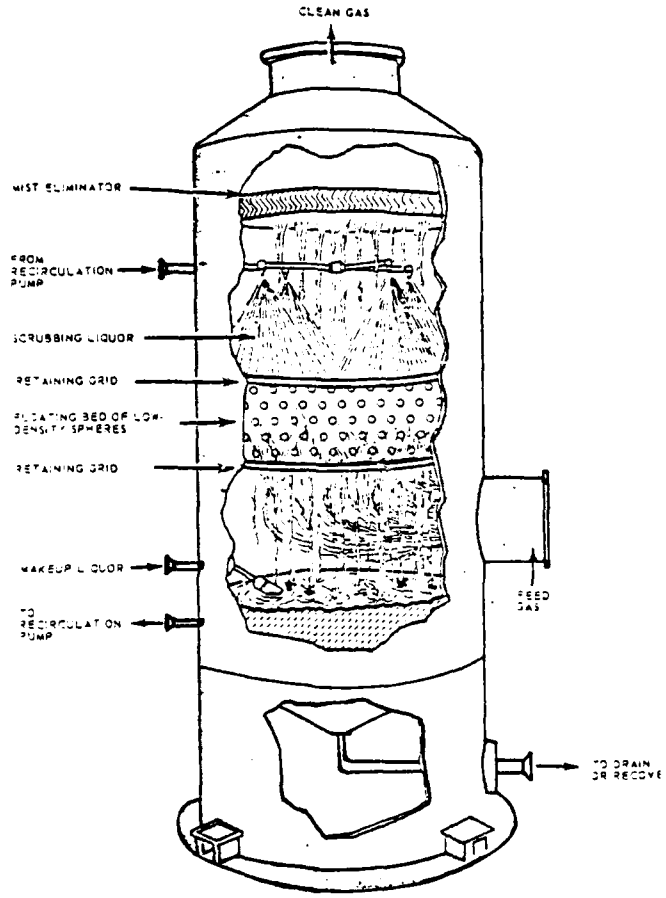
DISADVANTAGES:

1. Very low scrubbing efficiencies for particles in 0 to 5 micron range.

CONTROL EQUIPMENT: Moving Bed Scrubbers

POLLUTANT TYPE: Gaseous and Particulate

ILLUSTRATION:



Moving Bed Scrubber

DESCRIPTION:

Moving-bed (fluid-bed) scrubbers incorporate a zone of movable packing where gas and liquid can intimately mix. The system shown in the figure uses packing consisting of low-density polyethylene or polypropylene spheres about 1.5 inch in diameter; these are kept in continuous motion between the upper and lower retaining grids. Such action keeps the spheres continually cleaned and considerably reduces any tendency for the bed to plug.

TYPICAL OPERATING CHARACTERISTICS:

1. Pressure drops typically range from 3 to 5 inches of water per stage.
2. Collection efficiencies are often in excess of 99% for particles down to 2 μm.
3. Collection of particles may be enhanced by using several moving bed stages in series.

ADVANTAGES:

1. Very efficient collection of fine particulates with simultaneous scrubbing of gases.

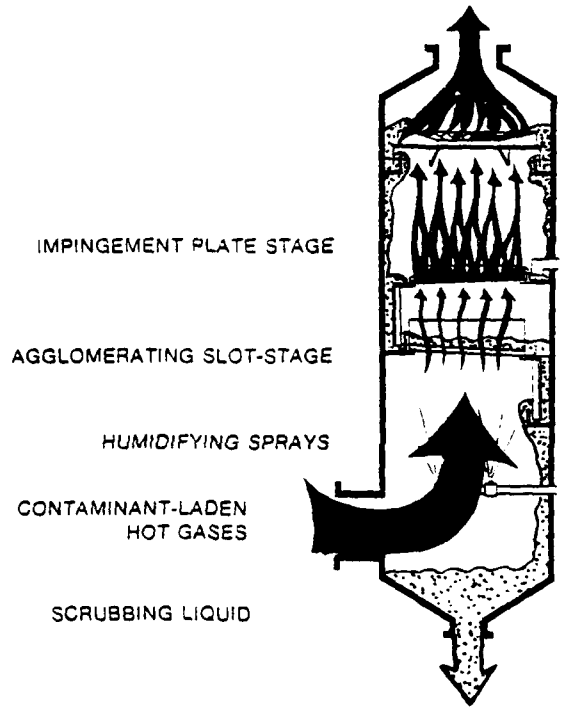
DISADVANTAGES:

1. There is some tendency for the bed to plug.
2. Relatively high pressure drop.

CONTROL EQUIPMENT: Impingement Plate Scrubbers

POLLUTANT TYPE: Gaseous and Particulate

ILLUSTRATION:



Impingement Plate Scrubber

DESCRIPTION:

Impingement plate scrubbers (see figure) utilize perforated plates with an impingement baffle over each perforation. The intention here is to expand the surface area of the liquid through use of the gas stream's kinetic energy. Gas flowing upward is divided into thousands of jets by the orifices. Each jet aspirates liquid from the blanket and creates a wetted surface on the baffle, located at the point of maximum jet velocity. The directed impingement on a wetted target dynamically precipitates particles and entraps them in the scrubbing liquid. On impingement, each jet forms minute gas bubbles which rise through and create turbulence in the liquid blanket, thus providing extremely close gas-liquid contact for maximum clearing. Continuous violent agitation of the blanket by the bubbles prevents settling of entrapped particles and flushes them away in the scrubbing liquid.

TYPICAL OPERATING CHARACTERISTICS:

1. Collection efficiencies for a single plate may range from 90-99% for 1 to 5 micron particulates.
2. Pressure drops from 1 to 3 inches of water per plate are typical.
3. Water requirements usually range from 3 to 5 gal/1,000 acfm of gas treated.
4. Gas velocities of 5 to 10 ft/sec through the tower are common.

ADVANTAGES:

1. Ability to collect particulates as well as gases.
2. Increasing the number of plates can improve mass transfer without purchasing a new piece of equipment.
3. Capable of achieving relatively high mass transfer coefficients.
4. Lighter in weight than packed towers.
5. Less susceptible to plugging than packed towers.

DISADVANTAGES:

1. Relatively high maintenance cost.
2. May create water disposal problem.
3. Product collected wet.
4. Particulates deposition may cause plugging of the plates.

S.3 DEVELOPMENT OF CONTROL METHODOLOGY

Control Requirements

Before applying a specific air pollution control methodology, it is neccesary to develop the control requirements. The degree of control required depends upon the uncontrolled emission rate and the desired emission rate according to the formula:

$$\text{Desired efficiency } (\%) = \left(1 - \frac{\text{desired emission}}{\text{actual emission}}\right) \times 100$$

The desired emission rate may consider possible future emission limitations as well as present ones.

The degree of control effected by the existing process equipment indicates the additional control necessary for meeting the projected emission levels. A specified degree of control for each pollutant for which there is a limitation will enhance the selection of the control system.

Process Changes

Process changes are sometimes viable methods of complying with emission limitations. A review of process changes on similar processes will reveal whether the concept is feasible. If an engineering analysis would disclose the methodology and magnitude of emission reduction through process changes, then the time and cost of such a study must be considered. A detailed description of how the study results will be evaluated and utilized is necessary before undertaking the study, particularly if the source is operating in violation of an emission regulation and a compliance schedule is in effect.

Some viable alternatives are raw material changes, fuel changes, process equipment changes, and operations changes. More detailed information on process changes as control methods can be found in Chapter 3.

Current Control Technology

In order to assess the control technology, a summary of existing methodology should be made and should include all the commonly used types of control equipment and their application, ranges of efficiency, utilities consumption. average initial cost, and operations/maintenance cost.

An evaluation of the development of new technology may be necessary if the existing technology is not sufficient to meet the control requirements.

Gathering of Emission Streams

Where several similar processes are vented, it is possible that the gathering of flow streams can be an advantage in providing emission control for all processes.

Control regulations must be closely examined since the requirements may change significantly if the flow streams from several points are gathered and vented at a single point.

If the gathering of streams appears feasible, the methodology to be used must be determined. An engineering study may be necessary to determine the methodology.

Control Technology from Similar Processes

The feasibility of utilizing up-to-date control technology from similar processes can be determined in cases where current technology does not fit the control requirements. Technology transfer is the method used to obtain information necessary for determining the feasibility of technological application.

Control concepts under development can be investigated to determine their applicability. The up-to-date control concepts should be summarized and feasibility studies for the utilization of technology transfer should be described.

Cost vs. Performance

A summary of the cost of all applicable control techniques that will meet the control requirements should be followed by detailed economic information, including iinitial costs and operating costs. The capital costs of the control system of any proposed process change begin with the costs of any required major process equipment or control apparatus and must also include such auxiliaries as pumps, fans, water pollution control facility and energy economizers. Included in installation costs are foundations, erection, piping, electricity, and other services.

Another cost that may be significant is that associated with modifying the production or manufacturing process to mate or interface the collector with that process. These initial costs are on-time costs.

The operating costs, which must also be considered, are associated with energy requirements, maintenance, by-product disposal, reagent, effluent purification, and labor. These expenses occur on a continuing basis. Also included on an occasional basis is the cost for operator training. Another consideration in evaluating the total cost of any control system is the time value of money, or the difference between the income that could be generated if the capital used for pollution control were invested in high quality bonds or even deposited at interest and any profit that can be attributed to the pollution control effort. Other contributions to the operating costs are overhead, taxes, insurance, deprecitation, and any return on investment (usually a negative value). All of these items must be considered with the initial and installation costs to determine the total annualized cost for each control process device being evaluated. This information is used in a cost and benefit analysis as an aid to control approach selection. Details of equipment and operating costs for control equipment can be found in Chapter 6.

If the study is complete, the conclusions may be evaluated and recommendations made for any additional efforts.

Selection of Best Control Approach Technology

After a thorough search of all alternatives has been performed, the selection of the best control approach must be made. The method used to select the proposed control approach should be described, and the rationale behind the selection should be given. Potential problems should be listed and any additional necessary research and development should be described.

1. AIR POLLUTION EMISSIONS FROM SPECIFIC OPERATIONS AT ARMY INSTALLATIONS AND SELECTION OF CONTROL EQUIPMENT — Page

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CHAPTER 1

AIR POLLUTION EMISSIONS FROM SPECIFIC OPERATIONS AT ARMY INSTALLATIONS AND SELECTION OF CONTROL EQUIPMENT

This chapter gives a detailed description of air pollution emissions and their sources at Army installations. It also discusses the alternatives available to control these emissions. Table S-1 is a one-page review of what is contained in the chapter.

There are four major areas of emission sources at Army installations:

1. steam generation
2. solid waste incineration
3. ammunitions manufacture
4. evaporative loss sources.

Each of the above is treated separately here. Different processes causing air pollution, within a given class, are identified and their operating characteristics are described. The resulting emissions are characterized and the control methods are discussed.

1.1 STEAM GENERATION--BOILERS AND HEATING PLANTS[1,2]

1.1.1 Process Characterization

The vast majority of combustion equipment is used to heat or vaporize water, or both. For convenience, industrial water heaters are considered together with boilers inasmuch as identical equipment is frequently used for both purposes. A sketch of a typical steam boiler is shown in Figure 1-1. The radiant section of the boiler is lined with boiler tubes on the walls, floor, and roof of the furnace enclosure. The boiler feed water is converted to saturated steam within these tubes through the radiant transfer of heat from the hot combustion gases within the furnace. Additional heat transfer tubes may be required on larger boilers to superheat the saturated steam and are usually included directly following the radiant section of the boiler. Larger boilers have an air preheater to transfer heat from the boiler exhaust to incoming combustion air, a heat-exchanger to preheat the feedwater, and an economizer. The types of boilers considered in this section include:

1. coal-fired stoker boilers
2. pulverized-coal-fired boilers
3. oil-fired boilers.

Once the fuel being used is ignited in the furnace, the process of combustion begins. The combustion process involves a fuel containing carbon (C) and hydrogen (H_2) burning in the presence of oxygen (O_2) to produce carbon dioxide (CO_2), water (H_2O), and heat. The rest of the process involves heat transfer from the combustion gases to the feed water. The gases are circulated in order to utilize as much heat as possible before being discharged to the atmosphere. Air pollution control devices are usually placed between the boiler and the induced-draft fan.

1.1.2 Coal-fired Boilers

Coal is one of the major fuels used in boilers to generate steam. Depending upon the mode of feeding, coal-fired boilers can be classified as (1) coal-fired stoker boilers or (2) pulverized-coal-fired boilers. Stokers of various types are designed to automatically feed coal uniformly onto a grate within the combustion zone of the furnace and to discharge the ash residue to a collection device. In pulverized coal boilers, the coal is finely ground and combined with a hot air stream. It is pneumatically conveyed to the boiler nozzles and injected into the boiler zone.

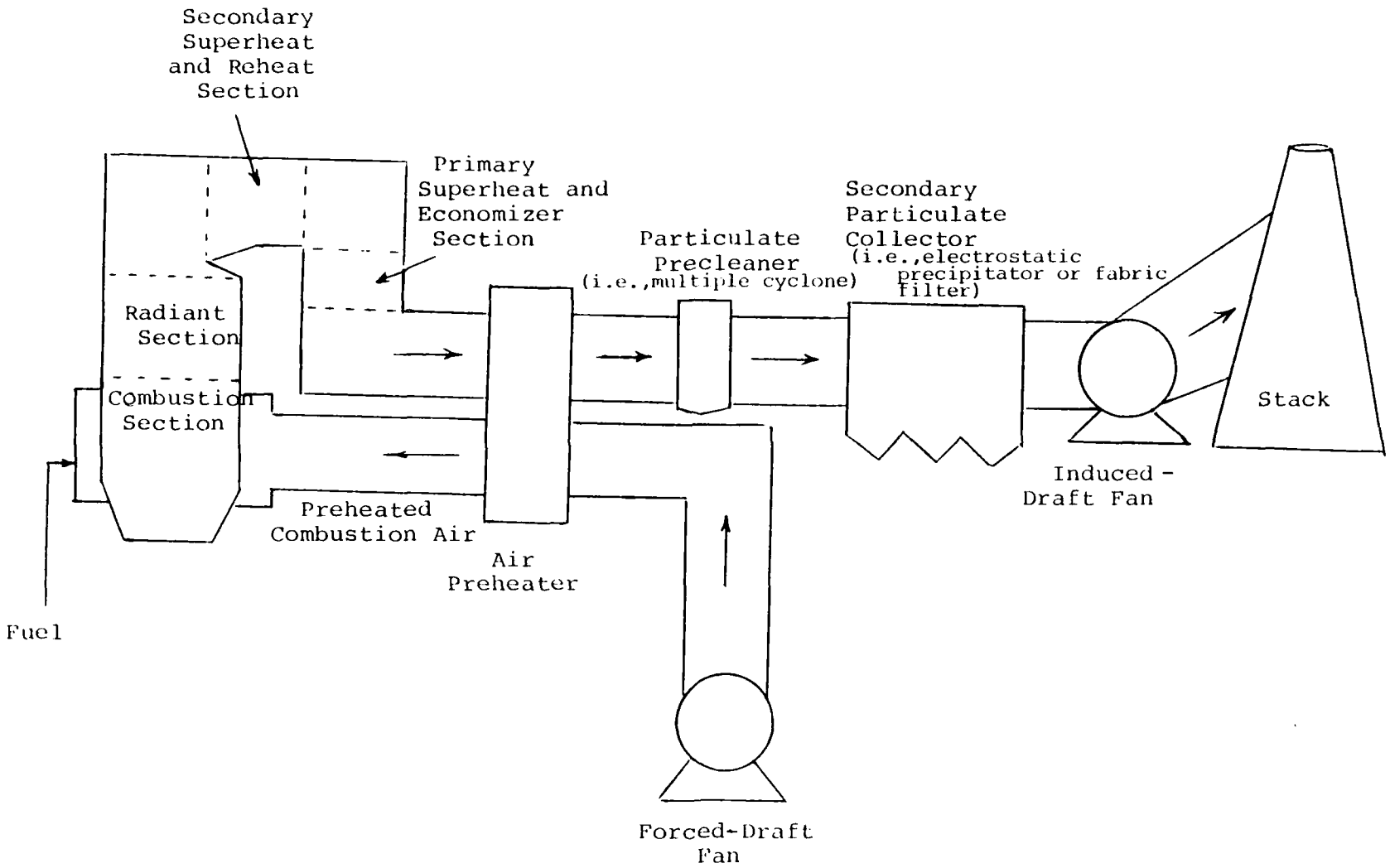

FIGURE 1-1

FLOW DIAGRAM OF BOILER WITH PARTICULATE CONTROL

1.1.2.1 Estimation of Pollutant Emissions

The emission factors[3] listed give representative pollutant emission rates for boilers without air pollution controls burning anthracite coal (Table 1-1), bituminous coal, (Table 1-2), and lignite (Table 1-3) based on boiler size and firing method. The accuracy of emission estimates based on emission factors depends on the boiler's age, condition, and operation. To use the emission factors, the ash, sulfur, and heat content of the coal used, as well as the boiler's maximum heat input, value must be known. The uncontrolled emission rate can be calculated by first determining the coal usage rate in tons per hour. This may be found by dividing the maximum heat input per hour (in Btu/hr) by the heat content of the coal (in Btu/ton). The uncontrolled emission rate can then be estimated by multiplying the usage rate by the proper emission factor from the tables. (Examples are given in the sections that follow.)

A. Particulate Emissions - Particulate emissions depend on the ash content of the coal used, the stokering system, the size of the boiler, and the particulate control device used. (In general, the size of the boiler is only important for bituminous coal. Table 1-2 shows three size range classes: greater than 100 x 10^6 Btu/hr, 10 to 100 x 10^6 Btu/hr, and less than 10 x 10^6 Btu/hr.) The proper emission factor is found by locating the size range and stoker system for the boiler on the table and looking for the corresponding emission factor for particulates. For example, the emission factor for a 24 x 10^6 Btu/hr boiler with a spreader stoker burning bituminous coal (from Table 1-2) is 13A lb/ton. The A is the weight percent ash in the coal. If the coal used had a 7 percent ash content, the emission factor would be 13 x 7 or 91 lb/ton. The uncontrolled particulate emissions for the boiler would be calculated by multiplying the fuel usage rate by the emission factor. The uncontrolled particulate emissions for the above-mentioned boiler would be calculated as follows:

Given that the bituminous coal used has 7 percent ash content and 13,000 Btu/lb,

$$24 \times 10^6 \frac{\text{Btu}}{\text{hr}} \times \frac{1 \text{ lb}}{13{,}000 \text{ Btu}} \times \frac{1 \text{ ton}}{2{,}000 \text{ lb}} = 0.92 \frac{\text{ton}}{\text{hr}}$$

(Fuel Usage Rate)

TABLE 1-1

EMISSIONS FROM ANTHRACITE COAL COMBUSTION WITHOUT CONTROL EQUIPMENT[3]

Type of Furnace	Particulate[a]		Sulfur Dioxide[b]		Sulfur Trioxide[b]		Hydrocarbons[c]		Carbon Monoxide[d]		Nitrogen Oxides[e]	
	lb/ton	kg/MT	lb/ton	kg/MT	lb/ton	kg/MT	lb/ton	kg/MT	lb/ton	kg/MT	lb/ton	kg/MT
Pulverized (dry bottom), no fly-ash reinjection	17A	8.5A	38S	19S	0.5S	0.25S	0.03	0.015	1	0.5	18	9
Overfeed stokers, no fly-ash reinjection[f]	2A	1A	38S	19S	0.5S	0.25S	0.2	0.1	(2 - 10)[g]	(1 - 5)	(6 - 15)[h]	(3 - 7.5)
Hand-fired units	10	5	36S	18S	0.8S	0.4S	2.5	1.25	90	45	3	1.5

[a]A is the ash content expressed as weight percent.
[b]S is the sulfur content expressed as weight percent.
[c]Expressed as methane.
[d]Based on bituminous coal combustion.
[e]Emitted as NO, but calculated as NO_2.
[f]Based on data obtained from traveling-grate stokers in the 12 to 180 Btu/hr (3 to 45 kcal/hr) heat input range. Anthracite is not burned in spreader stokers.
[g]Use high side of range for smaller-sized units [less than 10 x 10^6 Btu/hr (2.5 x 10^6 kcal/hr) heat input].
[h]Use low side of range for smaller-sized units [less than 10 x 10^6 Btu/hr (2.5 x 10^6 kcal/hr) heat input].

TABLE 1-2

EMISSION FACTORS FOR BITUMINOUS COAL COMBUSTION WITHOUT CONTROL EQUIPMENT[3]

Furnace Size, 10^6 Btu/hr Heat Input[a]	Particulate[b]		Sulfur Oxides[c]		Carbon Monoxide		Hydrocarbons[d]		Nitrogen Oxides		Aldehydes	
	lb/ton	kg/MT	lb/ton	kg/MT	lb/ton	kg/MT	lb/ton	kg/MT	lb/ton	kg/MT	lb/ton	kg/MT
Greater than 100 (utility and large industrial boilers)												
Pulverized												
General	16A	8A	38S	19S	1	0.5	0.3	0.15	18	9	0.005	0.0025
Wet Bottom	13A[e]	6.5A	38S	19S	1	0.5	0.3	0.15	30	15	0.005	0.0025
Dry Bottom	17A	8.5A	38S	19S	1	0.5	0.3	0.15	18	9	0.005	0.0025
Cyclone	2A	1A	38S	19S	1	0.5	0.3	0.15	55	27.5	0.005	0.0025
10 to 100 (large commercial and general industrial boilers)												
Spreader Stoker[f]	13A[g]	6.5A	38S	19S	2	1	1	0.5	15	7.5	0.005	0.0025
Less than 10 (commercial and domestic furnaces)												
Spreader Stoker	2A	1A	38S	19S	10	5	3	1.5	6	3	0.005	0.0025
Hand-fired Units	20	10	38S	19S	90	45	20	10	3	1.5	0.005	0.0025

---

[a] 1 Btu/hr = 0.252 kcal/hr.

[b] The letter A on all units other than hand-fired equipment indicates that the weight percentage of ash in the coal should be multiplied by the value given. Example: If the factor is 16 and the ash content is 10 percent, the particulate emissions before the control equipment would be 10 times 16, or 160 pounds of particulate per ton of coal (10 times 8, or 80 kg of particulates per MT of coal).

[c] S equals the sulfur content (see footnote b above).

[d] Expressed as methane.

[e] Without fly-ash reinjection.

[f] For all other stokers use 5A for particulate emission factors.

[g] Without fly-ash reinjection. With fly-ash reinjection use 20A. This value is not an emission factor but represents loading reaching the control equipment.

TABLE 1-3

EMISSIONS FROM LIGNITE COMBUSTION WITHOUT CONTROL EQUIPMENT[3a]

Pollutant	Type of Boiler: Pulverized Coal, lb/ton	Pulverized Coal, kg/MT	Cyclone, lb/ton	Cyclone, kg/MT	Spreader Stoker, lb/ton	Spreader Stoker, kg/MT	Other Stokers, lb/ton	Other Stokers, kg/MT
Particulate[b]	7.0A[c]	3.5A[c]	6A	3A	7.0A[d]	3.5A[d]	3.0A	1.5A
Sulfur Oxides[e]	30S	15S	30S	15S	30S	15S	30S	15S
Nitrogen Oxides[f]	14(8)[g,h]	7(4)[g,h]	17	8.5	6	3	6	3
Hydrocarbons[i]	<1.0	<0.5	<1.0	<0.5	1.0	0.5	1.0	0.5
Carbon Monoxide[i]	1.0	0.5	1.0	0.5	2	1	2	1

[a]All emission factors are expressed in terms of pounds of pollutant per ton (kilograms of pollutant per metric ton) of lignite burned, wet basis (35 to 40 percent moisture, by weight).
[b]A is the ash content of the lignite by weight, wet basis.
[c]This factor is based on data for dry-bottom, pulverized-coal-fired units only. It is expected that this factor would be lower for wet-bottom units.
[d]Limited data preclude any determination of the effect of fly-ash reinjection. It is expected that particulate emissions would be greater when reinjection is employed.
[e]S is the sulfur content of the lignite by weight, wet basis. For a high sodium-ash lignite (Na_2O >8 percent), use 17S lb/ton (8.5S kg/MT); for a low sodium-ash lignite (Na_2O <2 percent), use 35S lb/ton (17.5S kg/MT). For intermediate sodium-ash lignite, or unknown sodium-ash content, use 30S lb/ton (15S kg/MT).
[f]Expressed as NO_2.
[g]Use 14 lb/ton (7 kg/MT) for front-wall-fired and horizontally opposed-wall-fired units and 8 lb/ton (4 kg/MT) for tangentially fired units.
[h]Nitrogen oxide emissions may be reduced by 20 to 40 percent with low excess air firing and/or staged combustion in front-fired and opposed-wall-fired units and cyclones.
[i]These factors are based on the similarity of lignite combustion to bituminous coal combustion.

$$0.92 \frac{ton}{hr} \times 13\ (7) \frac{lb}{ton} = 84 \frac{lb}{hr}$$

(Particulate Emission Rate)

The uncontrolled emissions from a boiler can be reduced by the use of particulate emission control devices. Table 1-4 gives the range of collection efficiencies[3] for some of these devices when used on a particular type of boiler. All the control devices are discussed in more detail later in this manual; however, Table 1-4 can be used to estimate particulate emission from controlled boilers. The controlled emissions can be estimated using the following:

$$C = U - (U \times eff)$$

where C = the controlled emission rate in lb/hr
U = the uncontrolled emission rate in lb/hr
eff = the control device removal efficiency (from Table 1-4) expressed as a decimal.

For example, if the particulate emissions from the above mentioned spreader stoker boiler were controlled by a high efficiency cyclone, the controlled emission would be calculated as follows:

$$C = U - (U \times eff)$$
$$C = 84\ lb/hr - (84\ lb/hr \times 0.85)$$
$$C = 12.6\ lb/hr$$

B. Sulfur Oxides Emissions - Sulfur oxides (SO_x) emissions for coal-fired boilers depend on the sulfur content of the coal and on control methods used. Control of SO_x emissions from coal is discussed in Section 1.1.7. Uncontrolled SO_x emission can be estimated by multiplying the fuel usage rate by the proper emission factor. For anthracite and bituminous coal, the proper emission factor is 38S lb/ton, where S is the weight percent sulfur in the coal (Tables 1-2). For lignite, the proper emission factor is 30S, where S is the weight percent sulfur in the lignite (Table 1-3). Using the boiler in the particulate emission section as an example, the SO_x emissions would be estimated as follows:

Given that the sulfur content of the coal is 4 percent by weight and the fuel usage rate of 0.92 ton/hr,

TABLE 1-4

RANGE OF COLLECTION EFFICIENCIES FOR COMMON TYPES OF FLY-ASH CONTROL EQUIPMENT[3]

Type of Furnace	Range of Collection Efficiencies %			
	Electrostatic Precipitator	High-Efficiency Cyclone	Low-Resistance Cyclone	Settling Chamber Expanded Chimney Bases
Cyclone Furnace	65 to 99.5[a]	30 to 40	20 to 30	10[a]
Pulverized Unit	80 to 99.5[a]	65 to 75	40 to 60	20[a]
Spreader Stoker	99.5[a]	85 to 90	70 to 80	20 to 30
Other Stokers	99.5[a]	90 to 95	75 to 85	25 to 50

[a]The maximum efficiency to be expected for this collection device applied to this type source.

$$SO_x \text{ emissions} = 0.92 \frac{\text{ton}}{\text{hr}} \times 38(4) \frac{\text{lb}}{\text{ton}} = 140 \frac{\text{lb}}{\text{hr}}$$

C. Oxides of Nitrogen - Emission of the oxides of nitrogen (NOx) from coal-fired boilers is related to the stokering system used and the boiler size. Uncontrolled NOx emissions can be estimated by multiplying the fuel usage rate by the emission factor found in Table 1-1, 1-2, or 1-3. For example, the NOx emissions for the boiler described in A above would be calculated as follows:

The fuel usage rate (calculated in A above) is 0.92 ton/hr and the boiler is 24 x 10^6 Btu/hr spreader stoker. The emission factor from Table 1-2 is 15 lb/ton.

$$\text{NOx emissions} = 0.92 \text{ ton/hr} \times 15 \text{ lb/ton} = 13.8 \text{ lb/hr}$$

Emissions of carbon monoxide and hydrocarbons can be similarly estimated by using Tables 1-1, 1-2, and 1-3.

1.1.2.2 Coal-fired Stoker Boilers

Stokers of various types are designed to automatically feed coal uniformly onto a grate within the combustion zone of the furnace and to discharge the remaining ash residue to some collection device. Mechanical stokers can most easily be classified in three basic groups, dependent upon the direction and method by which the raw coal reaches the fuel bed and the nature of the bed. The three main groups include:

1. overfeed: traveling grate, vibrating grate, and oscillating grate
2. spreader
3. underfeed.

For the larger steam installation (400,000 to 5,000,000 lb/hr), stokers have not proven themselves as economical as pulverized-coal or cyclone furnaces. The spreader stoker dominates the middle-to-upper size of stoker units, while the smaller sizes are dominated by traveling-grate stokers. Other less commonly used stokers include the vibrating grate and underfeed stokers.

1.1.2.2.1 Overfeed Stokers--Traveling Grate

Traveling grate stokers consist of assembled links, grates, or keys joined together in an endless-belt arrangement at the base of the furnace. This flexible

grate passes around two sprokets located at the front and rear of the furnace. Coal, which is broken up into approximately 2-inch particles, is fed by gravity from a hopper onto the moving grate assembly and enters the front of the furnace, where it passes under an adjustable gate that regulates the thickness of fuel in the fuel bed. As hot furnace gases react with the hydrocarbons and other combustible distillation products from the coal, combustion takes place. The bed continues to burn as it moves along, and the bed thickness becomes correspondingly smaller, until at the far end of the furnace, the ash is dumped over the end of the grate into the ash pit when the grate arches over the rear sprocket.

The air necessary to support combustion is introduced either from above, below, or up through the grate supporting the fuel bed, via jets sometimes referred to as "overfire jets." Fuel used in this type of stoker can be anthracite, sub-bituminous, weakly caking bituminous coal, lignite, and occasionally coke. The stoker works best with non-caking fuels (high ash-fusion temperature) because air distribution is most uniform through an even bed of non-caked masses. Traveling grates are most commonly used at installations in the 5,000 to 75,000 pounds of steam per hour capacity range.

1.1.2.2.2 Overfeed Stokers--Vibrating Grate

The vibrating grate overfeed stoker burns low ranking (lignite and brown) coals with much higher efficiency than had been possible previously and with exceptionally low maintenance. Since its introduction in the United States, this stoker has found acceptance in both industrial and institutional plants and has burned a wide range of American coals successfully. As with traveling grate stokers, coal is hopper-fed, and the thickness of the bed is regulated by an adjustable gate. The vibrating force is provided by a generator located at the front of the stoker beneath the coal hopper. Air to the stoker is distributed through laterally disposed compartments beneath the stoker. During vibration, new coal is fed to the stoker as the entire fuel bed is simultaneously conveyed down the inclined grate, and ash is discharged into the ash hopper. To complete carbon burnout of the ash, the thickness of the ashbed portion is controlled by an adjustable ash-dam plate.

Fly ash may be slightly higher than the traveling grate stokers because of increased agitation of the fuel bed. The water-cooled vibrating grate stoker is an adaptation of the design; it is also capable of burning better grades of coal. Because of simplicity, inherent low fly-ash carry-over characteristics, and very low maintenance, this stoker has been steadily gaining acceptance and now has replaced larger size multiple retort underfeed stokers in the 75,000 to 300,000 pounds of steam per hour intermediate range.

1.1.2.2.3 Spreader Stokers

The spreader stoker employs either a mechanical spreader or jets of steam or air to throw the solid fuel into the furnace, where it falls on a grate that is either traveling or stationary. The coal is spread on the grate in such a fashion that it falls off at the end of the grate into an ash collection hopper and is removed from the furnace.

The essential difference between the spreader stoker and the traveling grate stokers is that the former utilizes a combination of suspension burning and a thin fast-burning fuel bed, whereas the latter employs a bed-thickness-load-related mass-bed burning principle. The spreader stoker is considered to be extremely sensitive to load fluctuations, since ignition is almost instantaneous on increase of firing rate, and burnout of the thin fuel bed can be rapidly effected. These stokers are also flexible in their ability to burn a wide range of fuels from high-rank Eastern bituminous to lignite or brown coals, as well as fine coals. Optimum coal sizing for this type of firing is 3/4 inch x 0 with about 50 percent passing a 1/4-inch screen.

Because there is some burning of the fuel in suspension, spreader stoker firing produces a much higher density of particulate matter in the flue gases than is the case with mass-burning equipment, such as a traveling grate. Dust collectors are required to remove the large particulate before the flue gases are discharged to the stack or reinject carbonaceous particulate back into the furnace to reduce combustible losses and to improve boiler efficiency (e.g., 2 to 3 percent).

Traveling grate spreader stokers are economically feasible for boiler capacities in the 75,000 to 400,000 pounds of steam per hour range. For larger capacities, the required grate area may run the boiler cost beyond that for other methods of coal firing.

1.1.2.2.4 Underfeed Stokers

In underfeed stokers, fuel is force-fed by rams to the fuel bed in small increments. The coal moves in a longitudinal channel, known as a retort, with the help of small pusher blocks that operate in conjunction with the ram. When the retort is full, the fuel is forced upwards and spills over the top on each side to form and feed the fuel bed. Air is supplied by tuyeres at each side of the retort and via air openings in the side grates.

As the coal rises, heat travels downward from the fuel burning above. Volatile gases are distilled off and

burned as they pass through the incandescent fuel bed. The rising fuel then ignites from contact with the actively burning bed. The pressure exerted by the incoming raw coal moves the bed gradually over the tuyeres and side grates while burning continues. Combustion is completed by the time the bed reaches the side-dump grates from which the ash is discharged to shallow pits.

A fairly wide variety of coals can be used with underfeed stokers. A desirable fuel is free-burning bituminous coal; however, caking coals are also desirable because the volatiles are released more slowly from caked masses than from a porous bed of coals. Anthracite can also be burned either separately or mixed with bituminous coal. Underfeed beds are inherently smoke-free because the air and fresh fuel flow concurrently, usually upwards. Low-fly ash carry-over is exhibited similar to other mass-burning furnaces.

The single- or double-retort horizontal underfeed stokers are generally designed to service boiler units producing up to 30,000 pounds of steam per hour continuously. Another type, the multiple-retort gravity-feed stoker, can be designed for capacities suitable for boiler units generating from 20,000 to 500,000 pounds of steam per hour. Burning rates are directly related to the ash-softening temperature. For coals with ash-softening temperatures below 2,400°F, the burning rates will be progressively reduced.

1.1.2.3 Pulverized Coal Boilers

In pulverized-fuel steam generators, the coal is finely ground to approximately a 200-mesh particle size in pulverizers. It is then combined with a hot air stream from the air preheater to accommodate drying and is pneumatically conveyed to the boiler nozzles, where it is injected into the burner zone of the boiler. Fine grinding of coal is necessary to assure complete combustion of the carbon and to avoid the deposit of ash and carbon on the heat absorbing surfaces. This uniform grinding before firing also produces fly-ash emissions that are somewhat homogeneous in terms of particle size distribution. This fact becomes important when considering what type of collection device would best control particulate emissions from pulverized coal boilers. Several arrangements for nozzle configuration exist, including tangential firing, horizontally-opposed burner, and front wall burner.

The major components of the direct-firing system, in general, are as follows:

1. Air heater to supply hot air to the pulverizers for drying the coal as it is pulverized.

2. Pulverizer fan, known as the primary-air fan, arranged either as an exhauster or a blower.
3. Pulverizer arranged to operate under suction or pressure.
4. Automatically controlled raw coal feeder.
5. Coal and air conveying lines.
6. Burners.

Because of the capital costs required for small pulverized coal plants, pulverized coal units smaller than 100,000 pounds of steam per hour are uneconomical. In larger units where the furnace size and configuration are less disproportionate, lower operating costs result from increased efficiency.

Although many types of coal can be used successfully in pulverized-coal-firing units, most pulverizers react adversely to handling high moisture coal and the grinding capacity will drop as the surface moisture increases.

The primary coal characteristics that influence the design and operation of this type of equipment are grindability, coal rank, coal moisture content, coal volatile matter, and ash content. The usual limits in these characteristics for pulverized-coal-firing are:

1. Maximum total moisture, as fired, 15 percent (although higher moisture content coal such as sub-bituminous and lignite may be used).
2. Minimum volatile matter on dry basis, 15 percent.
3. Maximum total ash on dry basis, 20 percent.

1.1.2.4 Particulate Emissions from Coal Combustion

Fly ash is largely produced by the combustion of coal. The mineral matter of the coal is converted to ash, part of which is discharged from the bottom of the furnace as a solid or as molten slag, and part of the coal mineral content is discharged from the furnace as particles suspended in the flue gas. The particles are mostly spherical; some are transparent or translucent and hollow (cenospheres). However, irregularly shaped and opaque particles may be present to a small extent if the fusion temperature of the ash is high or if combustion of the coal is incomplete. The sizes of the particles vary over a wide range from submicron sizes up to the largest sizes that can be suspended in the flue gas. The average or mass median particle diameter is usually 10 to 20 μm. Particles with diameters of 1 μm or less will typically account for 1 to 10 percent of the weight of the fly ash.

The particle size range and the average particle diameter depend on the type of furnace and how it is

operated. The coal-fired boilers are stoker-fired, or pulverized-coal-fired. Stoker-fired boilers are often used in industrial or small utility systems. About half the mineral content of the coal is emitted as fly ash from these boilers. Pulverized-coal-fired boilers are most widely used in electric utility power plants. They emit 70 to 100 percent of the coal ash in the form of fly ash. Stoker-fired boilers produce a coarser fly ash than pulverized-coal fired boilers. However, there is considerable overlap in the particle-size ranges of fly ash. Average particle size distributions resulting from coal combustion[4] are shown in Figures 1-2 and 1-3.

1.1.2.5 Particulate Control in Coal-fired Plants

Electrostatic precipitators (ESP) have been the most popular air pollution control system, often in series with a mechanical collector. However, the recent promulgation of a more stringent particulate emission regulation has resulted in the economic competitiveness of fabric filters. Electrostatic precipitators, fabric filter baghouses, and wet scrubbers all can be designed and operated to be highly efficient in collecting fly ash to the 99 to 99.9 percent level on an overall mass basis.

When compared on the basis of fractional collection efficiency, specifically for the collection of fly-ash particles below 1 μm in diameter, fabric filters have demonstrated collection efficiencies in excess of 99 percent. Electrostatic precipitators in the same particulate size range are not as effective as fabric filters. The collection efficiency of wet scrubbers for submicron particles can also be effective; however, the relatively high pressure drop requirements tend to eliminate scrubbers from consideration due to energy costs.

Nonuniform gas flow distribution, changes in dust loading of the gas, and the chemical and electrical properties of the fly ash have less effect on the collection efficiencies of a fabric filter or a wet scrubber than on the efficiency of an electrostatic precipitator.

1.1.2.5.1 Electrostatic Precipitators

From a practical as well as a theoretical standpoint, the factors governing precipitator performance include:

1. particle size distribution and concentration of the entering dust
2. area and type of collection electrodes
3. electrical conditions within the precipitator (current and voltage)
4. degree of precipitator sectionalization

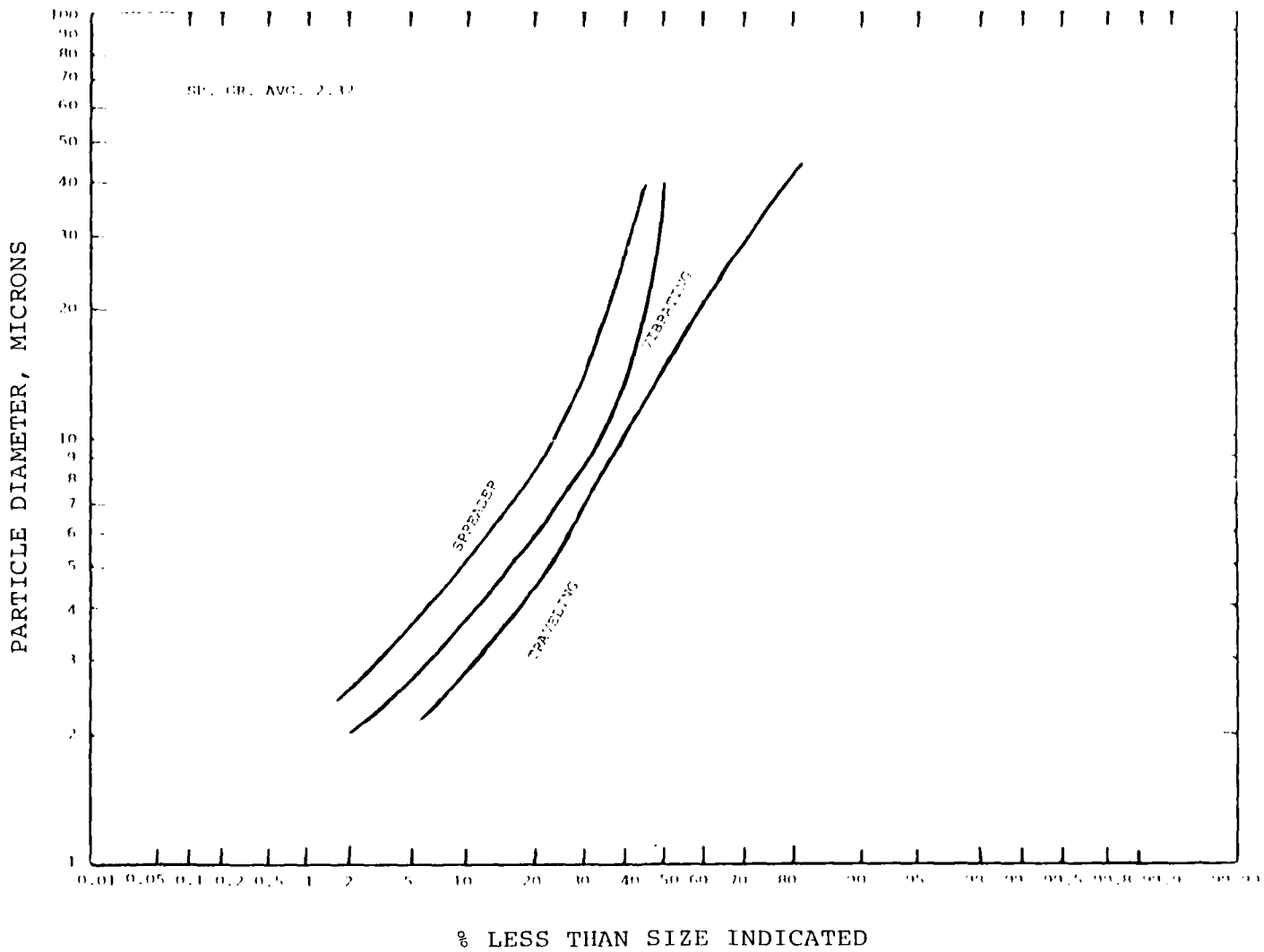

FIGURE 1-2

PARTICLE SIZE DISTRIBUTION STOKERS[4]

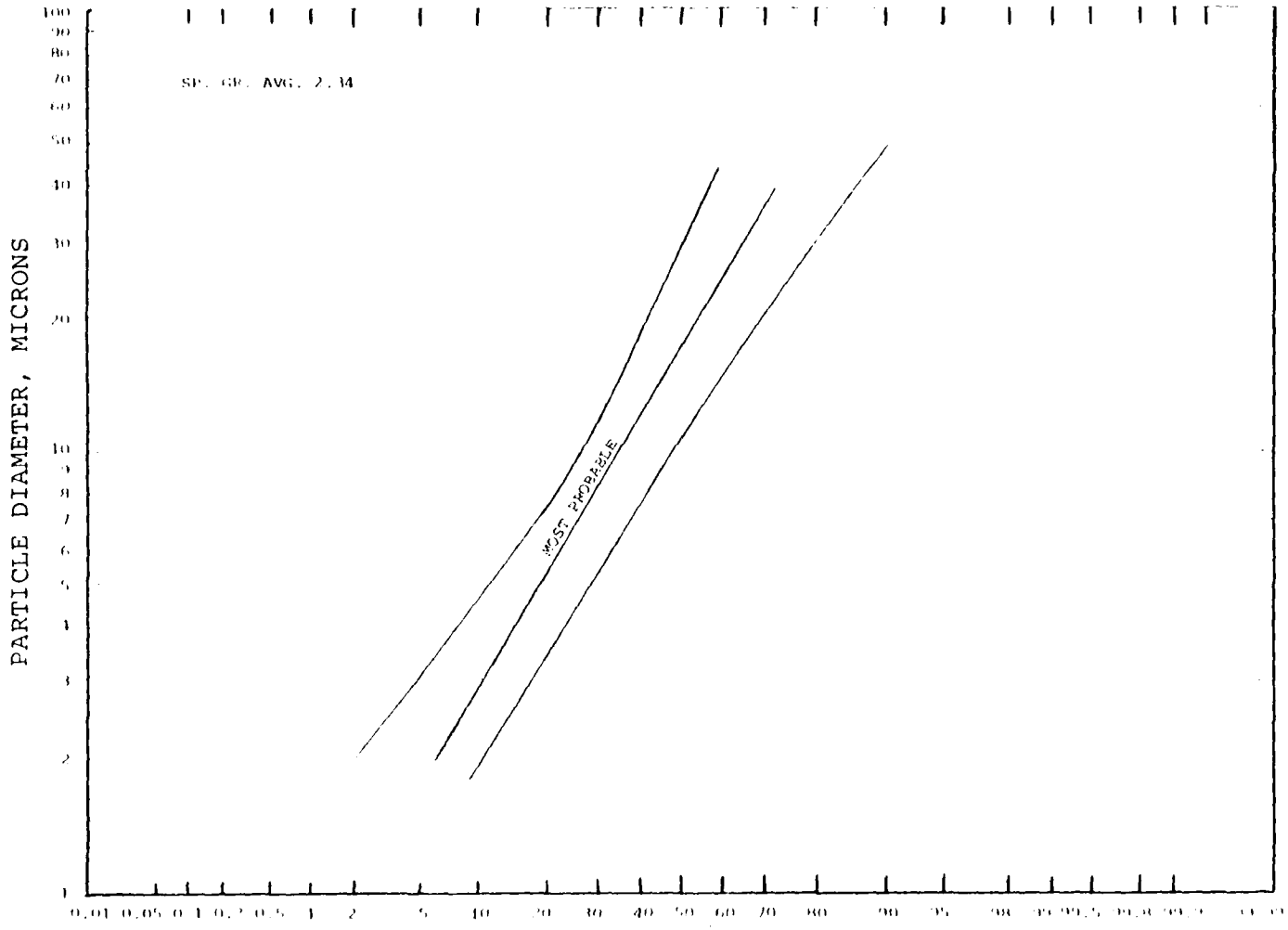

FIGURE 1-3

PARTICLE SIZE DISTRIBUTION OF PULVERIZED COAL[4]

5. degree of reentrainment of the dust
6. uniformity of the gas flow.

As higher efficiencies are required, a larger percentage of the finer size particulate must be collected. This change requires a proportionately larger precipitator size (collecting area) to go, for example, from an efficiency of 99 to 99.9 percent than to go from 95 to 99 percent. Some typical operating characteristics of coal-fired boilers controlled by ESP's[4] are presented in Table 1-5.

The determination of necessary plate area and type of collection electrodes involves consideration of the fly-ash characteristics and the amount of electrical energization equipment required to realize a desired collection efficiency. Precipitators used for fly-ash removal are generally installed following the air preheater, at which point the flue gas temperature is typically about 300°F. Some precipitators are installed ahead of the air preheater (hot side), where the flue gas temperatures are usually around 600 to 800°F, and the fly-ash resistivity is sufficiently low so that the current is not resistivity limited.

Fly-ash resistivity is a property affected principally by the composition of the coal. Combustion of coal containing sulfur results in sulfur oxides, principally sulfur dioxide (SO_2), in the flue gases. Except perhaps for that in the form of metal sulfates, all of the sulfur in the coal appears in the flue gases. About 1 percent of the SO_2 is oxidized to SO_3, which combines with the moisture present in the flue gas to form sulfuric acid vapor. At the temperatures at which most precipitators operate (around 300°F or lower), some of the sulfuric acid vapor present is condensed or is adsorbed on the fly-ash surface. The presence of this sulfuric acid on the fly-ash reduces the electrical resistivity of the ash in relation to the quantity of sulfuric acid present and the temperature of the flue gases. In general, coals containing sulfur in quantities greater than about 1.5 percent produce fly ash with resistivities in a range that will enable satisfactory collection at temperatures around 300°F in a conventional precipitator.[3] Coals with sulfur in quantities less than 1.5 percent generally produce ash with moderate to relatively high resistivities. These require either comparably larger plate areas (specific collecting area) or precipitator design and installation on the "hot side."

Another factor influencing precipitator performance is reentrainment of collected dust, or bypassing of the dust-laden gases around the electrified regions. Reentrainment can take place in a variety of ways. Dust being precipitated can strike the dust layer with sufficient velocity to dislodge additional particles. Rapping

TABLE 1-5

OPERATING CHARACTERISTICS OF PULVERIZED-COAL-FIRED BOILERS CONTROLLED BY ELECTROSTATIC PRECIPITATORS[4]

Type of Precipitator	Coal Sulfur (%)	Stack Gas Flow Rate (1000 acfm)	Stack Gas Temp. (^{o}F)	Precipitation Rate Parameter (ft/sec)	Collection Efficiency	Collecting Area (Ft^2/1000 acfm)
Cold	0.9	385	285	0.341	98.0	191
Hot	0.7-5.0	714	655	0.292	99.0	250
Hot	1.2-1.4	163	705	0.282	99.3	295
Hot	1.2-1.4	440	775	0.289	99.3	288
Cold	2.2-9.0	202	306	0.279	97.0	208
Cold	0.7	575	320	0.164	99.6	560
Cold	0.6	286	300	0.217	99.0	314
Cold	0.6	415	300	0.217	99.0	355
Cold	1.9	300	400	0.344	96.3	158

of the plates can cause dust to be reentrained into the gas stream. Dust settling into the hoppers can also result in reentrainment of dust already collected. Gas sneakage resulting from maintanence of the electrical clearance above and below the electrodes as well as gas sweepage through the hoppers also influence performance.

The influence of gas sneakage and reentrainment is reduced considerably by proper baffling and by proper design of rapping gear. Reentrained dust is generally not dispersed as individual particles, so the redispersed dust is more readily collected than would be indicated by the sizes of the discrete particles. Dust reentrained from the first field is readily collected in the subsequent fields. The reentrained dust appearing at a precipitator outlet is principally that resulting from reentrainment of dust collected in the last field.

Nonuniform gas flow is also an important factor in precipitator performance. The size of the flue gas ducts and space limitations make it difficult to achieve good uniformity of gas flow. As a standard, the Industrial Gas Cleaning Institute has recommended that the gas entering a precipitator should be such that 85 percent of the local velocities should be within 25 percent of the mean and no single point should differ more than +40 percent.

1.1.2.5.2 Fabric Filters

Factors that can influence fabric filter performance on fly-ash collection include the fabric structure, air-to-cloth ratio, maximum pressure drop before cleaning, method of cleaning, cleaning frequency, intensity of cleaning, and flue gas temperature, and humidity. These factors can be broadly classified into two categories: factors that are basic to the design of a fabric filter system for optimum performance and factors that relate to the behavior of the system as installed. During the design of a baghouse, questions of fabric, method of cleaning, air-to-cloth ratio, operating temperature, and humidity need to be considered. Once the baghouse is operating, cleaning rate, duration and intensity of cleaning, and other variables related to the operation of the collector are adjusted to keep pressure drop before cleaning below the maximum allowable.

Other occurrences can influence the collection efficiency of the baghouse. Bags can be blinded or clogged during boiler start-up. Temperature excursions through acid dew points may corrode the baghouse structure and adversely affect the life and performance of the

bags. Another potential problem is air preheater failure, which can drive temperatures above safe operating limits fo the bags.

Fabric filter performance is not influenced by some of the factors that are considered critical in electrostatic precipitator and scrubber operation. Woven fabric filters are relatively insensitive to inlet dust loading. Also, fly-ash resistivity is of little consequence for fabric collectors since collection does not depend on electrostatic attraction. In contrast to the operation of a scrubber, fabric filter collection efficiency is not a function of particle wettability or pressure drop.

The pressure drop across a baghouse is of concern in order to minimize total system energy requirements, which can seriously impact operating costs. An operating pressure drop of 3 to 4 inches (water gauge) is typical for fly-ash collection; however, some baghouses may require up to 10 inches water gauge. Pressure drop numbers are values averaged over the filtering cycle. When filtration begins, the pressure drop increases nonlinearly until a filter cake is built and then increases linearly thereafter. Some typical operating characteristics of coal-fired plants controlled by fabric filters[4] are presented in Table 1-6.

1.1.2.5.3 Wet Scrubbers

The information available on scrubbers indicates that they can be designed to remove fly-ash efficiently on an overall mass basis. Unfortunately, pressure drop requirements may be excessive and eliminate the wet scrubbers from consideration on the basis of operating costs. Furthermore, the particulate is collected wet and requires further treatments, increasing the operating costs even more.

In making the particle-size measurements on scrubber installations, interference is usually encountered from the entrainment of liquid droplets in the gas emerging from the scrubber, even though the scrubbers are usually equipped with entrainment separators. This has been a common problem in sampling fly-ash scrubber effluents. As a consequence, it is difficult to compare the data obtained with similar data on other installations.

1.1.2.6 Classification of Coals

The type and amounts of pollutants from coal-fired plants are strongly dependent upon the type of coal being used. It is, therefore, necessary to fully understand the different types of coals available. As a guide, three

TABLE I-6

OPERATING CHARACTERISTICS OF COAL-FIRED BOILERS CONTROLLED BY FABRIC FILTER UNITS [4]

Type of Boiler	Coal Sulfur (%)	Coal Ash (%)	Stack Gas Flow Rate (1000 acfm)	Stack Gas Temp. (°F)	Air Cloth (acfm/Ft/^{2})	Pressure Drop (In. WG)	Type of Fabric	Type of Cleaning
Pulverized	1.8	22	900	325	2.0	2.5	Woven glass fiber	Reverse flow
Pulverized	NA	NA	200	360	2.4	4.5	Woven glass fiber	Reverse flow
Pulverized	1.0	11	550	325	2.0	3.0	Woven glass fiber	Reverse flow
Stoker	0.7	12	258	360	2.8	4.5	Woven glass fiber	Reverse flow and shake
Stoker	1.0	10	60	325	4.0	2.8	Woven glass fiber	Reverse flow

standard methods of classifying coal have been adopted in the United States. These classifications are:

1. rank (degree of metamorphism or progressive alteration, in the natural series from lignite to anthracite)
2. grade (quality determined by size designation, calorific value, ash, ash-softening temperature, and sulfur)
3. type or variety (determined by the nature of the original material and subsequent alteration thereof).

Other less significant methods of coal classifications are by its use or suitability for specific purposes or types of combustion equipment, and by various trade systems set up to meet particular conditions in a given area or time.

1.1.2.6.1 Classification by Rank

The most well known of the classification systems is by rank, as defined by ASTM D388-38, in which coals are arranged according to fixed carbon content and calorific value (Btu content) calculated on the mineral-matter-free basis. The higher ranking coals are classified according to fixed carbon on a dry basis; the lower ranking coals, according to Btu content on a moist basis (containing the natural bed moisture). Agglomerating and weathering indexes are used to differentiate between certain adjacent groups. The classification system summarized in Table 1-7 is based on the ASTM rank classification system.[5] Several parameters other than those appearing in the ASTM system have been included. These values are included only to give approximate ranges for areas of classification not as carefully considered in the ASTM system. Chosen values are based on experimental findings using many types of coal from various parts of the country. The complete standard located in ASTM Standards on Coal and Coke contains more in-depth details of classification, methods of sampling, analysis, testing, and calculation of test results. A graphic representation of the ASTM rank classification system[5] with some additions is depicted in Figure 1-4.

1.1.2.6.2 Classification by Grade

Classification by grade provides a designation system indicating size, Btu content, ash content, ash-softening temperature, and sulfur content of various coals. Type classification includes categories such as common banded coal, splint coal, cannel coal, and boghead coal. For more detailed classification and/or descriptions of coal classification systems, the reader is advised to see ASTM Standards on Coal and Coke, 1948.

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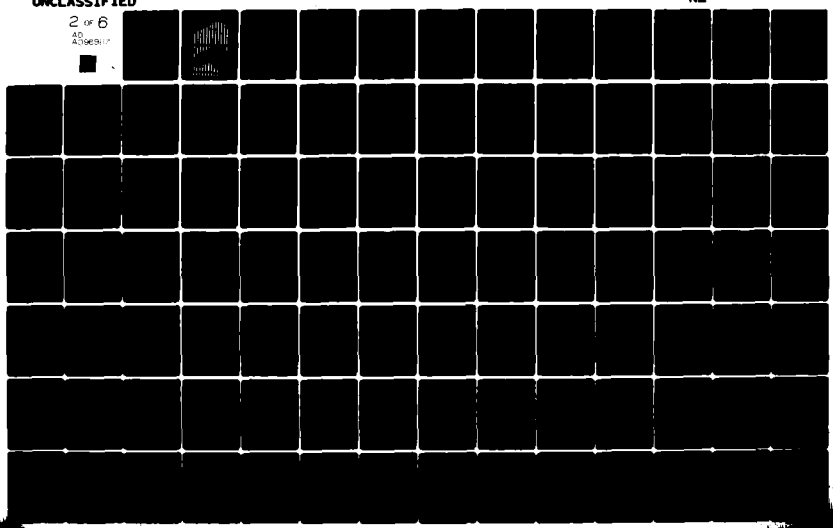


TABLE 1-7

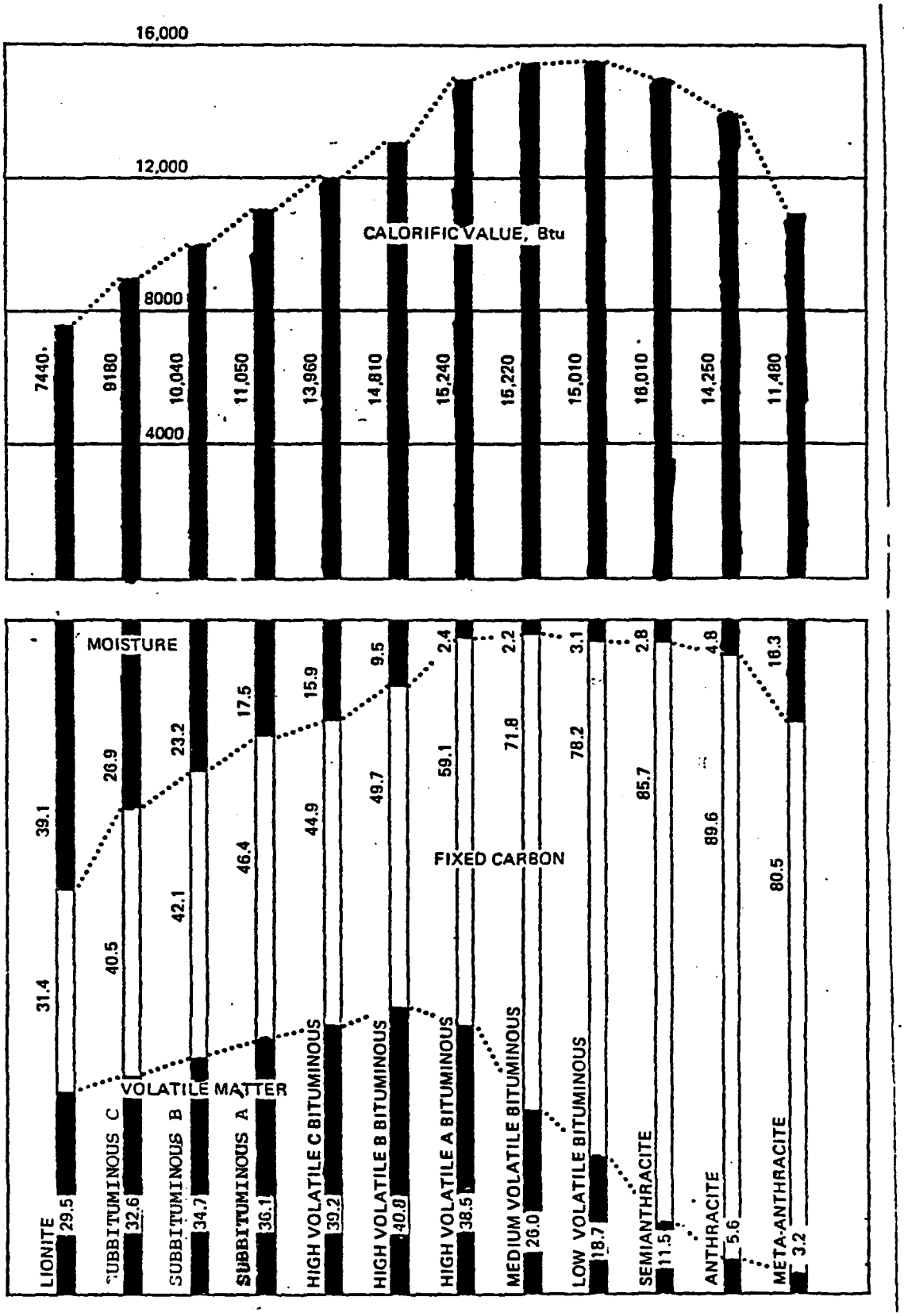
COAL CLASSIFICATION (MOIST, MINERAL-MATTER-FREE)[5]

	Btu/lb	Fixed Carbon	Volatile Matter	Bed Moisture	Ash
I. Anthracite					
1. Meta Anthracite	12,500*	98	2*	2-3	5-10
2. Anthracite	15,000*	92-98*	2-8*	2-3	10-15
3. Semi Anthracite	15,500*	86-92*	8-14*	2-3	10-15
II. Bituminous					
1. Low Volatile Bituminous	15,200*	78-86*	14-22*	3-4	5-10
2. Medium Volatile Bituminous	15,200*	69-78*	22-31*	3-4	5-10
3. High Volatile A Bituminous	14,000*	69*	31*	4-6	5-10
4. High Volatile B Bituminous	13-14,000	43-50	31-26	6-10	10-15
5. High Volatile C Bituminous	12,100	38-43	31-36	10-15	10-15
III. Sub-Bituminous					
1. Subbitum. A	11-13,000	35-42	31-36	15-20	10-15
2. Subbitum. B	9,500-11,000	35-42	30-33	20-25	10-15
3. Subbitum. C	8,300-9,500	35-42	30-33	20-25	10-15
IV. Lignite					
1. Lignite	<8,300	<35	<30	>25	10
2. Brown Coal	<8,300	<35	<30	>25	10

* Dry basis.

Figure 1-4

GRAPHIC REPRESENTATION OF ASTM COAL CLASSIFICATION[5]



Grindability is the term used to measure the ease of pulverizing a coal in comparison with a standard coal chosen as 100 (unity) grindability. A description of the method of testing to determine grindability of a coal can be found in ASTM Standard Method D409.

The ASTM Standard Method D720 is used for obtaining information regarding the free-swelling property of a coal. Since it is a measure of the behavior of coal when heated rapidly, it may be used as an indication of the coking characteristic of the coal when burned as a fuel.

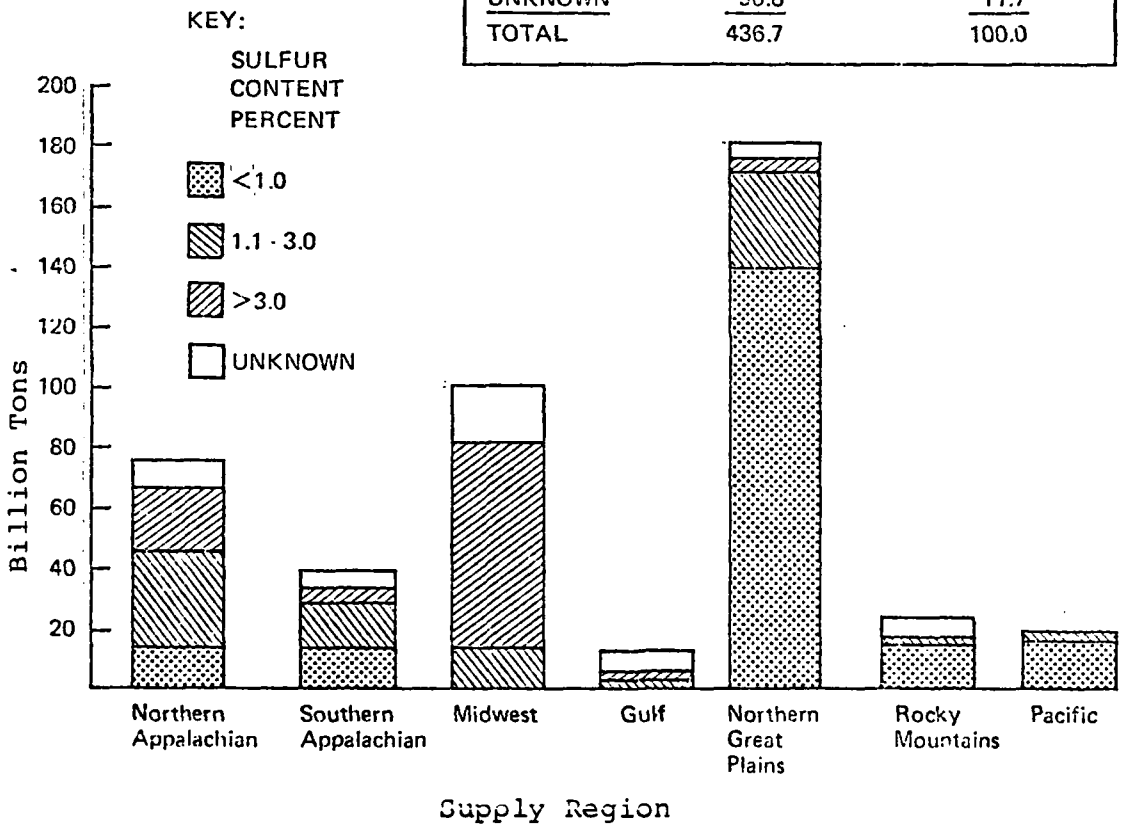
1.1.2.6.3 Classification by Variety

Sulfur in coal is a major concern with regard to the environment and the effects of combustion on it. The geographic region in which the coal is mined has a great deal to do with the sulfur content of a coal. As an example, 80 to 90 percent of Midwestern coal is high sulfur coal (>3.0% S), whereas 80 percent of the Northern Great Plains coal is low sulfur coal (<1.0% S). Figure 1-5 illustrates the estimated domestic coal reserves[5] and the levels of sulfur contained in those coals. The burning of low sulfur coal can, in many instances, eliminate the need for costly SO_2 scrubbers and still maintain appropriate air quality standards. However, the supply of low sulfur coal is, like all resources, a finite one, and conversion to other coals containing higher or lower levels of sulfur should be anticipated at some point during the life of the combustion device.

FIGURE 1-5

COAL RESERVES[5]

TOTAL DOMESTIC COAL RESERVES		
SULFUR LEVEL, PERCENT	AMT COAL AVAILABLE, BILLION TONS	PORTION OF TOTAL COAL RESERVED, PERCENT
<1.0	200.2	45.8
1.1 - 3.0	93.0	21.3
>3.0	92.7	21.2
UNKNOWN	50.8	11.7
TOTAL	436.7	100.0



1.1.3 Oil-fired Boilers

An oil-fired boiler uses atomization to burn the fuel oil in the furnace. The oil is dispersed into the furnace as a fine mist which exposes a large amount of oil particle surface to the combustion air to assure prompt ignition and rapid combustion. The two most popular ways of atomizing fuel oil are by use of steam or air. Arrangements of nozzle configuration are the same as pulverized coal furnaces: tangential, horizontally opposed, and front wall.

Compared with coal, fuel oils are relatively easy to handle and burn. Heating is not required for the lighter oils, and even the heavier oils are relatively simple to handle. There is not as much ash-in-bulk disposal problem as there is with coal, and the amount of ash discharged from the stack is correspondingly small. In the atomized state, the characteristics of oil approach those of a gas, with consequent similar explosion hazards.

Because of its relatively low cost compared with that of lighter oils, No. 6 fuel oil is the most widely used for steam generation. Its ash content ranges from about 0.01 to 0.5 percent, which is very low compared with coal. However, despite this low percentage content, ash containing compounds of vanadium, sodium, and sulfur can be responsible for a number of serious operating problems.

Fuel oils generally available on the Eastern seaboard are produced from varying amounts of Venezuelan and Middle East crudes, depending on the relative quantities of shipments and on the blending at the refineries to meet No. 6 fuel oil viscosity specifications. Fuel oils used on the Gulf and West Coasts are produced largely from domestic crudes, although they may contain some Venezuelan crude. Due to the distribution complexity, it is difficult, if not impossible, to identify the specific source of fuel oils as fired.

1.1.3.1 Estimation of Pollutant Emissions

Table 1-8 presents emission factors[3] for fuel oil combustion without control equipment, as pounds of pollutant emitted per 1,000 gallons of oil burned. Estimation of emissions using this table requires the knowledge of sulfur and heat content of the oil used, as well as maximum heat input to the boiler. The oil usage in gallons per hour can be computed by dividing the maximum heat input (Btu/hr) by the heat content of the oil (Btu/gal). The uncontrolled emission rate (lb/hr) can then be obtained by the multiplication of oil usage (gal/hr) with the corresponding emission factor. The following example explains the estimation procedure more clearly.

Consider No. 6 fuel oil with 0.8 percent sulfur content by weight and the heat content of 180,000 Btu/gal.

For a power plant residual oil boiler with maximum heat input of 25×10^6 Btu/hr, the uncontrolled emission rates can be calculated as follows:

Fuel Usage = 25×10^6 Btu/hr ÷ 180,000 Btu/gal
= 1.389×10^2 gal/hr

A. Particulate Emissions

Emission Factor = 10(S) + 3 lb/1,000 gal
(from Table 1-8) = 10 (0.8) + 3 lb/1,000 gal
= 11 lb/1,000 gal

Emission Rate = (1.389×10^2 gal/hr) x (11 lb/1,000 gal)
= 1.528 lb/hr

B. Sulfur Dioxide Emissions

Emission Factor = 157(S) lb/1,000 gal
(From Table 1-8) = 157 (0.8) lb/1,000 gal
= 125.6 lb/1,000 gal

Emission Rate = (1.389×10^2 gal/hr) x (125.6 lb/1,000 gal)
= 17.446 lb/hr

Sulfur trioxide emissions can be calculated similarly.

C. Nitrogen Oxides Emissions

From Table 1-8, assuming the boiler not to be tangentially fired,

Emission factor = 105 lb/1,000 gal
Emission Rate = (1.389×10^2 gal/hr) x (105 lb/1,000 gal)
= 14.585 lb/hr

Emissions of carbon monoxide and hydrocarbons can be calculated in a similar fashion using Table 1-8.

1.1.3.2 Particulate Emissions from Oil Combustion

Residual oil typically has an ash content of 0.1 percent. When residual oil combustion is highly efficient, the resulting particulate emissions are con-

TABLE 1-8

EMISSION FACTORS FOR FUEL OIL COMBUSTION WITHOUT CONTROL EQUIPMENT[3]

Pollutant	Type of Boiler[a]		
	Power Plant Residual Oil	Industrial and Residual Oil	Commercial Distillate
	lb/1,000 gal	lb/1,000 gal	lb/1,000 gal
Particulate[b]	c	c	2
Sulfur Dioxide[d]	157S	157S	142S
Sulfur Trioxide	2S	2S	2S
Carbon Monoxide[e]	5	5	5
Hydrocarbons[f] (total, as CH_4)	1	1	1
Nitrogen Oxides (total, as NO_z)	105(50)[g]	60	h

[a]Power Plant = Greater than 250 million Btu/hr.
Industrial = Greater than 14 million Btu/hr, less than 250 million Btu/hr.
Commercial = Greater than 0.5 million Btu/hr, less than 15 million Btu/hr

[b]Filterable Particulate = That which is measured by the front half of an EPA = 5 train.

[c]No. 6 fuel oil = 10(S) + 3 lb/1000 gal.
No. 5 fuel oil = 10 lb/1000 gal.
No. 4 fuel oil = 7 lb/1000 gal.

[d]S = Sulfur content in the oil, percent by weight.

[e]Carbon monoxide emissions may increase by a factor of 10 to 100 if the boiler is improperly operated or not well maintained.

[f]Hydrocarbon emissions are generally negligible unless the boiler is improperly operated or not well maintained.

[g]Tangentially fired boilers = 50 lb/1,000 gal.
All other boilers = 105 lb/1,000 gal.
Reduced load = .5 to 1 percent NO_x reduction for every 1 percent reduction in boiler load.

[h]When nitrogen content of oil is less than 0.5 percent by weight = 400 $(N)^2$ lb/1,000 gal.
When nitrogen content is greater than 0.5 percent = 120 lb/1,000 gal.
N = Nitrogen content in the oil, percent by weight.

stituted almost entirely of inorganic ash, which occurs as oxides, chlorides, or sulfates. Frequently, however, this material consists of unburned carbonaceous solids that tend to be sticky and hygroscopic. The latter condition probably arises from the presence of calcination products and condensed sulfuric acid.

On a mass basis, the particulate emissions from an uncontrolled residual oil-fired boiler are of the same order as those from a highly controlled (>95 percent removal efficiency) coal-fired boiler. Stack tests have indicated that between 85 and 90 weight percent of the particles liberated by uncontrolled residual oil combustion are less than 1 micron in diameter, while usually less than 10 percent of those liberated by coal combustion are less than 1 micron. Because submicron particles are highly efficient light scatterers, uncontrolled residual oil-fired boilers often have high plume opacities. Due to the submicron size of the particles emitted, relatively efficient particulate collection equipment may be necessary in order to improve plume appearance. Besides reducing plume opacity, a high efficiency collector reduces the chance of acid smut discharge during soot blowing operations. Smuts are created by the formation and/or collection of sulfuric acid upon particle deposits that lie on furnace, duct, and stack liner surfaces. The sulfuric acid dew point of a stack gas increases with increasing sulfur trioxide and, to a lesser extent, increasing water vapor concentration. Generally speaking, the higher the sulfur content of a fuel oil, the more SO_3 is formed, and subsequently, the higher the sulfuric acid dew point.

1.1.3.3 Particulate Control in Oil-fired Plants[6]

1.1.3.3.1 Electrostatic Precipitators

When designing an electrostatic precipitator to control the particulate emissions from oil-fired boilers, the hygroscopicity, resistivity, and size distribution of the particulate must be considered. Due to hygroscopicity, solids buildup in hoppers, high tension electrodes, insulators, and collecting curtains is a problem. These solids, when allowed to contact cool surfaces, absorb moisture, become difficult to remove, and cause arcing and shorts. This problem can be remedied by keeping deposition surfaces hot and preventing the solids from "setting up." By locating the precipitator on the hot side of the air preheater, ash buildup on high tension wires and collection curtains in minimized. Buildup on

insulator bushings can be avoided by the use of hot air ventilation. Solids in hoppers can be kept mobile either by heating the wall surfaces or using a wet bottom flush-out system.

With respect to particulate emissions from oil-fired boilers, stack gas temperature and sulfur content of the oil affect the resistivity of the noncombustible portion of these solids; however, the balance of these solids is composed of highly conductive combustible carbonaceous solids. As a result of these carbonaceous solids, the resistivity of the particulate emissions is usually less than that for coal: 10^7 to 10^9 ohm-cm for oil versus 10^9 to 10^{13} ohm-cm for coal. In some cases, these solids are so conductive that they do not retain a charge and subsequently prevent the field from becoming saturated. Another problem is that these solids, upon deposition on collecting curtain surfaces, sometimes lose their charge to the curtain and become reentrained in the gas stream. The extremely fine size of the particles generated by oil combustion makes efficient collection by electrostatic precipitation difficult. Collection efficiency is improved through the employment of high voltage, large collection curtains, lower superficial gas velocity, and high retention times.

Corrosion due to high dew point, reentrainment of collected particulate matter, and fire hazards due to the combustible solids in ash hoppers are three problems that must also be considered. Reentrainment of particulate matter can be abated by optimizing rapping frequency and intensity. Problems associated with the hygroscopicity and combustibility of the collected particulate matter can be eliminated with a fly-ash injection system that keeps hot gas going through the hopper. The problem of solids combustion can also be combatted by using steam quenching devices that are activated by temperature. Corrosion problems can be eliminated by locating the precipitator on the hot side of the air preheater and by heating precipitator surfaces.

1.1.3.3.2 Fabric Filters

Although relatively undemonstrated (except for a single industrial installation), baghouses may have the potential to be effective in the reduction of particulate emissions from oil-fired boilers. The key to their success will depend mainly on the degree of operation and maintenance problems resulting from the difference in fly-ash properties (between coal and oil fly ash).

1.1.3.3.3 Wet Scrubbers

Although scrubbers can be used successfully to remove particulate emissions from oil-fired boilers, some potential problems must be faced and rectified. In order to realize efficient particulate removal, high pressure drops through the scrubber must be maintained. This results in high operational costs. Also, proper treatment of the scrubber discharge liquor presents a very real ecological and economic problem.

1.1.4 Emission Characterization

The combustion of a fuel for the generation of steam or hot water results in the emission of SO_x, NOx, CO, hydrocarbons, and particulate matter. The respective amounts of these emissions formed are dependent upon variables occurring within the combustion process. The inter relationships of these variables do not permit direct interpretation by current analytical methods. Therefore, most emission estimates are based upon factors compiled through extensive field testing and are related to the fuel type, the boiler type and size, and the method of firing. Although the use of emission factors based on the above parameters can yield an accurate first approximation of on-site boiler emissions, these factors do not reflect individual boiler operating practices or equipment conditions, both of which have a major influence on emission rates. A properly operated and maintained boiler requires less fuel to generate steam efficiently thereby reducing the amount of ash, nitrogen and sulfur entering the boiler and the amount of ash, hydrocarbons, nitrogen oxides and sulfur oxides exiting in the flue gas stream.

Table 1-9 contains typical operating data for stoker-fired boilers[7] and Table 1-10 contains data for oil-fired boilers.[7] In addition, actual emission data are provided for each type of boiler. The particulate and gaseous emissions listed are amounts measured after any air pollution control device.

1.1.5 Particulate Emissions

The particulate loadings in stack gases depend primarily on combustion efficiency and on the amount of ash not collected or deposited within the boiler. A boiler firing fuel with a high percentage of ash will have particulate emissions that depend more on the fuel ash

TABLE 1-9

OPERATING AND EMISSION CHARACTERISTICS OF COAL-FIRED STOKER BOILERS[7]
(FROM ACTUAL TESTS)

Type of Stoker	Control Device	Actual Heat Input Rate (million Btu/hr)	Actual Output (1,000 lb/hr)	Steam Rate	Stack Gas Flow Rate (Scfm)	Stack Gas Temp. (°F)	Particulate Emissions (lb/million Btu)	NO_x Emissions (lb/hr)	SO_2 Emissions (lb/hr)
Spreader	M/C	2.70	-		3,250	390	1.80	-	19.0
		4.50	-		3,340	485	1.40	-	28.0
		5.60	-		2,960	460	0.90	-	37.0
		5.70	4.00		4,450	454	0.98	-	-
		6.00	-		2,840	415	1.00	-	33.0
		7.60	-		2,870	425	0.80	-	42.0
		9.40	-		2,540	440	0.50	-	48.0
		10.00	-		3,070	460	0.60	-	58.0
		27.30	19.77		5,140	471	0.17	-	-
		34.20	-		19,324	152	0.87	-	31.1
		51.00	35.00		17,990	500	1.61	34.0	101.0
		72.00	50.00		24,430	610	1.21	62.0	168.0
		81.80	-		26,000	318	0.54	-	-
Underfeed	Cyclone	17.30	-		16,442	421	0.37	18.9	18.9
		32.40	-		13,243	451	0.32	18.9	30.3
Traveling Grate Stoker	M/C	143.00	75.0		31,630	495	37.18	-	470.00
		176.00	135.0			981	121.40	129.4	-

Dash indicates information not available.

M/C = Multiple cyclone.

TABLE 1-10

OPERATING AND EMISSION CHARACTERISTICS OF UNCONTROLLED OIL-FIRED BOILERS[7]

Type of Fuel	Actual Heat Input Rate (million Btu/hr)	Actual Steam Output Rate (1000 lb/hr)	Stack Gas Flow Rate (scfm)	Stack Gas Temp. (°F)	Particulate Emissions (lb/million Btu)
No. 4	2.42	-	869	-	0.09
No. 4	2.45	-	1,229	-	0.14
No. 4	2.65	-	848	-	0.04
No. 4	3.25	114.22	1,032	273	0.08
No. 6	27.00	-	9,789	-	0.28
No. 6	35.20	-	16,860	545	0.36
No. 6	198.00	-	58,481	-	0.32
No. 6	198.00	143.08	58,560	263	0.11

TABLE 1-11

FACTORS AFFECTING PARTICULATE EMISSIONS[8]

	Level of Particulate Emissions	
	High	Low
Fuel characteristics		
High ash content	X	-
High moisture content	X	-
Sizing		
High degree of atomization or pulverization	-	X
Boiler characteristics		
High combustion rate	X	-
Dry bottom ash collection	X	-
Wet bottom ash collection	-	X
Flue-gas recirculation	X	-
Method of firing		
Tangential	-	X
Horizontal	X	-
Spreader stoker	X	-
Boiler operation		
Improper oil pressure	X	-
Improper oil viscosity	X	-
High combustion air temperature	-	X
Nonuniform air flow (between register or air compartments)	X	-
High excess air	X	-
Low furnace temperature	X	-
Equipment condition:		
Worn burner (nozzles, sprayer plates, etc.)	X	-
Unclean or slagging boiler tube surfaces	X	-

content and the furnace ash collection or retention time than on combustion efficiency. In contrast, a boiler burning a low ash content fuel will have particulate emissions that depend more on the combustion efficiency the unit can maintain. Therefore, particulate emission estimates for boilers burning low ash content fuels will depend more on unit condition and operation. Factors[8] affecting particulate emissions are presented in Table 1-11.

1.1.6 Sulfur Oxide Emissions

During combustion, sulfur is oxidized in much the same way carbon is oxidized to CO_2. Therefore, almost all of the sulfur contained in the fuel will be oxidized to SO_2 or SO_3 in efficiently operated boilers. Field test data show that in efficiently operated boilers, approximately 98 percent of the fuel-bound sulfur will be oxidized to SO_2, 1 percent to SO_3, and the remaining 1 percent sulfur will be contained in the fuel ash. Boilers with low flue gas stack temperatures may produce lower levels of SO_2 emissions due to the formation of sulfuric acid.

1.1.7 SO_2 Control for Coal- and Oil-fired Plants[9]

The sulfur dioxide emission standards for industrial boilers may be met in a variety of ways. The use of low sulfur coal or oil is, of course, a distinct alternative. Assuming an allowable emission factor of 1.2 lb SO_2 per 10^6 Btu, the allowable sulfur content in a coal of heating value of 12,000 Btu/lb will be about 0.7 percent.

In a number of coal-producing regions in the continental United States, some coals are amenable to reduction of their sulfur contents by so-called physical techniques. These techniques utilize differences in physical properties between the coal and certain portions of the sulfur to separate this sulfur from the coal prior to transportation to the power plant.

When a high sulfur fuel is used in a power plant, it is necessary to control SO_2 emission by using a flue gas desulfurization system.

Although numerous processes have been proposed for flue gas desulfurization, only three are generally considered commercial for post combustion control of SO_2 from boilers and heating plants: (1) limestone slurry, (2) lime

scrubbing, and (3) double alkali. All the remaining are in various stages of development and beyond the scope of this report. Dry FGD is, however, briefly described.

1.1.7.1 Limestone Scrubbing

Flue gas is first cleaned of particulate matter in an electrostatic precipitator or equivalent. SO_2 is then reacted with $CaCO_3$ in a wet scrubber (most commonly a spray tower) to form $CaSO_3$ with typically 50 to 90 percent of the sulfite oxidized to sulfate. After thickening and vacuum filtration, the resulting 70 percent solids cake of $CaSO_3/CaSO_4$ must be transported by truck to a landfill area or a suitable disposal site. SO_2 removal efficiency typically ranges from 70 to 90 percent.

1.1.7.2 Lime Scrubbing

Lime Scrubbing uses lime slurry in a wet scrubber (most commonly a spray tower) to remove SO_2 from the flue gas. Both $CaSO_3$ and $CaSO_4$ are produced and are removed from the process in the form of a sludge stream that is thickened and filtered. SO_2 removal ranges from 70 to 90 percent depending on the SO_2 concentration in the flue gas.

1.1.7.3 Double Alkali Process

The double alkali process scrubs with a Na_2SO_3 buffer solution and then reacts the clear solution with lime or limestone to precipitate $CaSO_3$. The purpose of separating scrubbing from precipitation has been to eliminate scaling problems in the scrubber. Since the sludge removed from the system contains 4 to 5 percent sodium, soda ash must be added to the system to replace these losses. Removal efficiencies have been better than 90 percent for SO_2.

1.1.7.4 Dry FGD

Dry flue gas desulfurization involves contacting a sulfur-containing flue gas with an alkaline material and results in a dry waste product for disposal. This process includes (1) systems that use spray dryers for a contactor, with subsequent baghouse or electrostatic precipitator collection of waste products; (2) systems that involve dry injection of alkaline material into contact with flue gas, and subsequent baghouse or electrostatic precipitator collection; and (3) other varied dry systems which include concepts such as addition of alkaline

material to a fuel prior to combustion or contacting flue gas with a fixed bed of alkaline material. Of these systems, spray drying is currently the only one being developed on a commercial scale.

1.1.8 Nitrogen Oxides Emissions

The level of nitrogen oxides (NOx) present in stack gases depends upon many variables. Furnace heat release rate, temperature, and excess air are the major variables affecting NOx emission levels, but they are not the only ones. Therefore, while the emission factors presented later in this section may not totally reflect on-site conditions, they are useful in determining if a NOx emission problem may be present. Factors that influence NOx formation[8] are shown in Table 1-12.

1.1.9 NOx Control for Coal- and Oil-fired Plants

Nitrogen oxides control is generally achieved by improved combustion control and combustion modifications such as:

- low stoichiometric combustion, selective burner service and two staged combustion
- flue gas recirculation
- reduced combustion air preheat temperature
- low excess air
- reduced load operation.

1.1.10 Carbon Monoxide and Hydrocarbons Emissions

Carbon monoxide is produced in a combustion process because of incomplete combustion of carbon. A small portion of the hydrocarbons remains unburned and is discharged to the atmosphere with the other combustion products. Both of these pollutants carry useful heat. Hence, the boilers are operated in such a way that their emission remains at the minimum possible level. No control equipment is, therefore, necessary to remove them.

TABLE 1-12

FACTORS AFFECTING NO_x EMISSIONS[8]

Increasing Variable	Effect on NO_x Emissions: Increase	Effect on NO_x Emissions: Decrease
Heat release rate	X	-
Heat quenching rate (water cooled furnaces)	-	X
Boiler capacity	X	-
Boiler load	X	-
Excess air	X	-
Furnace temperature	X	-
Combustion air temperature	X	-
Fuel nitrogen content	X	-
Fuel oxygen content	X	-
Fuel moisture content	-	X
Distance between burners	-	X
No. of burners per unit	-	X
Fuel/air mixing (turbulence)	X	-
Time available for particle burnout (tangentially fired boilers)	-	X

1.2 SOLID WASTE INCINERATION[10]

1.2.1 General Waste Incinerators

1.2.1.1 Process Characterization

The most common type of general waste incinerators consists of a refractory-lined chamber with a grate upon which refuse is burned. In some recent heat recovery incinerators, water tubes line the furnace walls to generate hot water and/or steam. Combustion products are formed by heating and burning of refuse on the grate. In most cases insufficient underfire air is used to prevent unburned particles and fly ash from blowing out of the combustion chamber. In order to enable complete combustion, additional overfire air is admitted above the burning waste, promoting complete combustion.

In multiple-chamber incinerators, gases from the primary chamber flow to a small secondary mixing chamber where more air is admitted, and more complete oxidation occurs. As much as 300 percent excess air may be supplied in order to promote oxidation of combustibles. Auxiliary burners are sometimes installed in the mixing chamber to increase the combustion temperature. Many small-size incinerators are single-chamber units in which gases are vented from the primary combustion chamber directly into the exhaust stack. Single-chamber incinerators of this type do not meet modern air pollution codes, unless a very efficient afterburner is installed in an insulated chamber and the gas velocity is slow enough to ensure adequate residence time.

The combustion process in multiple-chamber incinerators proceeds in two stages--primary or solid fuel combustion in the ignition chamber, followed by secondary or gaseous phase combustion. The secondary combustion zone is composed of two parts, a downdraft or mixing chamber and an up-pass expansion or combustion chamber.

The multiple-chamber incineration process begins in the ignition chamber and includes the drying, ignition, and combustion of the solid refuse. As the burning proceeds, the moisture and volatile components of the fuel are vaporized and partially oxidized in passing from the ignition chamber through the flame port connecting the ignition chamber with the mixing chamber. From the flame port, the volatile components of the refuse and the products of combustion flow down through the mixing chamber,

TABLE 1-13

OPERATING DATA ON MULTIPLE-CHAMBER INCINERATORS[11]

	Multiple-Chamber Incinerator							
Capacity, lb/hr	50	50	350	750	1,000	1,000	2500	6000
Charging rate, lb/batch	2-4	2-4	30	20	75	Mech.[a]	400	650
Refuse composition, %/weight								
Paper	100	69	0	71	83	0	100	65
Garbage	0	31	0	17	17	0	0	0
Wood	0	0	100	12	0	100	0	35
Auxiliary fuel, standard cubic ft/hr-gas								
Primary chamber	0	165	0	0	0	0	0	0
Mixing chamber	0	165	0	1,125	2,850	0	b	0
Combustion air, % of total								
Overfire air	85	45	55	79	50	20	60	70
Underfire air	15	10	3	7	20	4	3	10
Secondary air (mixing chamber)	0	45	42	14	30	76	37	20
Orsat gas analysis, %/volume								
Carbon dioxide	4.8	6.4	8.4	6.0	7.4	5.8	2.2	6.3
Oxygen	13.8	6.3	11.2	12.6	9.9	14.7	18.3	9.4
Carbon monoxide	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Nitrogen	81.4	87.3	80.4	81.4	82.7	79.5	79.5	84.3

[a] Mech. = mechanized feed.

[b] Oil at 2.5 gallons per hour.

TABLE 1-14
CLASSIFICATION OF WASTES

Type	Description	Contents
0	Trash, a mixture of highly combustible waste such as paper, cardboard cartons, wood boxes, and combustible floor sweepings, from commercial and industrial activities. The mixtures contain up to 10% by weight of plastic bags, coated paper, laminated paper, treated corrugated cardboard, oily rags,and plastic or rubber scraps.	10% moisture, 5% incombustible solids and a heating value of 8,500 Btu per pound as fired.
1	Rubbish, a mixture of combustible waste such as paper, cardboard cartons, wood scrap, foliage and combustible floor sweepings, from domestic, commercial and industrial activities. The mixture contains up to 20% by weight of restaurant or cafeteria waste, but contains little or no treated papers, plastic,or rubber wastes.	25% moisture, 10% incombustible solids, and a heating value of 6,5000 Btu per pound as fired.
2	Refuse, consisting of an approximately even mixture of rubber and garbage by weight.	50% moisture, 7% incombustible solids, and a heating value of 4,300 Btu per pound as fired.
3	Garbage, consisting of animal and vegetable wastes from restaurants, cafeterias, hotels, hospitals, markets, and like.	Up to 70% moisture, up to 5% incombustible solids, and a heating value of 2,500 Btu per pound as fired.
4	Human and animal remains, consisting of carcasses, organs, and solid organic wastes from hospitals, laboratories, and like installations.	Up to 85% moisture, up to 5% incombustible solids, and a heating value of 1,000 Btu pound as fired.
5	Gaseous, liquid,or semi-liquid by-product waste, such as tar, paint, solvent sludge and fumes,from industrial operations.	Btu values must be determined by the individual materials to be destroyed.
6	Solid by-product waste, such as rubber, plastic,and wood waste, from industrial operations.	Btu values determined by the individual materials to be destroyed.

TABLE 1-15

CLASSIFICATION OF INCINERATORS

Class	Description
I	Portable, packaged, completely assembled, direct-fed incinerators, having not over 5 ft^3 storage capacity, or 25 lb per hr burning rate, suitable for Type 2 waste.
IA	Portable, packaged or job-assembed, direct-fed incinerators 5 to 15 ft^3 primary chamber volume, or a burning rate of 25 lb per hr up to, but not including 100 lb per hr of Type 0, Type 1, or Type 2 waste, or a burning rate of 25 lb per hr up to, but not including 75 lb per hr of Type 3 waste.
II	Flue-fed, single chamber incinerators with more than 2 ft^2 burning area, for Type 2 waste. This type of incinerator is served by one vertical flue functioning both as a chute for charging waste and for carrying the products of combustion to atmosphere. This type of incinerator has been installed in apartment or multiple dwellings.
IIA	Chute-fed multiple chamber incinerators for apartment buildings, with more than 2 ft^2 burning area, suitable for Type 1 or Type 2 waste. (Not recommended for industrial installations.) This type of incinerator is served by a vertical chute for changing waste from two or more floors above the incinerator and a separate flue for carrying the products of combustion to atmosphere.
III	Direct-fed incinerators with a burning rate of 100 lb per hr and over, suitable for Type 0, Type 1, or Type 2 waste.
IV	Direct-fed incinerators with a burning rate of 75 lb per hr or over, suitable for Type 3 waste.
V	Municipal incinerators suitable for Type 0, Type 1, Type 2, or Type 3 wastes, or a combination of all four wastes, and are rated in tons per hr or tons per 24 hr.
VI	Crematory and pathological incinerators, suitable for Type 4 waste.
VII	Incinerators designed for specific by-product wastes, Type 5 or Type 6.

TABLE 1-16

COMPARISON BETWEEN AMOUNTS OF EMISSIONS FROM SINGLE-NUMBER AND MULTIPLE-CHAMBER INCINERATORS [11]

Emission	Multiple Chamber	Single Chamber
Particulate matter, grains/scf at 12% CO_2	0.11	0.9
Volatile matter, grains/scf at 12% CO_2	0.07	0.5
Total, grains/scf at 12% CO_2	0.19	1.4
Total, lb/ton refuse burned	3.50	23.8
Carbon monoxide, lb/ton of refuse burned	2.90	197 to 991
Ammonia, lb/ton of refuse burned	0	0.9 to 4
Organic acid (acetic), lb/ton of refuse burned	0.22	< 3
Aldehydes (formaldehyde), lb/ton of refuse burned	0.22	5 to 64
Nitrogen oxides, lb/ton of refuse burned	2.50	< 0.1
Hydrocarbons (hexane), lb/ton of refuse burned	< 1	--

TABLE 1-17

EMISSIONS FROM MULTIPLE-CHAMBER INCINERATORS [12]

	Multiple-Changer Incinerator							
Capacity, lb/hr	50	50	350	750	1,000	1,000	2,500	6,000
Maximum opacity of stack gases, %	10	0	0	45	10	0	15	0
Smoking time, minutes	1	0	0	1	2.5	0	9	0
Particulates, number total, number grains/scf								
At stack conditions	0.099	0.058	0.024	0.075	0.083	0.052	0.019	0.092
Particulates, filter collection only, grains/ scf	0.068	0.039	0.023	0.055	0.045	0.025	0.009	0.066
Total corrected to 12% carbon dioxide	0.270	0.300	0.038	0.205	0.248	0.116	0.113	0.200
Filter collection only, number corrected to 12% carbon dioxide	0.185	0.182	0.033	0.130	0.119	0.053	0.057	0.126
Total minus filter collection, grains/ scf								
Corrected to 12% carbon dioxide	0.085	0.118	0.005	0.075	0.129	0.063	0.056	0.074
Gaseous emissions, number grains/ scf								
Carbon monoxide	-	-	0.0	-	-	0.0	-	-
Nitrogen oxides	-	-	0.032	0.0001	-	0.028	-	-
Aldehydes	-	-	0.002	0.0001	-	0.002	-	-
Organic acids	-	-	0.021	-	-	0.009	-	-
Sulfur dioxide or trioxide	-	-	-	-	-	-	-	-

where secondary air is introduced. The combination of adequate temperature and additional air, augmented by mixing chamber or secondary burners as necessary, assists in initiating the second stage of the combustion process. Turbulent mixing, resulting from the restricted flow areas and abrupt changes in flow direction, furthers the gaseous-phase reaction. In passing through the curtain wall port from the mixing chamber to the final combustion chamber, the gases undergo additional changes in direction, accompanied by expansion and final oxidation of combustible components.

Fly ash and other particulate matter are collected in the combustion chamber by impingement and settling. Depending upon the specific circumstances, a wet scrubber and/or a fabric filter or an electrostatic precipitator may be required for air pollution control. The gases finally discharge through a stack.

1.2.1.2 Incinerator and Waste Classifications

Operating conditions[11] for several multiple-chamber incinerators, ranging in capacity from 50 to 6,000 lb/hr, are shown in Table 1-13.

As a guide, mixtures of waste most commonly encountered have been classified into types of waste together with the Btu values and moisture contents of the mixtures. A concentration of one specific waste in the mixture may change the Btu value and/or the moisture content of the mixture. A concentration of more than 10 percent by weight of catalogues, magazines, or packaged paper will change the density of the mixture and affect burning rates. Similarly, incinerators have been classified by their capacities and by the type of wastes they are capable of incinerating.

For information, the standards of the Incinerator Institute of America are given for the classification of wastes and incinerators in Tables 1-14 and 1-15, respectively.

1.2.1.3 Emission Characterization

Air pollution emissions from multiple-chamber and single-chamber incinerators[11] are compared in Table 1-16. Stack emission data for several typical multiple-chamber incinerators,[12] ranging in capacity from 50 to 6,000 lb/hr, are shown in Table 1-17.

Operating conditions, refuse composition, and basic incinerator design have a pronounced effect on emissions. Air may be introduced from beneath the chamber, from the side, or from the top of the combustion area. As underfire air is increased, an increase in fly-ash emissions occurs.

Erratic refuse charging disrupts the combustion bed and cause a subsequent release of large quantities of particulates. Large quantities of uncombusted particulate matter and carbon monoxide are also emitted for an extended period after charging of batch-fed units because of interruptions in the combustion process. In continuously fed units, furnace particulate emissions are strongly dependent upon grate type. The use of rotary kiln and reciprocating grates results in higher particulate emissions than the use of vibrating or traveling grates.

Emissions of oxides of sulfur are dependent on the sulfur content of the refuse. Carbon monoxide and unburned hydrocarbon emissions may be significant and are caused by poor combustion resulting from improper incinerator design or operating conditions. Nitrogen oxide emissions increase with an increase in the temperature of the combustion zone, an increase in the residence time in the combustion zone before quenching, and an increase in the excess air rates to the point where dilution cooling overcomes the effect of increased oxygen concentrations.

1.2.1.4 Emission Factors

Table 1-18 shows emission factors for general waste incinerators.[3]

Hydrochloric acid emissions were found to be approximately 1.0 to 1.8 lb/ton of feed. The level can be sharply increased in areas where large quantities of plastics are consumed. Methane levels may range from 0.004 to 0.4 lb/ton of feed.

Single-chamber incinerators have been found to have particulate emissions from 14 to 35 lb/ton of material burned. By contrast, the particulate discharges from well-designed multiple-chamber incinerators average 4.5 lb/ton of refuse burned.

1.2.2 Pathological-Waste Incinerators

1.2.2.1 Process Characterization

Pathological waste includes all, or parts of, organs, bones, muscles, other tissues, and organic wastes of human or animal origin.

A pathological-waste incinerator is a multiple-chamber incinerator with the following modifications to allow for the presence of a relatively high percentage of moisture:

TABLE 1-18

EMISSION FACTORS FOR GENERAL WASTE INCINERATORS WITHOUT CONTROLS[3,a]

Incinerator	Particulates		Sulfur Oxides[b]		Carbon Monoxide		Organics[c]		Nitrogen Oxides	
	lb/ton,	kg/MT	lb/ton,	kg/MT	lb/ton,	kg/MT	lb/ton,	kg,MT	lb/ton,	kg/MT
Municipal chamber,										
Multiple uncontrolled	30	15	2.5	1.25	35	17.5	1.5	0.75	3	1.5
With settling chamber and water spray system[e]	14	7	2.5	1.25	35	17.5	1.5	0.75	3	1.5
Industrial/commercial										
Multiple chamber	7	3.5	2.5[f]	1.25	10	5	3	1.5	3	1.5
Single chamber	15	7.5	2.5[f]	1.25	20	10	15	7.5	2	1
Controlled air	1.4	0.7	1.5	0.75	Neg	Neg	Neg	Neg	10	5
Flue-fed single chamber	30	15	0.5	0.25	20	10	15	7.5	3	1.5
Flue-fed (modified)[g]	6	3	0.5	0.25	10	5	3	1.5	10	5
Domestic single chamber										
Without primary burner	35	17.5	0.5	0.25	300	150	100	50	1	0.5
With primary burner	7	3.5	0.5	0.25	Neg	Neg	2	1	2	1

Neg means negligible

[a] Average factors given based on EPA procedures for incinerator stack testing.

[b] Expressed as sulfur dioxide.

[c] Expressed as methane.

[d] Expressed as nitrogen dioxide.

[e] Most municipal incinerators are equipped with at least this much control

[f] Based on municipal incinerator data.

[g] With afterburners and draft controls

1. The use of a solid hearth instead of grates.
2. The provision for heating the hearth by passing the combustion products from the mixing chamber through a chamber beneath the hearth before they exit to the combustion chamber. (Unless the incinerator operates continuously for several hours, little if any advantage is gained by an underhearth chamber.)
3. A side charging door, which is necessary for frequent charging of large individual components of pathological waste.

The combustion process is the same as in general-refuse incinerators. The reader should, therefore, refer to that section for a process description. Additional detailed description and diagrams can be obtained from Reference 11.

1.2.2.2 Operating Characteristics

Operating conditions for several multiple-chamber incinerators[12] are shown in Table 1-17. Charging rates typically range from 19.2 to 160 lb/hr of feed material.

1.2.2.3 Emission Characterization

Visual emissions of fly ash are not evident from pathological-waste incinerators. Air contaminants, as solid, liquid, and gaseous emissions,[11] are given in Table 1-19. The stack effluent from a well-designed incinerator will not be highly objectionable from the standpoint of odors when freshly killed or frozen animals are being cremated. However, cremation of decayed animal matter will produce objectionable odors that will not be entirely eliminated by the incinerator.

1.2.2.4 Emission Factors

Table 1-20 shows emission factors[3] for pathological-waste incinerators without controls.

1.2.2.5 Medical/Infectious Waste Incinerators

1.2.2.5.1 Process Characterization and Operating Characteristics

Multiple-chamber incinerators are generally used for disposal of medical or infectious wastes. The wastes that may be found in the main stream of the solid waste system include garbage, rubbish, ashes, special wastes, and reusables, such as linens and plastic packaging items. A description of the process can be found in the General Waste Incinerator section.

TABLE 1-19

EMISSIONS FROM PATHOLOGICAL-WASTE INCINERATORS WITHOUT GAS WASHERS[11]

Type of Waste	Human	Tissue	Animals					
Batch destruction rate to dry bone and ash, lb/hr	19.2	64	62	35	99	137	149	160
Particulates,								
gr/scf	0.014	0.017	0.032	0.015	0.0936	0.013	0.024	0.0202
gr/scf at 12% CO_2 (CO_2 from refuse only)	0.240	0.400	0.183	0.106	0.295	0.260	0.240	0.135
Organic acids,								
gr/scf	0.006	0.0003	0.010	NA	0.013	0.0033	0.0018	0.0002
lb/hr	0.010	0.003	0.034	NA	0.050	0.075	0.012	0.002
lb/ton	1.04	0.093	1.10	NA	1.01	1.10	0.161	0.025
Aldehydes,								
gr/scf	NA	0.008	0.013	0.004	0.006	0.0032	0.012	0.010
lb/hr	NA	0.076	0.041	0.014	0.020	0.072	0.082	0.12
lb/ton	NA	2.37	1.32	0.80	0.40	1.05	1.10	1.50
Nitrogen oxides,								
ppm	42.7	35	134	111	131	60	165	102
lb/hr	0.085	0.29	0.37	0.29	0.099	1.2	0.94	1.1
lb/ton	8.86	9.05	12.0	16.6	2.00	17.5	12.6	13.7
Opacity, %	0	0	0	0	0	0	0	0

TABLE 1-19 (Cont.)

Type of Waste	Human Tissue		Animal					
Batch destruction rate to dry bone and ash, lb/hr	19.2	64	62	35	99	137	149	160
Auxiliary fuel								
Primary, scfm	190	700	530	600	640	800	1,020	1,800
Mixing, acfm	185	230	170	300	260	600	480	600
Gas flow, scfm	260	1,150	380	370	450	2,640	780	1,400
Gas temperature, °F	410	307	590	950	800	346	1,020	910
Stack gases, %								
CO_2	3.4	2.1	5.6	6.3	7.6	1.6	4.9	5.0
O_2	12.5	16.5	9.8	7.7	4.8	17.7	10.8	10.8
CO	0.0009	0.0	0.004	0.0	0.02	0.0	0.0	0.0
N_2	74.0	74.8	71.5	71.9	67.2	75.5	71.2	73.1
H_2O	10.1	6.6	13.1	14.1	20.4	5.2	13.1	11.1

TABLE 1-20

EMISSION FACTORS FOR PATHOLOGICAL-WASTE INCINERATORS WITHOUT GAS WASHERS[3,a]

Incinerator Type	Particulates		Sulfur Oxides[b]		Carbon Monoxide		Organics		Nitrogen Oxides[d]	
	lb/ton	kg/MT	lb/ton	kg/MT	lb/ton	kg/MT	lb/ton	kg/MT	lb/ton	kg/MT
Pathological	8	4	Neg	Neg	Neg	Neg	Neg	Neg	3	1.5

[a] Average factors given based on EPA procedures for incinerator stack testing.

[b] Expressed as sulfur dioxide.

[c] Expressed as methane.

[d] Expressed as nitrogen dioxide.

Typical operating data[7] are shown in Table 1-21. Information on the types and amounts of waste produced at medical facilities[13] is given in Table 1-22.

1.2.2.5.2 Emission Characterization

Emisson data from medical waste incinerators are not generally available. Some emission data are presented in Table 1-21, but US EPA emission factors have not been developed. Problems result from the incineration of chlorinated plastics in the form of hydrochloric acid gases; the emission rates depend on the content of the refuse as well as the charging rates.

1.2.3 Explosive Waste Incinerators[14]

1.2.3.1 Process Characterization

Explosive-waste incinerators are specialized units designed to dispose hazardous material. Three explosive waste disposal systems currently considered for use at various Army installations are:

1. simplified incineration technique for pollution abatement (SITPA II) system
2. fluidized bed incinerator
3. rotary kiln.

The waste materials include propellants, explosives, pyrotechnics (PEP), and munitions (e.g., grenades).

1.2.3.1.1 SITPA II System

The SITPA II system consists of an APE 1236 deactivation furnace, a cyclone separator, flame arrestor, and a baghouse. Munitions are fed to the unit by means of a conveyor belt. Upon entering the furnace, the munitions move through the retort toward the flame by means of spiral flights, which are an integral part of the retort casting. As the munitions approach the flame and become heated, they either detonate or burn freely, depending upon the munition configuration and characteristics. High order detonations are contained by the retort wall. The spiral flights provide physical separation of munitions or groups of munitions, discouraging propagation of detonations and defeating fragments generated by detonation. Metal parts and components of the munitions are then discharged from the furnace and may be recycled.

1.2.3.1.2 Fluidized Bed Incinerator

The fluidized bed incinerator consists of a vertical metal cylinder containing a bed material such as alumina

TABLE 1-21

OPERATING AND EMISSION DATA FROM MEDICAL INCINERATORS[7]

Incinerator Type	Type of Waste	Actual Charging Rate (lb/hr)	Particulate Emissions (gr/SDCF at 12% CO_2)	Stack Volumetric Flow Rate (Scfm)	Stack Temp. (°F)	Stack O_2 (%)	Stack CO_2 (%)	Excess Air (%)
Combustible Model 400	General medical	281	0.177	551	1,170	8.4	11.00	-
Single chamber with after burner	General refuse	287	0.164	471	1,375	8.4	10.90	67
Multiple chamber	Type 0-2	128	0.32	347	1,309	11.0	8.50	113
Pathological	General medical	43	0.15	1,683	167	-	1.15	-
Pathological	General medical	45	0.23	1,671	193	-	1.04	-

TABLE 1-22

BREAKDOWN OF DAILY DISPOSABLE WASTE PRODUCTION (LB/DAY) BY TYPE OF WASTE FROM MEDICAL FACILITIES[13]

Size of Medical Facility (No. of beds)	Sharp Needles, Etc.	Patho-logical Surgical	Rubbish	Non-combustibles	Garbage (non-grindable)[a]	Animal Carcasses	Food Waste (grindable)	Total Disposable Waste
3,000	75	1,000	16,200	1,500	1,800	25	2,000	23,200
428	3	TR[b]	540	75	150	-	330	1,098
715	22	156	6,569	465	660	220	950	9,062
1,188	40	4	2,760	725	875	20	1,100	5,544
259	8	115	717	80	160	10	210	1,350
725	20	6	1,722	250	475	23	1,860	4,376
232	5	TR	362	80	110	-	150	732

[a] Predominantly garbage mixed with substantial quantities of paper, plastics, metal, etc.

[b] TR = Totally reused.

(Al_2O_3). Heated air is forced up through the bed, producing a lifting of the bed material. Upon reaching steady state conditions, the bed is expanded vertically and the material becomes fluidized. Along with the flow of air and heat, slurried explosives are continuously injected into the bed. In the fluidized bed concept, most of the combustion takes place in a reducing (starved-air) atmosphere. Only about 70 percent of the necessary stoichiometric air is injected through the bed; an additional 50 percent is introduced near the top of the bed. This starved-air primary combustion condition is intended to reduce NOx formation.

1.2.3.1.3 Rotary Kiln

An alternate deactivation furnace is the rotary kiln facility developed by the Radford Army Ammunition Plant, Virginia. The PEP-water slurry is prepared by grinding bulk explosives or propellants under water to a particle size of 0.1 inch or less. It is then discharged into the constantly rotating kiln. After the water evaporates from the slurry, the explosive or propellant ignites and burns.

Diagrams and additional detailed descriptions of these incinerators can be found in "PEP and PEP-Contaminated Wastes Disposal Technology," Goldberg, Robert; Wood, James; U.S. Army Environmental Hygiene Agency, APCA Paper No. 79-5.3, June, 1979.

1.2.3.2 Operating Characteristics

1.2.3.2.1 SITPA II System

A breakdown of various munitions[14] demilitarized with the SITPA II are shown in Table 1-23. The operational characteristics[14] of the SITPA II are shown in Table 1-24.

1.2.3.2.2 Fluidizied Bed Incinerator

Operating characteristics for the Picatinny fluidized bed incinerator[14] are shown in Table 1-25. Classification of this type of unit as an incinerator or as a process is largely at the discretion of the state or local regulatory authorities.

1.2.3.2.3 Rotary Kiln

The required feed mixture ratio of approximately 3 parts water and 1 part PEP is maintained in the lower section of the vessel. The normal feed rate is 250 lb/hr of PEP and 750 lb/hr of water for a total slurry feed of 1,000 lb/hr. The 8-feet long by 5-feet diameter kiln

TABLE 1-23

EMISSIONS FROM THE SITPA II SYSTEM [14]

Item	50 Caliber Cartridge	Assembled Booster	Base Detonating Fuse	50 Caliber Cartridge
Abbreviation	API M8	M 21A4	M66A2	M17
Disposal Rate (item/min)	120	7	10	160
BEFORE AP CONTROLS[a]				
Particulate Emissions (lb/hr)	81.9-357.1	12.8-24.0	-	-
AFTER AP CONTROLS				
Particulate Emissions (lb/hr)	0.3-0.5	0.2-0.3	0.1-0.2	0.2
Visible Emissions (max %)	40	0	0	40
Visible Emissions (avg %)	5-10	0	0	5-10

[a] Air pollution (AP) controls are a cyclone followed by a baghouse.

TABLE 1-24

OPERATING DATA FOR THE SITPA II SYSTEM[14]

Item	50 Caliber Cartridge	Assembled Booster	Base Detonating Fuse	50 Caliber Cartridge
Abbreviation	API M8	M21A4	M66A2	M17
lb/item[a]	0.24	0.93	2.11	0.17
Disposal Rate (item/min)	120	7	10	10
BEFORE AP CONTROLS[b]				
Temperature (°F)	194-291	288-298	-	-
Oxygen (%)	19.8	20.5-20.6	-	-
AFTER AP CONTROLS				
Temperature (°F)	169-207	230-235	110-182	228-232
Flow Rate (Scfm)	1,880-2,200	1,760-2,290	2,590-4,490	2,370-2,480

[a] Weight includes combustibles and noncombustibles.

[b] Air pollution (AP) controls are a cyclone followed by a baghouse.

TABLE 1-25

OPERATING DATA FROM A FLUIDIZED BED INCINERATOR[14]

	Fluidized Bed Incinerator								
Slurry Composition	100-H_2O	25-CB[a]	22-CB	20-CB	15-CB	5-CB	25-TNT	15-TNT	5-T
Feed Rate (lb/hr)	1,160	1,248-1,816	1,444	1,625	1,180-1,600	1,160-1,540	1,824	1,600	1,5
Fuel Usage (gal/min)	1.05-1.15	0.70-1.08	0.85-1.12	0.80	0.80-0.98	1.07-1.15	1.19	0.93	0.8
Grid Temperature (°F)	825-883	895-1,300	1,180-1,425	857	800-964	913-950	935	870	927
Bed Temperature (°F)	1,150-1,630	1,525-1,700	1,700-1,950	1,510	1,617-1,686	1,713-1,775	1,580	1,663	1,5
Cyclone Temperature (°F)	1,667-1,683	1,533-1,700	--	1,442	1,600-1,720	1,725-1,738	1,643	1,728	1,7
Primary Input Air Flow (scfm)	900-1,000	500-1,800	700-1,200	600	500-800	500-800	500	500	5
Secondary Input Air Flow (scfm)	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,0

[a]CB = composition B explosive (40% TNT, 60% RDX).

cylinder is not equipped with flights. It is heated to 1,600°F by a burner using No. 2 fuel oil fed in from the end opposite the PEP feed. The rotary cylinder is inclined slightly toward the burner or discharge end to allow ash to move toward the cleanout door located beneath the burner. The effluent stream is passed through an oil-fired afterburner, a precooler, and a marble bed wet-scrubber.

1.2.3.3 Emission Characterization

1.2.3.3.1 SITPA II System

Emission data for the SITPA II are shown in Table 1-23. As can be seen, visible emissions are less than 10 percent opacity. Also, particulate emissions can be controlled to more than 97 percent efficiency[14].

1.2.3.3.2 Fluidized Bed Incinerator

Emission data for this incinerator[14] are shown in Table 1-26. Particulate emissions are calculated in terms of grains per DSCF corrected to 12 percent CO_2 without the contribution of auxiliary fuel (typical incinerator emissions evaluation), and in terms of pounds per hour which is typical of many process standards. When compared against the New Jersey incinerator standard of 0.1 grain DSCF corrected in 12 percent CO_2 without the contribution of auxiliary fuel, the fluidized bed incinerator unit exceeded the standard in all cases except when burning a slurry composed of 25 percent composition B. With respect to visible emissions, the fluidized bed unit is in compliance with a 20 percent opacity standard except when burning a slurry of 22 percent and 25 percent composition B.

Total NOx emissions range from about 150 to 450 ppm during various test runs.

1.2.3.3.3 Rotary Kiln

Emission data are available from the 250 lb/hr system at the Radford Army Ammunition plant.[7] Particulate emissions ranged from 0.012 to 0.031 grain/SCF corrected to 12 percent CO_2. Gaseous emissions ranges in ppm were hydrocarbons, 10-80, H_2S, 0-80, NO, 11-200, NO_2, 0-180, SO_2, 0-70. The variations are primarily due to different explosive feeds.

1.2.3.4 Emission Factors

Emission factors for explosive-waste incinerators have not been developed. Therefore, it will be necessary to consider each source individually. Also, source testing must be performed to obtain information for compliance and/or control equipment design.

TABLE 1-26

EMISSIONS FROM A FLUIDIZED BED INCINERATOR[14]

	Fluidized Bed Incinerator								
Slurry Composition (% type)	100-H_2O	25-CB[a]	22-CB	20-CB	15-CB	5-CB	25-TNT	15-TNT	5-
Feed Rate (lb/hr)	1,160	1,248-1,816	1,444	1,625	1,180-1,600	1,160-1,540	1,824	1,600	1,
Particulate Emissions (grains/scf)[b]	0.42-1.03	1.06-1.59	0.35-0.41	0.75	0.50-1.54	0.44-0.49	1.12	0.63	0.
Particulate Emissions (lb/hr)	5.90-8.24	0.19-8.62	1.80-3.35	6.75	3.85-6.84	4.29-5.53	6.10	4.94	4.
Visible Emissions (avg. %)	50-100	5-45	10-30	5	10	5	10	10	5
NO_x Concentration (ppm)		336-402	192-530	436	187-215	173	336	295	28
SO_2 Concentration (ppm)	12	23-33	--	--	--	--	--	--	--

[a] CB = composition B explosive (40% TNT, 60% RDX).

[b] Corrected to 12% CO_2 without the contribution of auxiliary fuel.

1.2.3.5 Control Systems--Incinerators

Although air pollution control systems incorporating water scrubbing will remove some water soluble gases, air pollution control equipment now in operation on incinerators has been designed primarily to remove particulate matter. The remainder of this section deals solely with the role of such equipment in removing particulate matter.

1.2.3.5.1 Wet Scrubbers

About 20 percent of the incinerator plants built since 1960 have been equipped with wet scrubbers. Most scrubbers in municipal incinerators operate at pressure drops of 5 to 7 inches of water with efficiencies about 90 percent.[15] High energy scrubbers, such as the venturi scrubber, may operate at over 99 percent efficiency with pressure drops in the range of 20 to 30 inches of water.[15] Water requirements range from about 5 to 15 gal/1,000 acfm of gas.

The main disadvantage of scrubbers is the corrosion from the absorption of materials in the gas stream, frequently causing the effluent scrubber water to become highly acidic. This problem requires the use of corrosion-resistant materials or the neutralization of the acid by alkali additives. Experience has shown that treatment and recirculation of scrubber water is difficult to accomplish. Some scrubber water is lost by evaporation into the effluent gas stream. The hot, saturated gas stream from the stack contacts air at ambient temperature and results in an unsightly vapor plume. Condensation of the vapor can cause settling of droplets and particles on local structures and vehicles, with damaging results.

1.2.3.5.2 Cyclones

The cyclone configurations that have been used for incinerators are, in order of decreasing particulate removal efficiency:

a. The multiple-cyclone system with numerous small-diameter (less than 12 inches) cyclone units installed in a tube sheet.
b. The multiple-cyclone system of larger diameter (over 24 inches) units installed in clusters, with flue gas manifold to the inlets of the individual cyclones, and the outlets manifolded to a common duct.
c. Single or double cyclone units of very large diameter (over 3 ft) with a single or split flue duct at the inlet and outlet.

Under ideal operating conditions, the smaller diameter cyclone system can attain 80 percent collection efficiency on incinerator ash.[15] However, plugging of the tubes by adherent fly-ash, or by particles wetted by an upstream gas-cooling system, will greatly impair their collection efficiency. The larger diameter cyclone systems are usually free of plugging, and can achieve efficiencies up to about 70 percent.[15]

1.2.3.5.3 Fabric Filters

In the past, incinerator system designers have avoided fabric filter systems for incinerators because of the assumption that initial costs and bag replacement costs would be prohibitive; greater control of combustion would be necessary to prevent formation of soot and tarry condensates, and closer gas temperature control would be necessary to prevent thermal destruction of the fabric, or condensation on the fabric, with the possibility of fabric plugging and subsequent damage to the collector. However, advancements in thermal control and fabric materials are resulting in further evaluation of fabric filters. A number of systems have been installed and have demonstrated sustained, satisfactory operation. Fabric filters are used extensively on SITPA II.

1.2.3.5.4 Electrostatic Precipitators

Electrostatic precipitators have been used for a number of years on incinerators with excellent results, particularly in European incinerator plants that recover heat. Efficiencies in the range of 96 to 99.6 percent have been achieved at gas pressure drops below 0.5 inches of water. Electrical power requirements are in the range of 200 to 400 watts per 1,000 acfm of gas treated. Inlet temperatures usually range from 350° to 700°F. A number of municipal incinerators in the United States are either equipped with electrostatic precipitators, or will be so equipped in the future.

1.3 AMMUNITION MANUFACTURE

1.3.1 TNT Manufacture[16]

1.3.1.1 Introduction

The explosives industry as a whole includes companies that manufacture organic nitration products and formulate mixtures of chemicals with explosive properties. Some 135 chemicals or formulations useful as explosives have been identified. Seventy-five are used industrially in mining, quarrying, excavating, and loosening oil and gas formations; 45 find primary application in military weapons, rockets, missiles, and space vehicles; and 15 are used for both purposes.

Commercial firms produce primarily ammonium nitrate-based explosives, dynamites, and nitroglycerin; TNT, HMX, and RDX are primarily military explosives; and nitroglycerin and nitrocellulose are components of military propellants.

Only production of TNT is treated in this section. Of the three military explosives (TNT, RDX, and HMX), TNT is the most extensively produced and more information concerning this process is available. Also, all of these manufacturing processes involve nitration of an organic molecule using nitric acid, indicating similarities between the processes.

1.3.1.2 Process Description[16]

Trinitrotoluene (TNT) is produced by the continuous processes pioneered by Canadian Industries Ltd. (CIL) at the Volunteer Army Ammunition Plant (VAAP). The manufacturing operation involves several direct processing steps. Nitration of toluene generates a raw product that is purified to obtain TNT. Spent acid from the nitration system and fumes from both nitration and purification are collected and recycled. The continuous nitration and purification (N&P) lines require spent acid tanks and settling tanks as well as provisions for handling red water wastes evolved in purification. Finishing and packaging operations are cleaner, but do generate a reddish-colored water waste from the dust scrubber. A schematic diagram[16] of the VAAP process flow is shown in Figure 1-6. This plant contains six such product lines rated at 50 tons/day each.

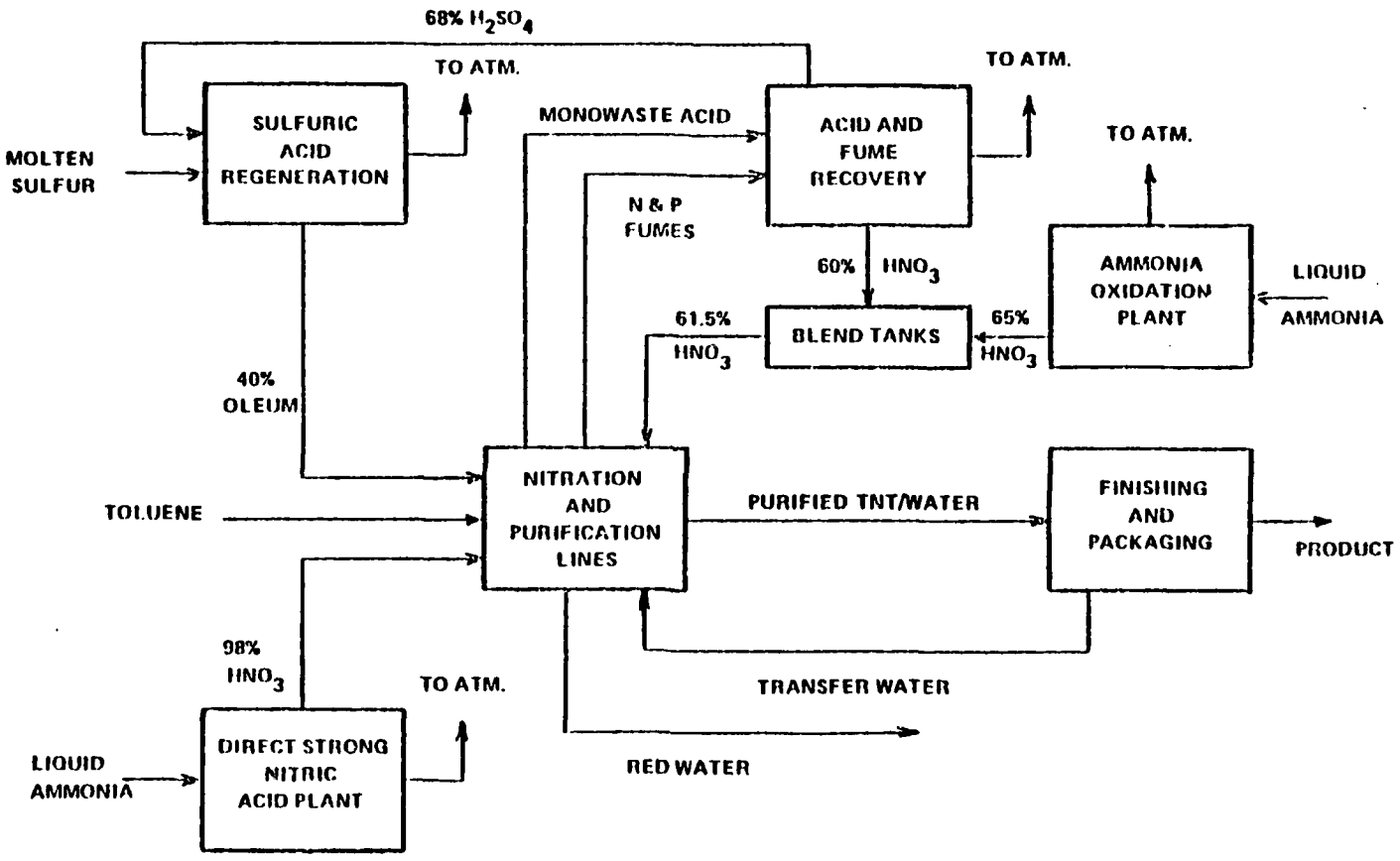


FIGURE 1-6

VOLUNTEER ARMY AMMUNITION PLANT PROCESS FLOW SCHEMATIC[16]

Processes supporting the manufacture of TNT are concerned primarily with the manufacture and treatment of nitric and sulfuric acids. Weak nitric acid is made either by an acid concentration system or at VAAP, by a direct strong nitric acid (DSN) plant. The nitric acid concentration/sulfuric acid concentration (NAC/SAC) unit strengthens weak nitric acid from the ammonia oxidation plant. Oleum and strong sulfuric acid are produced from weak sulfuric acid and sulfur in a special kind of oleum plant known as a sulfuric acid regeneration unit. The weak sulfuric acid is derived indirectly from the nitration lines. The waste acid drawn from the first nitrators is a mixture of nitric and sulfuric acids, with various nitrobodies. At VAAP, a portion of the acid and fume recovery (AFR) process is responsible for separating the acids.

Four processes inherent to the TNT manufacturing process are characterized as separate emission sources. They are:

- ammonia oxidation plant (AOP)
- acid fume recovery systems (AFR)
- direct strong nitric acid process (DSN)
- sulfuric acid regeneration (SAR)

These processes will be explained and characterized in the following subsections.

1.3.1.3 Ammonia Oxidation Plant (AOP)

1.3.1.3.1 Process Description

AOP's are responsible for the production of weak (60 to 65 percent) nitric acid from liquid ammonia and ambient air. The process is primarily dependent on the catalytic oxidation of NH_3 to NO followed by reaction of NO to NO_2 and subsequent absorption by water to HNO_3. Figure 1-7 provides a flow diagram[16] for a typical AOP. Operating and emission characteristics can be found in the nitric acid plant section of this chapter.

The most significant emission from any ammonia oxidation plant is NOx. The older plants emitted large amounts of NOx as an orange plume. Newer AOP units and modified older plants have sharply reduced NOx emissions attributable to weak HNO_3 production.

VAAP has a new 383-ton weak (60 percent) HNO_3/day AOP for use in TNT production. The apparent overcapacity (with respect to six continuous TNT lines) allows for the use of the older batch-process lines or the future installation of additional continuous lines.

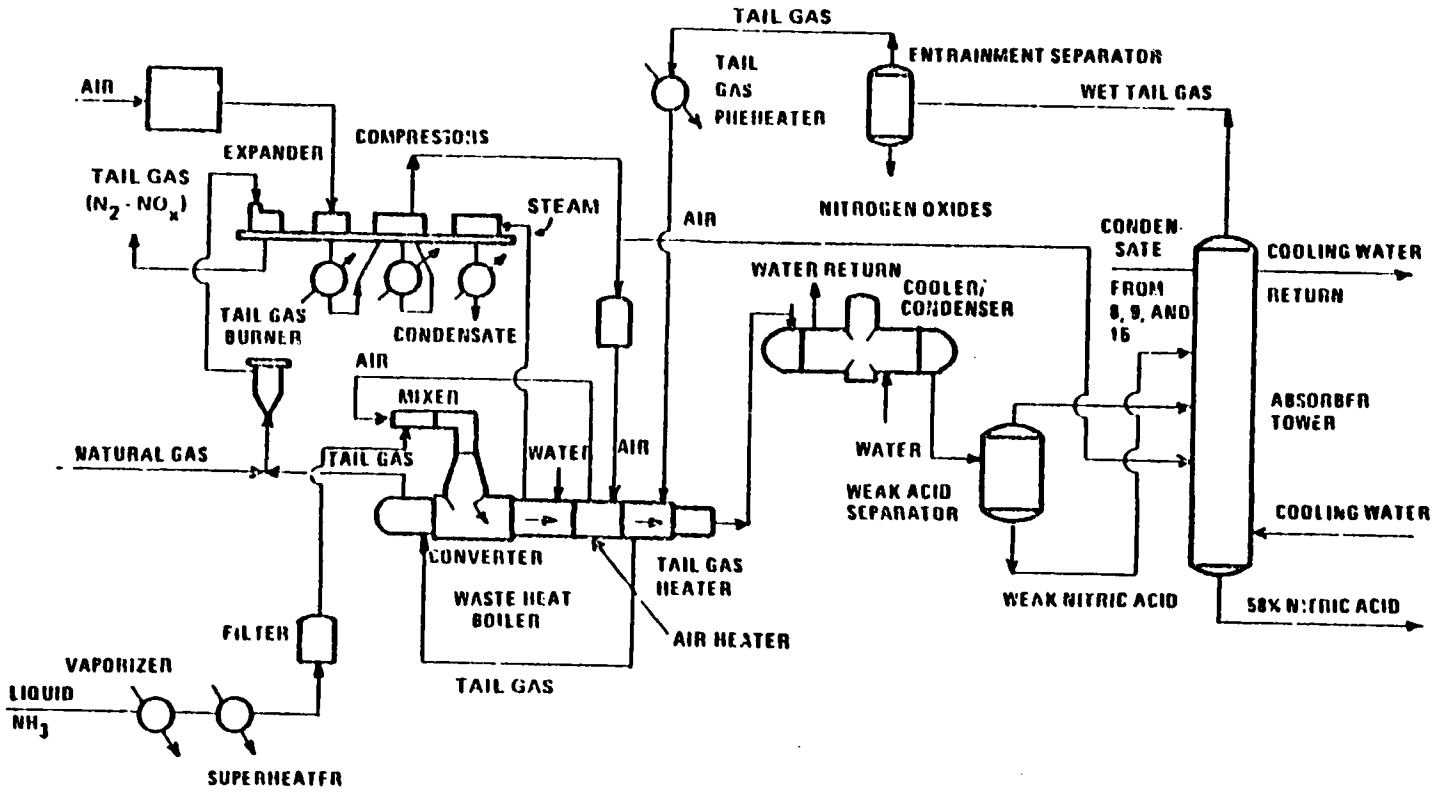


FIGURE 1-7

AMMONIA OXIDATION PLANT PROCESS FLOW SCHEMATIC [16]

1.3.1.3.2 Emission Characterization

At full production capacity, the AOP at VAAP produces 383 tons/day of weak nitric acid. Stack gas flow rate is 27,744 acfm at 662°F and controlled NOx emission rate is 25.5 lb/hr. Uncontrolled NOx emission rate is calculated to be 475 to 525 lb/hr.

1.3.1.3.3 Emission Factors

The only source for atmospheric emissions from the AOP is tail gas from the absorption tower. The National Emission Data System listing shows an emission factor of 5.0 lb./ton HNO_3 for the high pressure process described here.

1.3.1.3.4 Control Systems

At VAAP, a catalytic combustor system is used for NOx control. This is a commonly used control system for nitric acid absorber tail gas (see nitric acid plant section of this chapter).

1.3.1.4 Acid and Fume Recovery Process (AFR)

1.3.1.4.1 Process Description

The AFR plants recover and recycle waste acids and fumes from VAAP's N&P lines. The AFR plants have two major sections, acid treatment and fumes processing. The acid treatment part of the plant receives monowaste acid from the nitration lines of the continuous TNT process. The fumes processing plant collects fumes (primarily from the nitration lines) to extract NOx and form nitric acid. The two sections of the plant interact by use of common nitric acid concentration and bleaching units. A general diagram of VAAP's acid and fume recovery unit[16] appears in Figure 1-8.

The spent acid from the N&P lines contains both nitric and sulfuric acid. The acid processing unit utilizes the higher volatility and lower affinity for water of the nitric acid to effect a separation. Major steps in acid processing include preheating, denitration and separation, condensation and bleaching, and cooling and storage of product acids.

1.3.1.4.2 Operating Characteristics

Although the fumes processing unit receives some input from the acids processing portion of the plant, most of the feed comes from the N&P lines. Table 1-27 describes the fumes from the N&P lines.

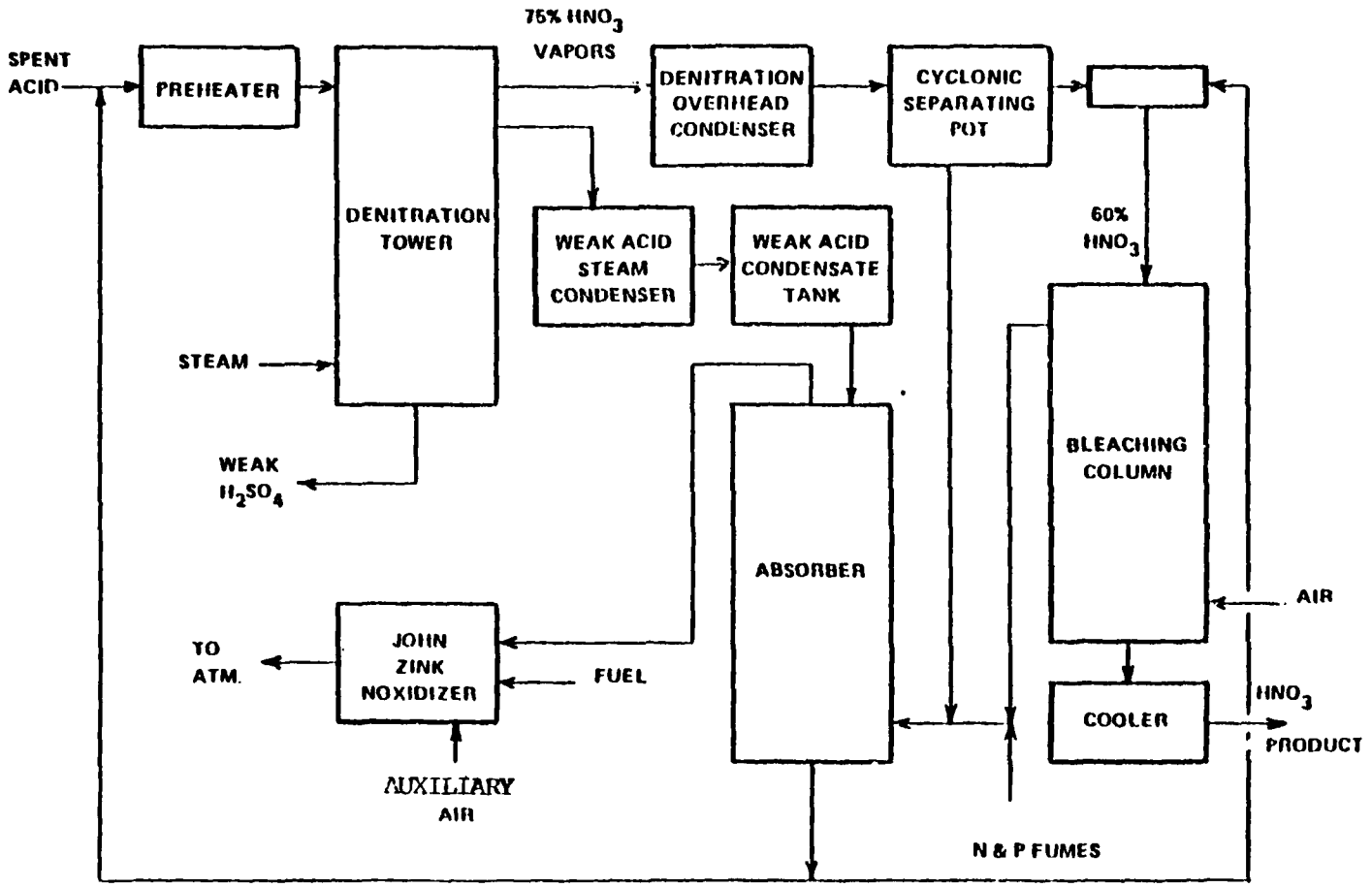


FIGURE 1-8

VAAP ACID AND FUME RECOVERY SYSTEM PROCESS FLOW SCHEMATIC [16]

1.3.1.5 Direct Strong Nitric Acid Process (DSN)

1.3.1.5.1 Process Description

The DSN process utilized at VAAP provides a means of producing strong nitric acid without an intermediate weak nitric acid product. The DSN process, like an AOP plant, oxidizes NH_3 to NO before conversion of NO to NO_2. At this point, however, the method differs. Rather than producing a weak HNO_3 stream, the NO_2 is used to produce the condensation product of two NO_2 molecules, N_2O_4. This dimer may then be reacted with water and oxygen to form strong HNO_3 directly. Figure 1-9 is a schematic diagram of the process.[16]

1.3.1.5.2 Operating Characteristics

The 336-ton/day, 98 percent DSN is designed for the following raw material rates (lb/hr): ammonia, 9,044; oxygen, 6,111; nitrogen, 188; water, 45; air, 108,500. The corresponding nitric acid production is 28,000 lb/hr (98 percent HNO_3).

1.3.1.5.3 Emission Characterization

Table 1-31 shows actual emissions from the direct strong nitric acid plant at the VAAP facility.

TABLE 1-31

DSN EMISSIONS AT VAAP

Source	Stack Gas Flow Rate (acfm)	Temp. (°F)	NO (lb/hr)	NO_2 (lb/hr)
DSN	24,705	59	15.38	4.08

Note: The unit operates 13 days/month.

The primary source of emissions from the DSN process is the condenser vent for the final absorption tower. Non-condensables from the bleaching tower condenser are fed into the suction side of the steam-driven compressor feeding the oxidation towers. Vapors from the reactor acid separators go to the raw acid tank, which is in turn vented to the N_2O_4 precondensers of the bleaching tower.

The rated capacity of the DSN plant is 340 tons, 98 percent HNO_3/day. VAAP's six continuous TNT lines demand only about 146 tons strong HNO_3/day; therefore, operation is projected at capacity for 13 days/month.

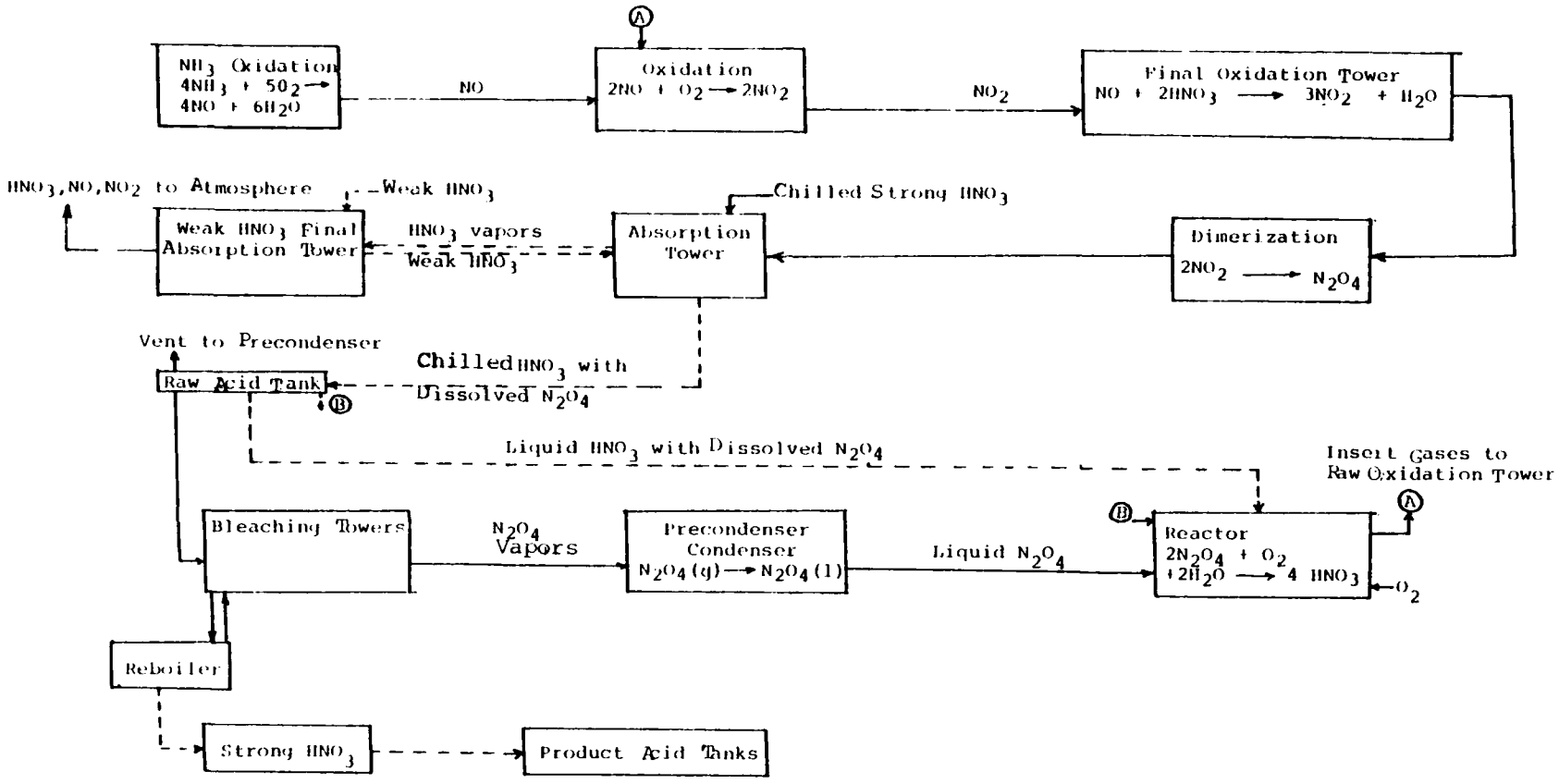


FIGURE 1-9

DIRECT STRONG NITRIC ACID PROCESS FLOW SCHEMATIC [16]

TABLE 1-27

DESIGN NITRATOR FUME INPUT TO THE FUME PROCESSING SEGMENT OF THE AFR PLANT, VAAP[16]

Compound	lb/hr at 100% Design Capacity
N_2	10,686
O_2	3,107
CO_2	1,499
NO_2	895
CO	519
$C(NO_2)_4$	86

Note: One AFR unit serves three lines.

Due to the interaction of the fume processing section with the acid treatment portion of the AFR unit, the product (nitric acid) of the fume processing system appears in the output from the acid treatment unit. The product consists of the two separated acid streams described in Table 1-28.

TABLE 1-28

AFR PRODUCTION RATES AT CAPACITY, VAAP[16]

Product	Rate, lb/hr at Design Capacity
68% H_2SO_4	55,103
60% HNO_3	9,427

Note: One AFR unit serves three TNT lines.

TABLE 1-29

AFR EMISSIONS AT VAAP[16]

Source	Stack Gas Flow Rate (acfm)	Temp. (°F)	NOx (lb/hr)
AFR No. 1	21,863	421	18.42
AFR No. 2	21,863	421	18.42

4,140 lb/day tetranitromethane (TNM) fed to each AFR.

TABLE 1-30

EMISSION FACTORS FOR AN AFR PROCESS[3]

	Nitrogen Oxides (NO_2, lb/ton)		Nitric Acid Mist (100% HNO_3, lb/ton)	
	Avg.	Range	Avg.	Range
Acid Recovery	8	(6.7-10)	1	(0.3-1.9)
Fume Recovery	3	(1-4.5)	0.02	(0.01-0.03)

1.3.1.4.3 Emission Characterization

Table 1-29 shows actual emissions from the AFR facility at the VAAP. The fume processing units exist at VAAP to reduce emissions from the N&P lines by capturing valuable airborne NOx. No pollution originates with the unit. However, the fume processing unit is the exit point for any NOx or TNM not recovered from the nitrator feed.

1.3.1.4.4 Emission Factors

Table 1-30 shows emission factors for the acid and fume recovery processes applicable to a continuous TNT process. The emission rate (lb/hr) for a given pollutant can be obtained by multiplying its emission factor (lb/ton) with the TNT production rate (ton/lb).

1.3.1.4.5 Control Systems

The gases from the TNT N&P lines are high in NOx. A considerable portion of nitrogen oxides is also evolved in the acid processing section of the AFR unit. To avoid potential pollution, the fumes processing unit applies absorption and incineration to reduce the level of NOx emissions from both sources. Absorption of NO_2 produces a useful product, HNO_3. Incineration, however, consumes fuel to convert NOx to N_2. Some of the energy consumed in incineration is recovered in a waste heat boiler.

Fumes are collected from several sources for reactant recovery and gas cleaning. The input blower system draws fumes from the TNT production lines, the nitric acid storage tanks, the denitrator overhead condenser, the acid steam condenser, and the bleacher. The blower can operate at various flow rates but is subject to surging below 50 percent capacity. Air may be bled to the blower intake to avoid surging.

1.3.1.5.4 Emission Factors

No emission factors have been developed for the DSN process at VAAP. However, generalized figures are available in the nitric acid plants, Section (1.3.3).

1.3.1.5.5 Control Systems

The type of control systems available for DSN plants is detailed in the nitric acid plant section of this chapter. The control system used at VAAP is a condenser system at the vent of the final absorption tower.

1.3.1.6 Sulfuric Acid Regeneration (SAR)

1.3.1.6.1 Process Description

The term SAR is applied to oleum plants that produce sulfuric acid from a feed stock that consists, at least partially, of sulfuric acid. In most SAR plants, the feed of a weak (68 percent is common) sulfuric acid to the combustor is supplemented by liquid sulfur for makeup purposes.

To produce oleum, a plant must receive a gas stream rich in SO_2. Before entering the converter system, the dilute sulfuric acid and molten sulfur must be reacted to yield SO_2 by burning the feed in a combustion furnace system. Here sulfuric acid, sulfur, and fuel oil are mixed with preheated air and burned to provide 8.5 percent SO_2. The heat from the combustion reaction is partially recovered in steam production. A waste heat boiler produces steam that is routed to another heat exchanger to be superheated. The steam supplies the SAR and provides an excess to be exported to the plant steam distribution system.

The chilled gas coolers (supplied with chilled water from a vacuum refrigeration system) operate at low temperatures to condense excess water from the gas stream. The condensate combines with SO_3 to produce a 2 to 3 percent H_2SO_4 liquor that is conveyed by an acid sewer to neutralization facilities. A mist eliminator prevents condensates from continuing through the plant. A packed drying tower is employed to completely dry the gas stream with 93 percent sulfuric acid liquor, sprayed against the rising gas stream.

The converter-heat exchanger system must provide the proper environment for the production of SO_3. The catalytic converter uses vanadium pentoxide to produce SO_3 by the simple reaction:

$$2SO_2 + O_2 \longrightarrow 2SO_3$$

The SO_3 generated in the converter must be absorbed to form oleum and acid products. Three absorption towers are required to remove the SO_3 from the gas stream. The first tower produces 40 percent oleum and uses 20 percent oleum as makeup. A second tower produces 20 percent oleum and uses 98 percent sulfuric acid as makeup. The last tower produces 98 percent sulfuric acid by the reaction:

$$SO_3 + H_2O \longrightarrow H_2SO_4$$

The makeup for the third tower is 93 percent sulfuric acid.

The gas stream may be further cleaned by using the SO_2 to produce sellite, an agent used to remove oxidizing impurities from TNT. The resulting sellite solution is used in the purification of TNT. A diagram of VAAP's single contact-single absorption (SC/SA) plant for sulfuric acid regeneration[16] appears in Figure 1-10.

1.3.1.6.2 Emission Characterization

At full production capacity, the SAR at VAAP produces 580 tons/day of 100 percent sulfuric acid and 53 tons/day of finished sellite with input rates of 550 tons/day of sulfuric acid and 25 tons/day of elemental sulfur.

Although there is only one major emission point in VAAP's SAR plant, there are three significant emission types: NOx, SO_2, and sulfuric acid mist. The NOx results from the combustion of sulfuric acid containing some HNO_3. The SO_2 emissions reflect incomplete conversion of SO_2 to SO_3 in the catalytic reactor. Exiting SO_2 must also escape the sellite production scrubber. Acid mists introduced in the absorption process are diminished through the use of mist eliminators.

At full capacity, the gas flow rate is 40,544 acfm at 86°F. The NOx emissions are 106.6 lb/hr; sulfuric acid mist, 25 lb/hr; SO_2, 90.7 lb/hr.

1.3.1.6.3 Emission Factors

Table 1-32 presents emission data for the VAAP's SAR unit, which were taken at 71.5 to 94.3 percent capacity operation.

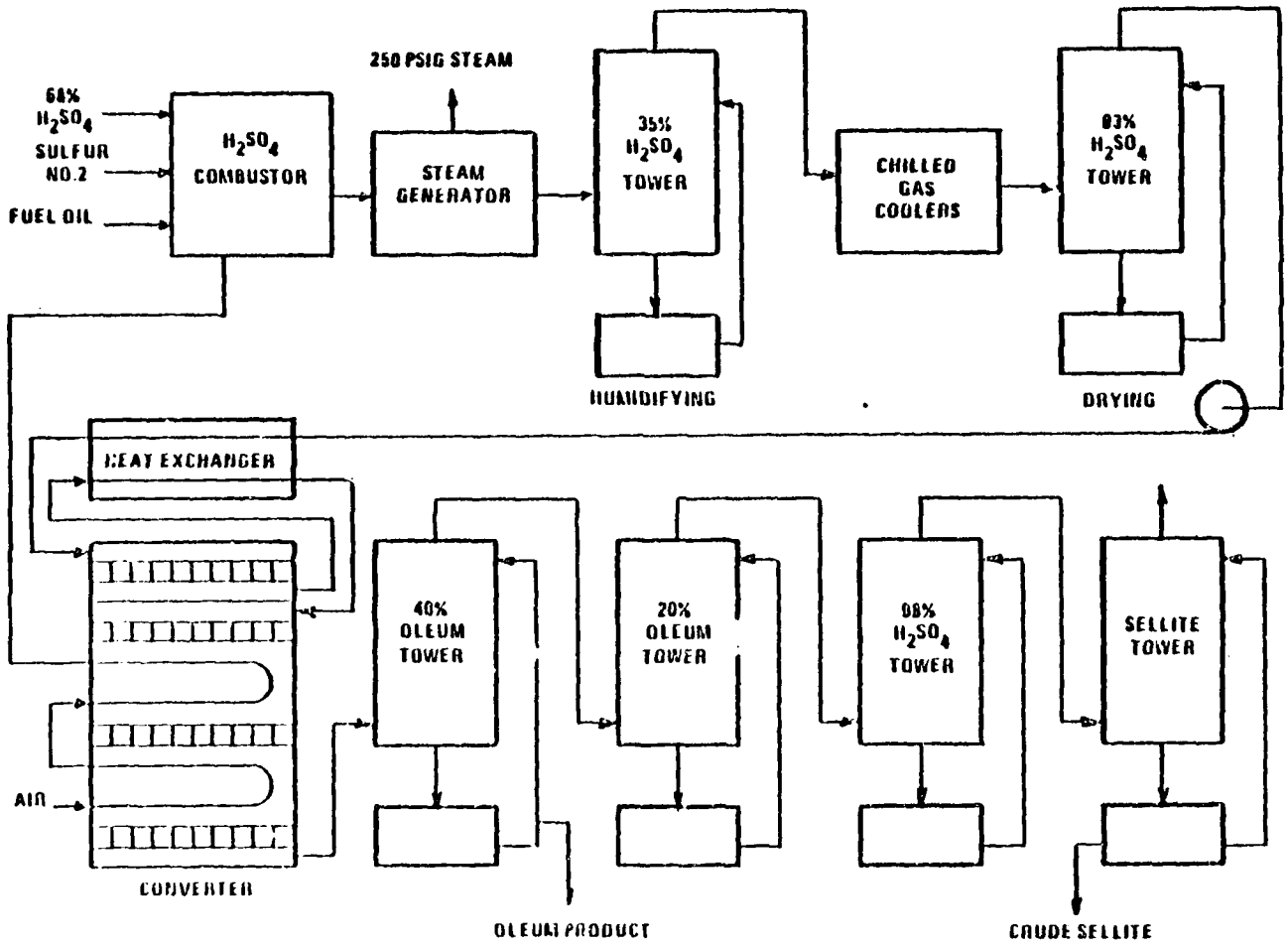


FIGURE 1-10

SINGLE CONTACT/SINGLE ABSORPTION SULFURIC ACID REGENERATION PLANT

TABLE 1-32

EMISSIONS RATE FOR THE VAAP SAR PLANT

Compound	Acid Production (tons/day)	Emissions (lb/ton H_2SO_4)
NO_2	546.8	4.05
SO_2	414.6	5.06
H_2SO_4	414.6	1.45

1.3.1.6.4 Control System Applications

The point of origin of emissions from the SAR plant is noted below, together with controls and appropriate points of measure.

Sources	Control	Emission Point
NOx resulting from combustion of H_2SO_4 containing some HNO_3	None	Exit stack from sellite tower
SO_2 from incomplete conversion of SO_2 to SO_3	Sellite tower	Exit stack from sellite tower
H_2SO_4 mist generated in absorption tower	Mist eliminators	Exit stack from sellite tower

Additional details on mist eliminator applications can be found in sulfuric acid plants section of this chapter.

1.3.2 Sulfuric Acid Plants (Contact Process)[17]

1.3.2.1 Process Characterization

All sulfuric acid is made by either the lead chamber or the contact process. The contact process accounts for more than 97 percent of the total sulfuric acid production in the United States, and so the only process discussed in this section. Contact plants are generally classified according to the raw materials charged to them: (1) elemental sulfur burning, (2) spent acid and hydrogen sulfide burning, and (3) sulfide ores and smelter gas burning plants. The relative contributions from each type of plant to the total acid production are 68, 18.5, and 13.5 percent, respectively.

All contact processes incorporate three basic operations, each of which corresponds to a distinct chemical reaction. First, the sulfur in the feedstock is burned to sulfur dioxide:

$$S + O_2 \longrightarrow SO_2 \qquad (1)$$

Then, the sulfur dioxide is catalytically oxidized to sulfur trioxide:

$$2SO_2 + O_2 \longrightarrow 2SO_3 \qquad (2)$$

Finally, the sulfur trioxide is absorbed in a strong, aqueous solution of sulfuric acid:

$$SO_3 + H_2O \longrightarrow H_2SO_4 \qquad (3)$$

1.3.2.2 Elemental Sulfur Burning Plants

Elemental sulfur, such as Frasch-process sulfur from oil refineries, is melted, settled, or filtered to remove ash and is fed into a combustion chamber. The sulfur is burned in clean air that has been dried by scrubbing with 93 to 99 percent sulfuric acid. The gases from the combustion chamber are cooled and then enter the solid catalyst (vanadium pentoxide) converter. Usually, 95 to 98 percent of the sulfur dioxide from the combustion chamber is converted to sulfur trioxide, with an accompanying large evolution of heat. After being cooled, the converter exit gas enters an absorption tower where the sulfur trioxide is absorbed with 98 to 99 percent sulfuric acid. The sulfur trioxide combines with the water in the acid and forms more sulfuric acid.

If oleum, a solution of uncombined SO_3 in H_2SO_4, is produced, SO_3 from the converter is first passed to an oleum tower that is fed with 98 percent acid from the absorption system. The gases from the oleum tower are then pumped to the absorption column where the residual sulfur trioxide is removed.

A schematic diagram of a contact-process sulfuric acid plant that burns elemental sulfur[17] is shown in Figure 1-11.

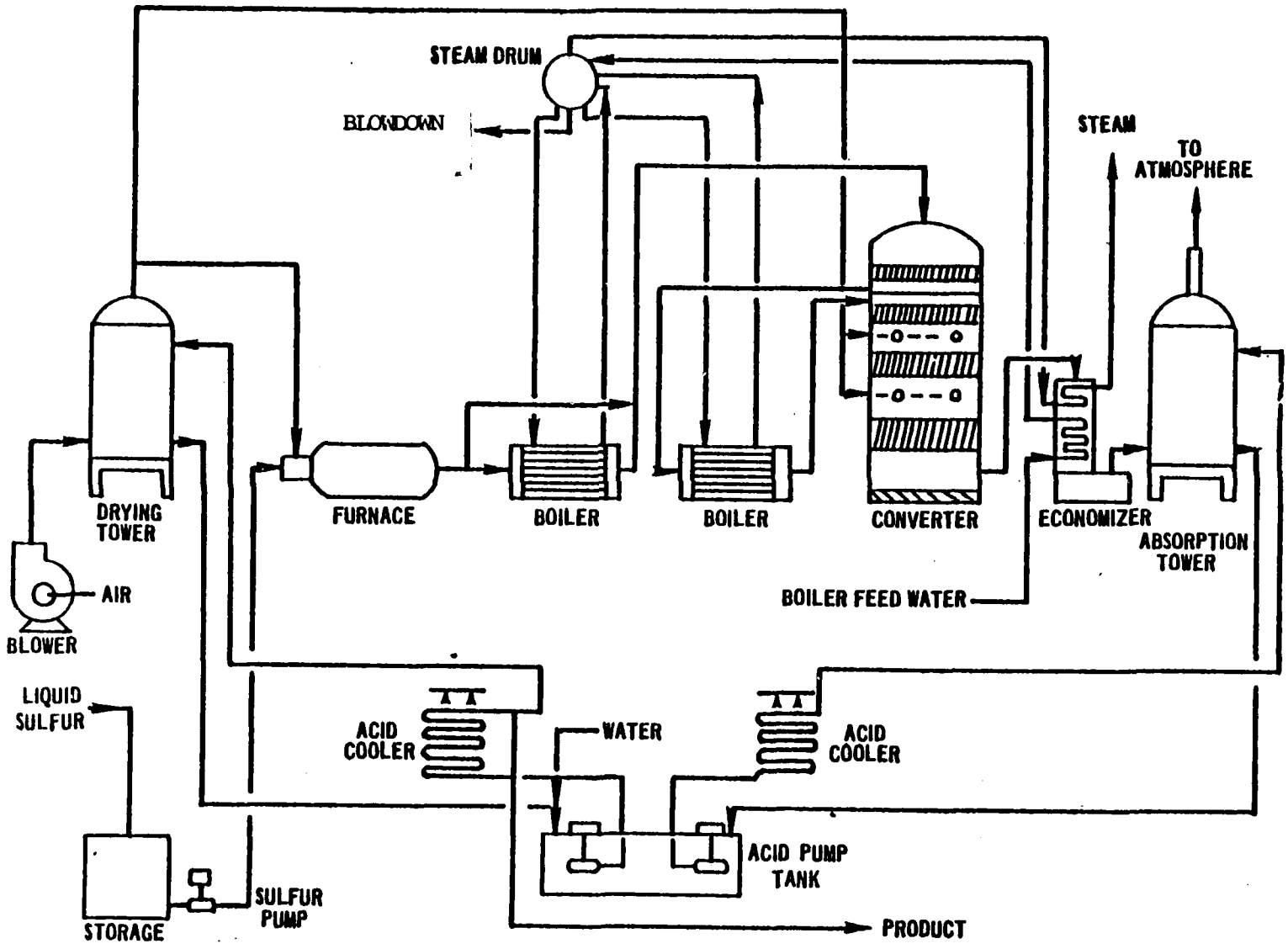


FIGURE 1-11

BASIC FLOW DIAGRAM OF CONTACT-PROCESS SULFURIC ACID PLANT BURNING ELEMENTAL SULFUR[17]

1.3.2.3 Spent Acid and Hydrogen Sulfide Burning Plants

Sulfuric acid can be produced from spent acid and hydrogen sulfide by two different processes. In one, the sulfur dioxide and other products from the combustion of spent acid and/or hydrogen sulfide with undried atmospheric air are passed through gas-cleaning and mist-removal equipment. The gas stream next passes through a drying tower. A blower draws the gas from the drying tower and discharges the sulfur dioxide gas to the sulfur trioxide converter. A schematic diagram of a contact-process sulfuric acid plant that burns spent acid[17] is shown in Figure 1-12.

In a "wet-gas plant," the wet gases from the combustion chamber are charged directly to the converter with no intermediate treatment. The gas from the converter flows to the absorber, through which 93 to 98 percent sulfuric acid is circulating.

1.3.2.4 Operating Characteristics

Operating data from sulfuric acid plants with and without mist eliminators[17] are shown in Tables 1-33 and 1-34, respectively. The data in these tables were supplied by the major manufacturers of sulfuric acid. Data from essentially all types of sulfuric acid plants are included.

1.3.2.5 Emission Characterization

1.3.2.5.1 General

Tables 1-35 and 1-36 contain emission data from sulfuric acid plants with and without mist eliminators.[17] Data from essentially all types of sulfuric acid plants are included. They represent results from 27 plants and include results from stack sampling programs conducted jointly by the Manufacturing Chemists Association and the Public Health Service.

The major source of emissions from contact sulfuric acid plants is the exit gas from the absorber. This gas contains unreacted sulfur dioxide, sulfuric acid spray and mist, and unabsorbed sulfur trioxide. Trace amounts of nitrogen oxides may also be present under some conditions, e.g., use of a raw material feed containing nitrogen compounds.

Unconverted sulfur dioxide gas, which is colorless, passes through the absorption system and is discharged to the atmosphere. The quantity of this gas emitted is a direct function of the degree of conversion of sulfur dioxide to sulfur trioxide and may vary from 0.1 to 0.5 percent by volume of the stack gases. During start-up or during some emergency shutdowns, higher concentrations will occur.

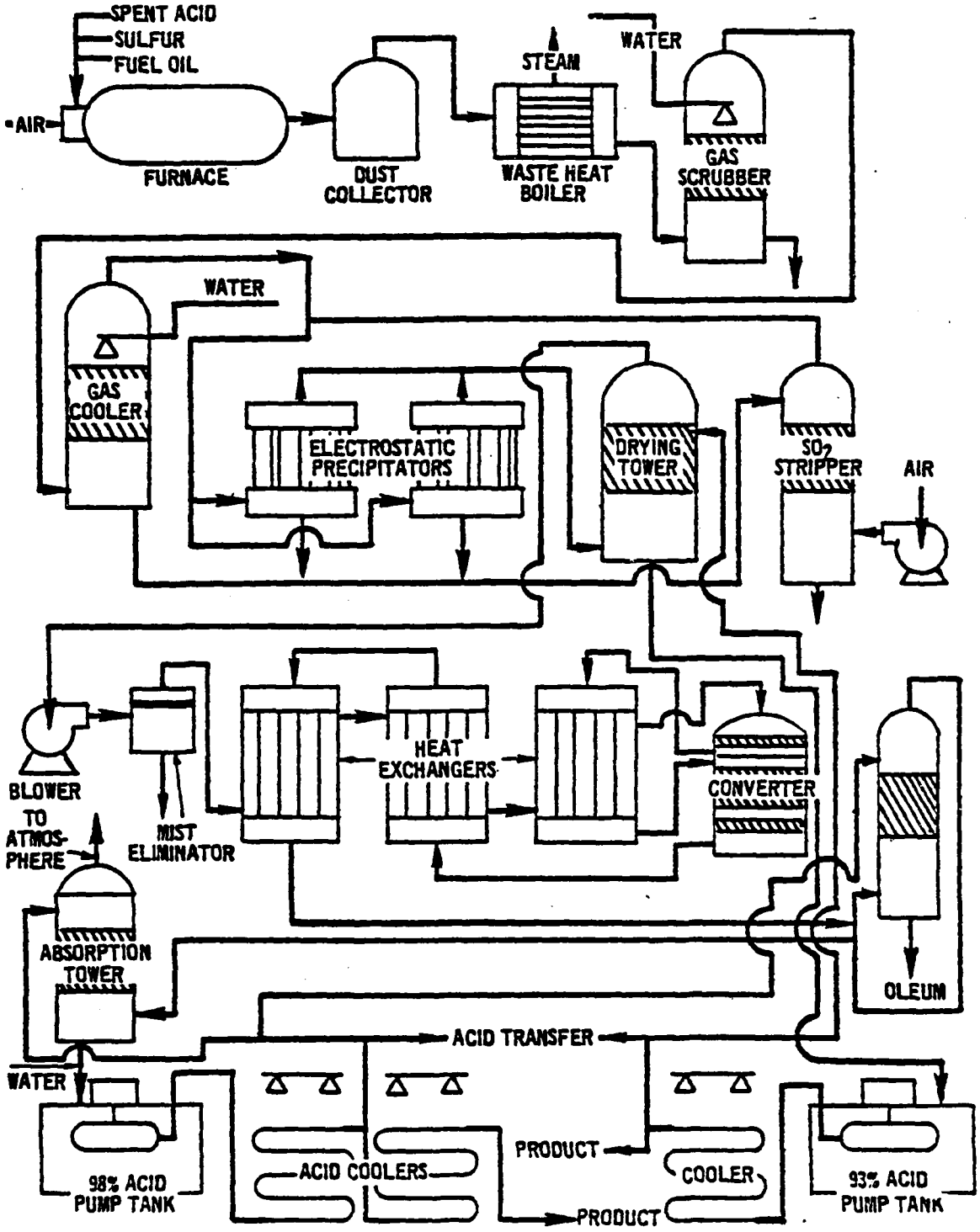


FIGURE 1-12

BASIC FLOW DIAGRAM OF CONTACT-PROCESS SULFURIC ACID PLANT BURNING SPENT ACID [17]

TABLE 1-33

TYPICAL OPERATING DATA FROM CONTACT SULFURIC ACID PLANTS WITH MIST ELIMINATORS[17]

Plant Type	Raw Material	H_2SO_4 Production (tons/day)	Oleum Mode (% of output)	Stack Gas Temp. (°F)	Stack Gas Flow Rate (Scfm)	SO_2 Entering Converter (vol %)	Conversion of SO_2 to SO_3 (%)
Sulfur burning							
- air dilution	Molten dark	961	0	186	58,000	8.0	98.5
- no air dilution	Molten dark	150	0	165	7,400	8.0	97.6
- no air dilution	Molten dark	150	13	166	7,400	8.0	97.5
Combination	Spent acid, H_2S, and sulfur	219	0	76	12,000	8.2	97.2
		133	56	76	7,000	8.4	98.2
		300	0	175	17,000	9.0	96.9
		265	0	160	14,000	7.2	98.0
		300	0	180	21,000	7.4	97.8

TABLE 1-34

TYPICAL OPERATING DATA FROM CONTACT SULFURIC ACID PLANTS WITHOUT MIST ELIMINATORS [17]

Plant Type	Raw Material	H_2SO_4 Production (tons/day)	Oleum Mode (% of output)	Stack Gas Temp. (°F)	Stack Gas Flow Rate (Scfm)	SO_2 Entering Converter (vol %)	Conversion SO_2 to SO_3 (%)
Sulfur burning, air dilution	Molten dark	735	0	212	45,000	8.0	97.8
		650	0	175	48,000	7.0	96.0
		120	50	115	7,600	8.0	97.5
		422	51	123	25,100	8.8	97.5
	Molten recovered	130	0	175	7,500	8.8	97.2
		100	0	173	6,500	8.0	97.8
		100	0	168	6,500	8.0	97.8
Sulfur burning, no air dilution	Molten recovered	325	43	148	16,700	10.5	95.7
		325	43	148	16,600	10.6	95.6
		162	35	140	11,000	7.7	98.5
		115	0	94	5,900	8.6	98.1

TABLE 1-34 (Cont)

Plant Type	Raw Material	H_2SO_4 Production (tons/day)	Oleum Mode (% of output)	Stack Gas Temp. (°F)	Stack Gas Flow Rate (Scfm)	SO_2 Entering Converter (vol %)	Conversion SO_2 to SO_3 (%)
Sulfur burning, no air dilution	Solid bright	210	0	-	12,000	9.1	97.7
	Solid dark	500	40	170	30,000	10.2	96.7
	Molten dark	310	100	105	17,000	9.0	97.6
	Molten dark	265	33	170	14,000	10.0	96.5
Metallurgical	Pyrite spent acid	285	25	190	18,200	8.0	97.5
	Pyrite	500	0	-	35,000	-	-
	By-product gas	100	-	-	5,000	-	-
Spent acid	Spent acid and sulfur	650	77	136	58,500	8.8	96.3
	Spent acid and sulfur	302	71.5	163	19,700	8.0	97.8
	Spent acid	900	0	145	62,000	7.5	95.9
	Spent acid, H_2S, and sulfur	-	18	178	34,000	6.7	97.3

TABLE 1-35

TYPICAL EMISSIONS FROM CONTACT SULFURIC ACID PLANTS WITH MIST ELIMINATORS[17]

Plant Type	Raw Material	H_2SO_4 Production (tons/day)	Stack Gas Temp. (°F)	Stack Gas Flow Rate (Scfm)	SO_2 Emission (lb/ton of H_2SO_4 produced)	Mist Eliminator[a]	Acid Mist leaving Absorber (lb/ton acid)	Acid Mist leaving Eliminator (lb/ton acid)
Sulfur burning - air dilution	Molten dark	961	186	58,000	21.64	W M	-	1.25
- no air dilution	Molten dark	150	165	7,400	24.00	W M	7.63	0.56
- no air dilution	Molten dark	150	166	7,400	25.33	W M	5.83	3.66
Combination	Spent acid, H_2S, and sulfur	219	76	12,000	36.53	ESP	1.23	0.03
		133	76	7,000	22.56	ESP	4.84	0.05
		300	175	17,000	46.67	GF	0.18-0.72	0.09-0.36
		265	160	14,000	21.13	GF	3.45	0.04
		300	180	21,000	34.00	GF	7.10	0.42

[a]Abbreviations: WM = wire mesh, ESP = electrostatic precipitator, GF = glass fiber.

TABLE 1-36

TYPICAL EMISSIONS FROM CONTACT SULFURIC ACID PLANTS WITHOUT MIST ELIMINATORS[17]

Plant Type	Raw Material	H_2SO_4 Production (tons/day)	Stack Gas Temp. (°F)	Stack Gas Flow Rate (Scfm)	SO_2 Emission (lb/ton) acid)	Acid Mist Emission (lb/ton acid)
Sulfur burning, air dilution	Molten dark	735	212	45,000	31.29	0.46
		650	175	48,000	58.46	3.63
		650	175	48,000	58.46	5.57
		120	115	7,600	36.67	-
		422	123	25,000	37.91	0.57
	Molten recovered	130	175	7,500	41.54	1.69
		100	173	6,500	34.00	0.80
		100	168	6,500	34.00	0.40
Sulfur burning, no air dilution	Molten recovered	325	148	16,700	69.54	0.37
		325	148	16,600	70.77	0.80
		162	140	11,000	22.22	0.49
		115	94	5,900	24.35	0.35

TABLE 1-36 (Cont.)

Plant Type	Raw Material	H_2SO_4 Production (tons/day)	Stack Gas Temp. (°F)	Stack Gas Flow Rate (Scfm)	SO_2 Emission (lb/ton acid)	Acid Mist Emission (lb/ton acid)
Sulfur burning, no air dilution	Solid bright	210	-	12,000	35.24	1.71
	Solid dark	500	170	30,000	61.60	1.16
	Molten dark	310	105	17,000	38.84	0.32
	Molten dark	265	170	14,000	56.60	6.26
Metallurgical	Pyrite spent acid	285	190	18,200	37.89	1.19
	Pyrite	500	-	35,000	-	7.36
	By-product gas	100	-	5,000	-	2.20
Spent acid	Spent acid and sulfur	650	136	58,500	85.23	2.40
	Spent acid and sulfur	302	163	19,700	33.11	2.25
	Spent acid	900	145	62,000	60.00	2.22

Emissions of sulfuric acid mist and spray usually vary from 3 to 15 mg/scf of gas; values ranging from as low as 1 mg to as high as 50 mg/scf have been observed. The appearance of a dense white plume at the absorber exit stack indicates the presence of a substantial number of small particles (i.e., less than 3 microns in diameter) and does not necessarily reflect the concentration of sulfuric acid mist present.

Unabsorbed sulfur trioxide usually constitutes a small part of the absorber exit gas. When discharged to the atmosphere, it is hydrated and forms a visible white plume of acid mist. Although the concentration of unabsorbed sulfur trioxide can vary appreciably, from 0.5 to 48 mg/scf of gas, it is usually closer to the lower figure and is a small part of the total acid mist emission.

1.3.2.5.2 Emission Factors--Sulfur Dioxide

Testing has shown that the mass of SO_2 emissions is an inverse function of the sulfur conversion efficiency (SO_2 oxidized to SO_3). This conversion is, in turn, affected by the number of stages in the catalytic converter, the amount of catalyst used, the temperature and pressure, and the concentrations of the reactants, sulfur dioxide and oxygen. The Environmental Protection Agency performance standard for new and modified plants is 4.0 pounds per ton (2 kg/MT) of 100 percent acid produced, maximum 2-hour average. As Table 1-37 indicates[3], achieving this standard requires a conversion efficiency of 99.7 percent in an uncontrolled plant or the equivalent SO_2 collection mechanism in a controlled facility. Most single absorption plants have SO_2 conversion efficiencies ranging from 95 to 98 percent.

1.3.2.5.3 Emission Factors--Acid Mist

Table 1-38 presents uncontrolled acid mist emission factors[3] for various sulfuric acid plants. These factors are dependent on the type of raw material used, since hydrocarbon impurities oxidize to water vapor during combustion. This water vapor is then available to form acid mist. The factors are also dependent on the amount of oleum produced since oleum plants produce greater quantities of finer, more stable mist. Finally, it should be emphasized that the percentage conversion of sulfur dioxide to sulfur trioxide has no direct effect on acid mist emissions.

TABLE 1-37

EMISSION FACTORS FOR SULFURIC ACID PLANTS[3]

Conversion of SO_2 to SO_3, (%)	SO_2 Emissions lb/ton of 100% H_2SO_4	SO_2 Emissions kg/MT of 100% H_2SO_4
93	96	48.0
94	82	41.0
95	70	35.0
96	55	27.5
97	40	20.5
98	27	13.0
99	14	7.0
99.5	7	3.5
99.7	4	2.0
100	0	0.0

Note: The following linear interpolation formula can be used for calculating emission factors for conversion efficiencies between 93 and 100 percent: emission factor (lb/ton acid) = -13.65 (percent conversion efficiency) + 1,365.

TABLE 1-38

ACID MIST EMISSION FACTORS
FOR SULFURIC ACID PLANTS WITHOUT CONTROLS[3]

Raw Material	Oleum Produced, % Total Output	Emissions Lb/Ton Acid	Emissions kg/MT Acid
Recovered Sulfur	0 - 43	0.35 - 0.8	0.175 - 0.4
Bright Virgin Sulfur	0	1.7	0.85
Dark Virgin Sulfur	33 - 100	0.32 - 6.3	0.16 - 3.15
Sulfide Ores	0 - 25	1.2 - 7.4	0.6 - 3.7
Spent Acid	0 - 77	2.2 - 2.7	1.1 - 1.35

Note: Emissions are proportional to the percentage of oleum in the total prouct. Use the low end of ranges for low oleum percentage and the high end of ranges for high oleum percentage.

1.3.2.6 Control Systems

1.3.2.6.1 Sulfur Dioxide Removal

Sulfur dioxide emissions can be reduced either through process modifications or tail-gas desulfurization. Process changes are more adaptable to new plants, whereas tail-gas treatments can be used on either new or existing plants.

1.3.2.6.1.1 Process Modifications

Two commercially available processes are modifications of the basic contact system, i.e., the dual absorption process and pressurized single-absorption process. Both are reportedly capable of reducing sulfur dioxide emission levels well below 500 ppm. However, there has been considerably more experience with dual absorption plants.

1.3.2.6.1.2 Tail Gas Desulfurization

Tail-gas desulfurization systems are capable of controlling sulfur dioxide emissions to almost any level. They include scrubbing with alkali solutions, dilute sulfuric acid, and hydrogen peroxide, and adsorption with molecular sieves.

The regenerative Wellman-Lord sodium sulfite system has been employed at acid plants as well as at oil-fired power plants. Sodium sulfite reacts with sulfur dioxide and forms bisulfite in the scrubber. Bisulfite is regenerated thermally to produce sulfur dioxide, which is fed back to the acid plant.

Gas scrubbing to remove sulfur dioxide with hydrogen peroxide produces weak acid directly (50% H_2SO_4 by weight). The weak acid is blended with other streams to increase product yield. As with sodium and ammonia systems, sulfur dioxide levels can be reduced below 100 ppm if necessary.

Sulfur dioxide can be absorbed from tail gases with molecular sieves. The sieves (synthetic zeolites) are desorbed thermally and sulfur dioxide is fed back to the converter to produce more acid. The process has been used successfully at a spent acid plant. Molecular sieves are capable of achieving sulfur dioxide levels below 100 ppm.

The US EPA has limited sulfur dioxide emissions from new plants to 4 pounds per ton of 100 percent acid, which is equivalent to an exit concentration of about 380 ppm for an elemental sulfur plant.

1.3.2.6.2 Acid Mist Removal

A number of devices of varying cost and efficiency are in use for removal of acid mist and spray from absorber tail gases. With any of them, relatively high efficiencies (over 90 percent) do not necessarily result in an invisible plume unless there are few particles less than 3 microns and inlet mist loading is not excessive. The following comments apply to devices used successfully on a commercial scale.

1.3.2.6.2.1 Wire-Mesh Mist Eliminators

The lowest first-cost device that effectively removes particles larger than about 3 microns diameter is the wire-mesh eliminator. Particle size and possibilities of corrosion from concentrated sulfuric acid mist must be carefully considered when selecting a wire-mesh eliminator. The eliminator is commonly constructed with two beds in series and operates with pressure drops of 1 to 3 inches of water. Test results for a two-stage wire-mesh eliminator, given in Table 1-35, show an acid-mist collection efficiency of 92.6 percent. The collection efficiency decreased to 37.3 percent, however, when oleum was produced. In this case, 62 percent of the particles were smaller than 3 microns. Although no plume was visible during production of 98 percent acid, a plume was plainly visible when oleum was also being produced.

1.3.2.6.2.2 Fiber Mist Eliminators

The high-efficiency glass-fiber mist eliminator is capable of operating with acid-mist collection efficiencies of over 99 percent. The acid-mist collection efficiencies for glass-fiber eliminators typically range from about 50 to 99.9 percent. The lower collection efficiencies were obtained with a glass-fiber unit specifically designed for collection at high velocity and medium efficiency. When a high-efficiency glass-fiber unit was utilized, an acid-mist collection efficiency of 99.9 percent was obtained for a tail-gas stream in which 38 percent of the particles, by weight, were 3 microns and smaller. The pressure drop for a high-efficiency glass-fiber mist eliminator is usually between 5 and 10 inches of water, but the system may be designed for higher or lower pressure drops, depending upon relative costs for power and equipment. The glass-fiber mist eliminator is also capable of maintaining a high mist-collection efficiency at varying tail-gas flow rates.

1.3.2.6.3 Electrostatic Precipitators

Electrostatic precipitators are highly efficient when used for collection of acid mist regardless of size of the

acid-mist particles. The acid-mist collection efficiencies for precipitators ranged from 92.2 to 99.9 percent.

Precipitators operate with pressure drops less than 0.5 inch of water and may be either of the wet or dry type. The dry type, which is suitable only for concentrated acid, is much less expensive but more susceptible to corrosion. Wet-type precipitators are suitable for use only with dilute acid and thus necessitate prior humidification of stack gases. Prehumidification also permits removal of sulfur trioxide by converting it to acid mist. The humidification step appreciably increases the cost of a wet-type installation.

1.3.3 Nitric Acid Plants[18,19]

1.3.3.1 Process Characterization

1.3.3.1.1 Weak Acid Production

Nearly all the nitric acid produced in the United States is manufactured by the high-pressure catalytic oxidation of ammonia (Figure 1-13).[18] Typically, this process consists of three steps, each of which corresponds to a distinct chemical reaction. First, a 1:9 ammonia-air mixture is oxidized at high temperature and pressure (6.4 to 9.2 atmospheres) as it passes through a platinum-rhodium catalyst, according to the reaction:

$$4NH_3 + 5O_2 \longrightarrow 4NO + 6H_2O \qquad (1)$$

After the process stream is cooled to 100°F (38°C) or less by passage through a cooler-condenser, the nitric oxide reacts with residual oxygen:

$$2NO + O_2 \longrightarrow 2NO_2 \rightleftharpoons N_2O_4 \qquad (2)$$

Finally, the gases are introduced into a bubble-cap plate absorption column where they are contacted with a counter current stream of water. The exothermic reaction that occurs is:

$$3NO_2 + H_2O \longrightarrow \underset{\substack{\text{Nitric Acid} \\ \text{50\% to 70\% Aqueous}}}{2HNO_3} + NO \qquad (3)$$

The production of nitric oxide in reaction (3) necessitates the introduction of a secondary air stream into the column to effect its oxidation to nitrogen dioxide, thereby perpetuating the absorption operation.

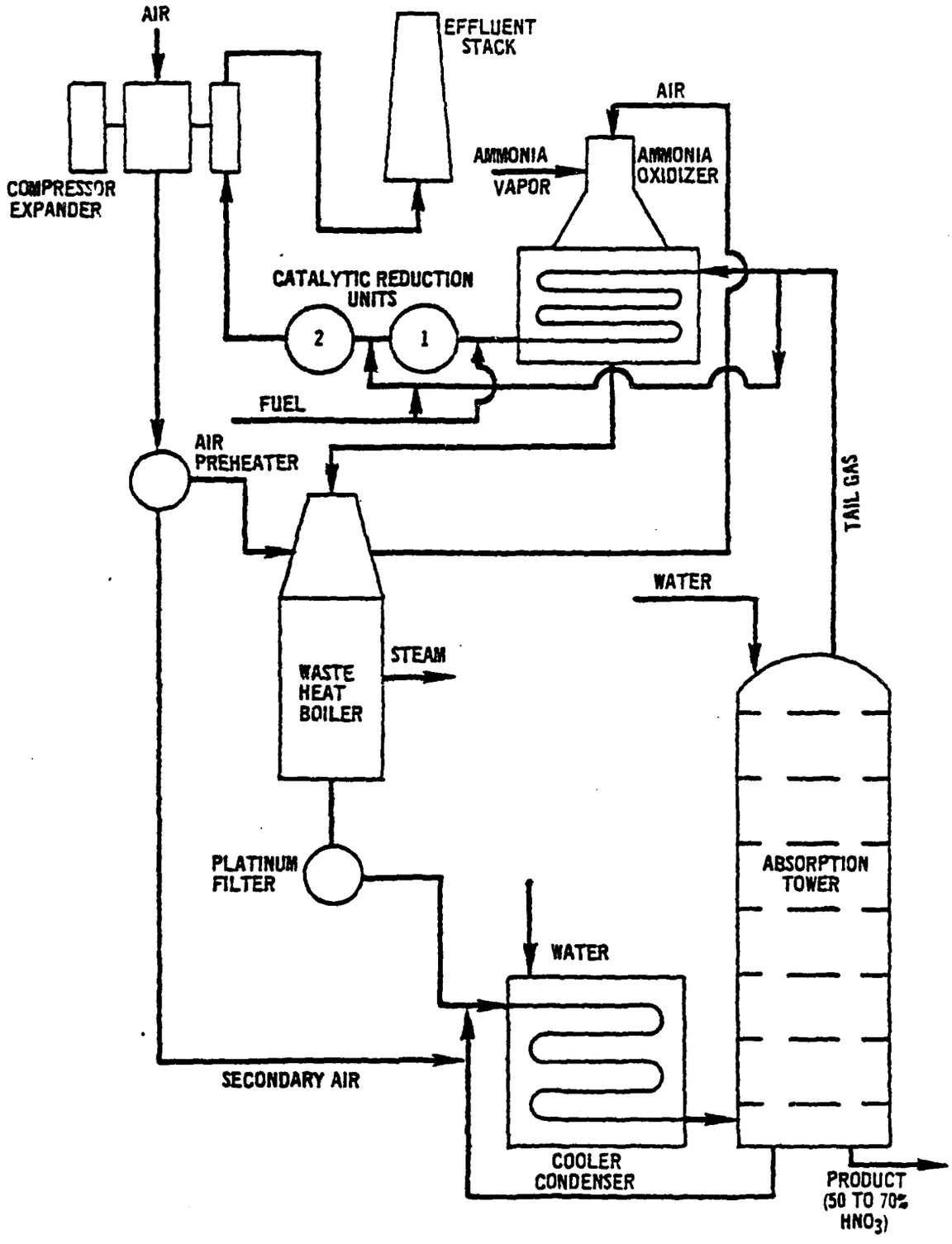


FIGURE 1-13

FLOW DIAGRAM OF TYPICAL NITRIC ACID PLANT USING PRESSURE PROCESS[18]

The spent gas flows from the top of the absorption tower to an entrainment separator for acid mist removal, through the ammonia oxidation unit for energy absorption from the ammonia stream, through an expander for energy recovery, and finally to the stack. In most plants, the stack gas is treated before release to the atmosphere by passage through either a catalytic combustor or, less frequently, an alkaline scrubber.

1.3.3.1.2 High Strength Acid Production

To meet requirements for high strength acid, the 50 to 70 percent acid produced by the pressure process is concentrated to 95 to 99 percent at approximately atmospheric pressure. The concentration process consists of feeding strong sulfuric acid and 60 percent nitric acid to the top of a packed column, where it is contacted by an ascending stream of weak acid vapor, resulting in the dehydration of the latter. The concentrated acid vapor that leaves the column passes to a bleacher and countercurrent condenser system to effect condensation of the vapors and separation of the small amounts of nitric oxides and oxygen that form as dehydration by-products. The by-products then flow to an absorption column where the nitric oxide mixes with auxiliary air to form nitrogen dioxide, which is, in turn, recovered as weak nitric acid. Finally, unreacted gases are vented to the atmosphere from the top of the column.

Table 1-39 presents typical operating data for 18 different nitric acid plants.[18]

1.3.3.2 Emission Characterization

Table 1-40 lists emission data for 11 different nitric acid plants[18] that use catalytic waste gas treatment equipment. Table 1-41 provides the same for seven other plants[18] that do not treat their waste gas. Operating conditions for all these plants are given in Table 1-39. Additional emission data[18] can be obtained from Figure 1-14.

The major source of atmospheric emissions is the absorption column in which nitrogen oxides are reacted with water to form nitric acid. The exit gas from the column contains unreacted nitrogen oxides (largely in the form of nitric oxide and nitrogen dioxide), oxygen, and nitrogen. Trace amounts of acid mist or vapor may also be present. Nitric oxide is a colorless gas; nitrogen dioxide is characterized by a reddish-brown color. The total concentration of nitrogen oxides normally ranges from 0.1 to 0.6 percent by volume of effluent prior to any treatment for control. Nitrogen dioxide accounts for about

TABLE 1-39

TYPICAL OPERATING DATA FROM NITRIC ACID PLANTS[18]

HNO_3 Rate 100% (tons/day)	Normal Acid Strength (%)	NH_3 Feed Rate (lb/hr)	NH_3 Oxidation Pressure (psig)	Absorption Column Exit Press. (psig)	Absorption Column Exit Temp. (°F)	Tail Gas Flow Rate (Scfm)
Controlled						
139.7	51	3,346	105	70	85	10,900
120.0	60	2,880	120	94	78	9,300
140.0	57	3,500	110	90	100	11,300
150.0	56	3,675	105	80	90	11,700
180.2	53	4,100	20	24	-	13,700
185.0	57	4,471	102	79	83	14,500
209.0	57	6,150	105	82	86	20,000
273.7	58	6,700	115	80	98	20,500
308.0	57	6,800	132	100	65	21,200
340.0	57	8,600	98	-	-	27,100
350.0	57	8,460	108	92	104	28,600

TABLE 1-39 (Cont.)

HNO_3 Rate 100% (Tons/day)	Normal Acid Strength (%)	NH_3 Feed Rate (lb/hr)	NH_3 Oxidation Pressure (psig)	Absorption Column Exit Press. (psig)	Absorption Column Exit Temp. (°F)	Tail Gas Flow Rate (Scfm)
Uncontrolled						
48.4	60.0	-	100	72	96	3,500
39.6	60.0	-	100	75	95	3,400
60.0	57.5	1,454	100	88	80	5,300
120.0	57.0	2,750	100	90	90	10,500
180.0	55.0	4,000	100	80	95	15,500
200.0	57.5	4,740	105	85	90	16,500
265.0	56.0	6,670	115	89	100	21,800

TABLE 1-40

TYPICAL EMISSIONS FROM NITRIC ACID PLANTS WITH CATALYTIC WASTE GAS TREATMENT EQUIPMENT [18]

HNO_3 Rate 100% acid (ton/day)	Tail Gas Flow Rate (Scfm)	Before Treatment			After Treatment			
		Temp. (°F)	Oxygen (vol %)	Total NO_x (vol %)[a]	Temp. (°F)	Oxygen (vol %)	Total NO_x (vol %)[a]	NO_x Emission of (lb/ton of acid produced, 100% acid basis)
139.7	10,900	840	3.5	0.21	1,250	-	0.04	5.7
120.0	9.300	375	2.4	0.10	915	0.0	0.0002	0.0
140.0	11,300	-	1.5	0.50	1,500	0.0	0.015	2.2
150.0	11,700	502	2.2	0.30	1,120	0.0	0.01	1.5
180.2	13,700	250	3.0	0.22	900	1.9	0.14	19.6
185.0	14,500	660	2.5	0.30	1,190	0.0	0.005	0.75
209.0	20,000	-	2.7	0.30	-	0.5	0.1	17.6
273.7	20,500	570	1.6	0.54		0.0	0.008	1.1
308.0	21,200	-	2.1	0.36	-	0.9	0.04	5.1
340.0	27,100	-	1.9	0.54	-	1.7	0.21	30.9
350.0	28,600	-	3.0	0.2	1,250	1.7	0.044	6.6

[a] Does not include N_2O.

TABLE 1-41

TYPICAL EMISSIONS FROM NITRIC ACID PLANTS WITHOUT WASTE GAS TREATMENT[18]

HNO_3 Rate 100% acid (ton/day)	Stack Gas Flow Rate (Scfm)	Oxygen Emission (vol %)	NO_x Emission (lb/ton 100% acid)
48.4	3,500	4.1	33
39.6	3,400	4.5	111
60.0	5,300	3.0	17
120.0	10,500	3.0	50
180.0	15,500	4.0	96
200.0	16,500	3.0	43
265.0	21,800	4.0	45

Based on 85 Scfm of Effluent
Per Daily Ton of Acid

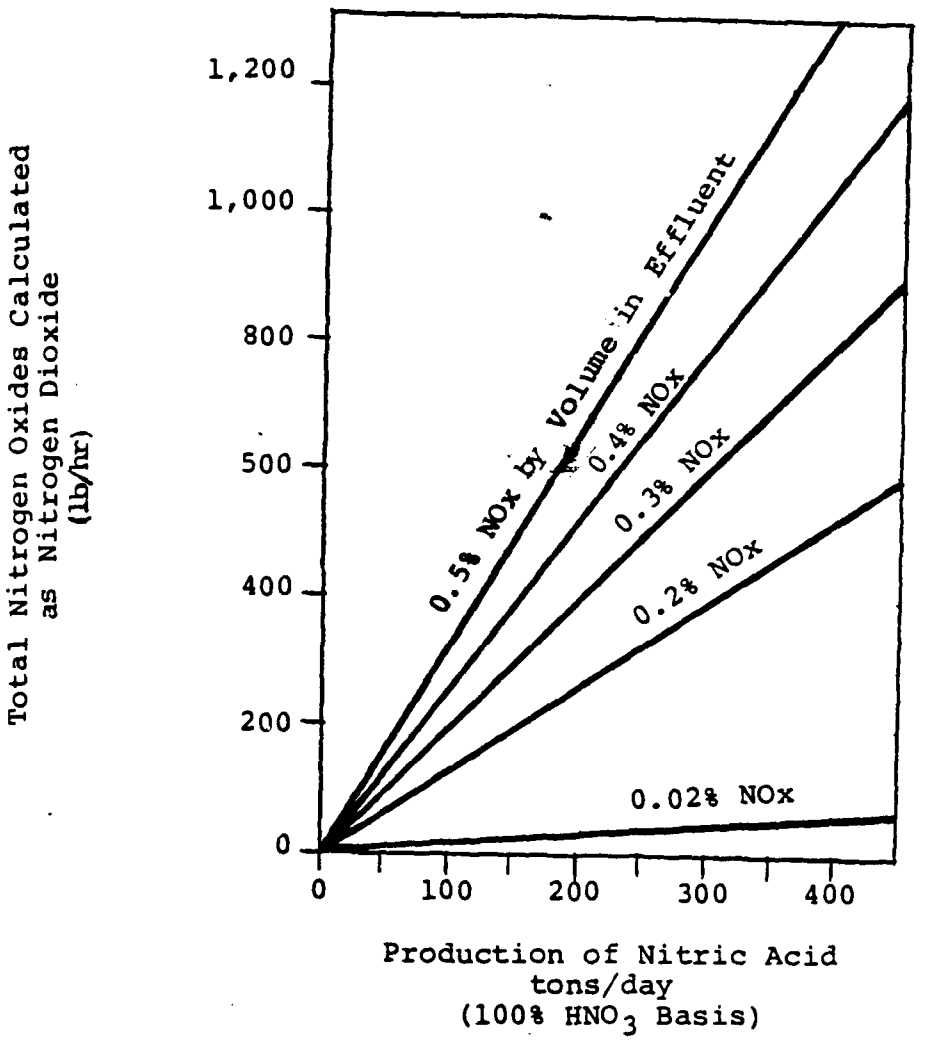


Figure I-14

TOTAL NITROGEN OXIDES (CALCULATED AS NITROGEN DIOXIDE) EMITTED PER HOUR VS. DAILY PRODUCTION OF NITRIC ACID[18]

one-half of the total nitrogen oxides. In the pressure process, the gases are generally reheated and discharged to the atmosphere at 400° to 500°F; any nitric acid mist present is then changed to its vapor state prior to discharge into the atmosphere.

1.3.3.3 Emission Factors

Nitrogen oxide emissions[3] (expressed as NO_2) are presented for weak nitric acid plants in Table 1-42. The emission factors vary considerably with the type of control employed, as well as with process conditions. For comparison purposes, the Environmental Protection Agency (EPA) standard for both new and modified plants is 3.0 pounds per ton of 100 percent acid produced (1.5 kilograms per metric ton), maximum 2-hour average, expressed as NO_2. Unless specifically indicated as 100 percent acid, production rates are generally given in terms of the total weight of product (water and acid). For example, a plant producing 500 tons (454 MT) per day of 55 weight percent nitric acid is really producing only 275 tons (250 MT) per day of 100 percent acid.

1.3.3.4 Control Systems[19]

The emission of nitrogen oxides may be reduced by catalytic reduction with certain fuels, by absorption, and by adsorption. Catalytic reduction is by far the most widely used method of abatement.

1.3.3.4.1 Catalytic Reduction

Catalytic reduction is particularly suited to the pressure ammonia oxidation process, in which the absorption tower tail gas is of uniform composition and flow, is under pressure, and can be reheated by heat exchange to the necessary reduction-system feed temperature. Efficiencies above 90 percent are possible,[19] and, in addition, a significant economic return can be realized through recovery of heat generated in the catalytic recovery unit.

Almost 50 percent of U.S.-built nitric acid plants employ this technology, using both one- and two-stage NOx reduction units. If a monolithic catalyst is used, pressure drop through the system is low. Capital costs are not excessive, and unit payback is possible by recovery of the heat generated by the tail-gas oxygen combustion. The plant thereby becomes self-sustaining, and in some cases a net exporter of energy. Ninety percent efficiency for NOx removal for a one-year catalyst life is obtainable.[19]

TABLE 1-42

TYPICAL NITROGEN OXIDE EMISSIONS FROM NITRIC ACID PLANTS[3]

Type of Control	Control Efficiency (%)	Emissions (NO_2)[a] lb/ton acid	kg/MT acid
Weak acid			
Uncontrolled	0	50-55[b]	25.0-27.5
Catalytic combustor (natural gas fired)	78-97	2-7[c]	1.0-3.5
Catalytic combustor (hydrogen fired)	97-99.8	0.0-1.5	0.0-0.75
Catalytic combustor (75% hydrogen, 25% natural gas fired)	98-98.5	0.8-1.1	0.4-0.55
High-strength acid	-	0.2-5.0	0.1-2.5

[a]Based on 100% acid production.

[b]Range of values taken from four plants measured at following process conditions: production rate, 120 tons (109 MT) per day (100% rated capacity); absorber exit temperature, 90°F (32°C); absorber exit pressure, 7.8 atmospheres; acid strength, 57%. Under different conditions, values can vary from 43 to 57 lb/ton (21.5 to 28.5 kg/MT).

[c]To present a more realistic picture, ranges of values were used instead of averages.

There are several disadvantages; the most notable is that optimum catalyst operating parameters limit absorber/acid production system operating flexibility. Natural gas prices (fuel for catalytic reduction) are also higher than in the past, or the gas may be simply unobtainable. If particulate catalysts (spheres, pellets) are used, pressure drops may be excessive, unless an expensive catalyst vessel design is used.

1.3.3.4.2 Extended Absorption

Both new and older plants have significant operating experience with the extended absorption system, which basically adds a second absorption tower to the plant. Additional acid is produced that can be credited to the production level. This system offers flexibility in absorber operation, and is not sensitive to production rate changes (within reasonable limits). Ninety percent NOx removal is routinely obtained.[19] The initial cost is not excessive, and the equipment life and maintenance is comparable to the first absorption tower. The two towers may also be physically combined.

The disadvantages include greater pressure drop and costs for utilities since two towers are used instead of one. The physical size may also make it difficult to retrofit older plants. No steam or power generation credit can be taken since heat is not in the process. The additional acid produced must be used, or a water pollution problem is generated.

1.3.3.4.3 Zeolite Adsorption

Acid-resistant molecular sieves (zeolites) can be used as a primary agent for NOx adsorption and recycling into the absorber. Installed nitric acid plant capacity using this system has now become significant. Reuse is made of the NOx tail gas without using a second absorber and, therefore, production credit can help pay for the unit. Sieve life exceeding one year can be guaranteed, and 90 percent NOx removal efficiency is routinely obtained.[19]

System flexibility (multiple adsorbers) can be built in to handle changes in plant operating conditions, but the system can be complex enough to cause maintenance problems. Pressure drops can be excessive, since multiple beds (silica gel, molecular sieve) are required in the adsorber gas train. The additional acid must be used and power/steam must be generated to produce it. Again, as with extended absorption, no steam and power generation credit is possible.

1.4 EVAPORATIVE LOSS SOURCES[20-22]

1.4.1 Metal Treating and Degreasing Operations[22]

1.4.1.1 Process Characterization

Before it becomes a finished product, metal goes through many steps, e.g., stamping, cutting, tapping, drawing, quenching, assembly, and finishing. These operations often deposit substances on the metal surfaces that must be removed before assembly or finishing operations such as enameling or painting. The focus from an emission control point is usually on the removal of greases for those metal finishing processes that require a surface ranging from relatively clean to very clean.

Degreasing or solvent metal cleaning employs non-aqueous solvents to clean the surface of metal articles. Metal work pieces are cleaned with organic solvents because water or detergent solutions exhibit a slow drying rate and have high electrical conductivity, high surface tension, a tendency to cause rusting, and a relatively low solubility for organic soils such as greases. A broad spectrum of organic solvents is available, such as petroleum distillates, chlorinated hydrocarbons, ketones, and alcohols. Although solvents may vary, there are basically three types of degreasers: (1) cold cleaners, (2) open top vapor degreasers, and (3) conveyorized degreasers.

1.4.1.1.1 Cold Cleaners

Cold cleaners are the simplest, least expensive, and most common type of degreaser. They are used for the removal of oil base impurities from metal parts in a batch-load procedure that can include spraying, brushing, flushing, and immersion. The cleaning solvent is generally at room temperature. Although it may be heated slightly, the solvent never reaches its boiling point. When parts are soaked to facilitate cleaning, it is not uncommon for the solvent to be agitated by pumps, compressed air, mechanical motion, or sonic waves.

There are several methods for materials handling in cold cleaning operations. Manual loading is used for simple, small-scale cleaning operations. Batch-loaded conveyorized systems are more efficient for complex,

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large-scale operations. They can be set to automatically lower, pause, and raise a workload. By dipping in a series of tanks, each with increasingly pure solvent or possibly a different solvent, a "cascade" cleaning system is established.

1.4.1.1.2 Open-Top Vapor Degreasers

The open-top vapor degreaser cleans by condensing vaporized solvent on the surface of the metal parts. The soiled parts are batch loaded into the solvent vapor zone of the unit. Solvent vapors condense on the cooler surface of the metal parts until the temperature of the metal approaches the boiling point of the solvent. The condensing solvent dissolves oil and grease, washing the parts as it drips down into the tank. Sometimes the cleaning process is modified with spraying or dipping.

To condense rising vapors and prevent solvent loss, the air layer or freeboard above the vapor zone is cooled by a series of condensing coils that ring the internal wall of the unit. Most vapor degreasers also have an external water jacket that cools the freeboard to prevent convection up hot degreaser walls. The freeboard protects the solvent vapor zone from disturbance caused by air movement around the equipment.

1.4.1.1.3 Conveyorized Degreasers

Conveyorized degreasers operate on the same principles as open top degreasers; the only difference is in materials handling. In conveyorized cleaners, parts may be dipped but manual handling is mostly eliminated. In addition, conveyorized degreasers are almost always hooded or covered.

There are many designs for conveyorized degreasers. These include monorail, cross-rod, vibra, ferris wheel, belt, and strip degreasers. Each conveying operation can be used with either cold or vaporized solvent. Conveyorized degreasers are used in a wide range of applications and are typically found in plants where there is enough production to provide a continuous stream of products to be degreased.

1.4.1.2 Operating Characteristics

Solvent concentrations in the air flow from solvent cleaners will vary considerably due to intermittent workloads. For example, open-top tank degreasers may operate only 25 percent of a shift, during which time the exhaust fan is always on. The expected concentration variation is from 50 to 1,000 ppm, with the average between 100 and 500 ppm.

Air flow rates range from 50 to 125 cfm per square foot of opening area. Temperatures of solvent degreasing emissions range from room temperature to 125°F.

1.4.1.3 Emission Characterization

Emissions occur due to evaporation from (1) solvent bath, (2) solvent carry-out, (3) agitation, (4) waste solvent evaporation, and (5) exhaust.

1.4.1.3.1 Solvent Bath

Solvent emissions resulting from bath evaporation include diffusion and convection losses. These losses are increased through failure to close the cover whenever parts are not being handled. Open-top vapor degreasers and conveyorized degreasers have a vapor/air interface at the top of the vapor zone. Here, evaporated solvent mixes with the air as a result of diffusion, drafts, and turbulence from parts being inserted and removed. Warm, solvent-laden air is carried upward to convection, and the solvent vapors diffuse into the room. Estimates for solvent diffusion emissions are 0.05 $lb/hr\text{-}ft^2$ (0.24 $kg/hr\text{-}m^2$), if no appreciable drafts cross the top of the tank. Conveyorized degreasers are normally enclosed, so convection and diffusion losses are minimized.

1.4.1.3.2 Solvent Carry-out

Carry-out emissions result from entrainment of liquid and vaporous solvent as clean parts are removed from the degreaser. This problem can be complicated by the shape of the part. Crevices and cupped portions may hold solvent even after the part appears to be dry. Carry-out emissions are usually the major emission from conveyorized degreasers because of the inherently large workload.

1.4.1.3.3 Agitation

Agitation of solvent in cold cleaners increases emissions. The extent of this increase depends on the use of a cover, the type of agitation, and adjustments to the agitation system. Emissions are normally insignificant if the cover is closed during agitation. However, if the cover is left open, emissions from all types of agitation are significant.

1.4.1.3.4 Waste Solvent

Solvent emissions due to spray evaporation are usually a problem in cold cleaners only. Increased emissions in open-top vapor degreasers are not a problem if sprays are kept below the condensing coil level. The amount of emissions will depend on the pressure and drop size of the spray, the volatility of the solvent, and the tendency to splash and overspray. Common practice is to keep the spray at a pressure less than 10 psig and in an unbroken fluid stream.

1.4.1.3.5 Exhaust Evaporation

Excessive exhaust emissions result when exhaust rates for open-top vapor degreasers and conveyorized degreasers are set too high. Disruption of the vapor/air interface can occur, causing solvent vapors to be carried out by the exhaust system. The average exhaust rate is 50 ft^3/min-ft^2 (15 m^3/min-m^2) of degreaser opening. However, this rate may be exceeded to comply with OSHA regulations on worker exposure levels. In any case, there should be a cover that closes beneath the exhaust intake vents to prevent withdrawal of solvent-laden vapor.

1.4.1.4 Emission Factors

Table 1-43 presents controlled and uncontrolled hydrocarbon emissions from degreasing operation.[3]

TABLE 1-43

TYPICAL EMISSION FACTORS FOR DEGREASING OPERATIONS

Type of Control	% Control	Metal Cleaned lb/ton
Uncontrolled	0	1.5
Refrigerated Cooling Coils	30-60	1.0-0.6
Use of Covers	25-40	1.1-0.9
Carbon Adsorption	40-70	0.9-0.5

1.4.1.5 Control Systems

The choice of techniques to control solvent cleaning emissions is relatively limited, and includes (1) condensation, (2) absorption, (3) incineration, and (4) carbon adsorption.

1.4.1.5.1 Condensation

Where low efficiency is acceptable, condensation of vapors is possible. Where air flow rates are low, vapor concentrations high, and the solvent expensive, condensation may be economical. In solvent cleaning, however, flows are normally too high and vapor concentrations too low for economical chilling operations.

1.4.1.5.2 Absorption

Absorption can also be used to control solvent emissions. However, for it to be effective, a scrubbing liquid (absorbent) must be available in which the solvent is either soluble or will react to form a less volatile compound. The absorbent in itself must not release undesirable vapors under operating conditions, and the solvent must be separable from the absorbent unless the mixture can be economically and safely wasted. For the halogenated hydrocarbons typically used in solvent cleaning, there are no known absorbent materials that meet the above criteria. Attempts have been made to use mineral oil to collect trichloroethylene vapors, but emissions from the mineral oil itself have been excessive.

1.4.1.5.3 Incineration

Incineration is another potential control technique. However, the halogens will burn to compounds such as phosgene, hydrochloric acid, and hydrofluoric acid, and unless these are removed in a second-stage collector, they may become a more serious source of air pollution than the original solvent vapors. Additional disadvantages of incineration are the loss of recovered-solvent benefits and the high fuel requirement for burning air streams with low concentrations of combustible vapors. Incineration, therefore, is generally too expensive for the low solvent concentrations from degreasers and even more so for secondary control of incinerator halogen emissions. It is not often a serious contender for degreaser emission control.

1.4.1.5.4 Carbon Adsorption

Adsorption, almost universally using activated carbon beds, is the most effective control system for halogenated and petroleum solvent cleaning emissions. The carbon adsorption technique uses regenerative activated carbon to remove gaseous molecules from an air stream in three distinct phases: adsorption, desorption, and disposal or recovery of the adsorbed material.

1.4.2 Metal Coating Operations

1.4.2.1 Process Characterization

Surface coating operations primarily involve the application of paint, varnish, lacquer, or paint primer for decorative or protective purposes. The basic processes used for coating include (1) spraying, (2) dip coating, (3) flow coating, (4) coil coating, and (5) masking.

1.4.2.1.1 Spraying

Typical spraying operations are performed in a booth, with a draft fan to prevent explosive or toxic concentrations of solvent vapors. Essentially, there are three spraying techniques (1) airless atomization, (2) air atomization, and (3) electrostatic spraying.

1.4.2.1.1.1 Airless Atomization

Airless spraying atomizes without air by forcing the liquid material through specifically designed nozzles under a pressure of 1,000 to 2,000 psi. On release to atmospheric pressure, some of the solvents in the surface coating vaporize and join with the straight hydraulic forces at the nozzle as atomizing agents. In general, with airless spraying, less solvent is volatilized in the spray booth than with air spraying, meaning that more solids may have to be removed later during air drying or baking. Emissions from airless atomization are similar to the solvent formula.

1.4.2.1.1.2 Air Atomization

Air atomization uses its own source, which may be heated, filtered and/or humidified, or treated in some other fashion. It is based on partial volatilization of the solvent blend and is likely to produce emissions high in low boiling compounds.

1.4.2.1.1.3 Electrostatic Spraying

Electrostatic spraying projects charged coating particles into an electrostatic field created by a potential difference of about 100,000 volts between the articles sprayed and spray grids 12 inches away. The

particles of wet paint from the spray gun enter this field with the same potential as the grids and are thus repelled by them and attracted to the article being sprayed.

1.4.2.1.2 Dip Coating

In dip coating operations, the object is immersed for the required time in a tank of the coating. When the object is removed, excess coating drains back into the tank, either directly or via a drain ramp.

1.4.2.1.3 Flow Coating

Articles that cannot be dipped due to their buoyancy, such as pressure bottles, are subjected to flow coating. Material is fed through overhead nozzles in a steady stream over the article. Excess coating drains by gravity from the coated object and is recirculated. Removal of excess coating material and solvents is aided by jets of heated air.

1.4.2.1.4 Coil Coating

Coil coating is a technique for applying finish to long flat strips or coils of metal, on one side or both, by means of rollers similar to those in a printing press. Three power-driven rollers are normally used. One of the rollers is partially immersed in the coating material. The coating is then transferred by direct contact to a second parallel roller. The object to be coated is run between the second and third rollers and is coated by the second roller.

1.4.2.1.5 Masking

Masking is a technique for applying coatings where sharp, clean edges are needed; for instance, for lettering, stripping, and two-color finishes. The areas to be left uncoated are masked with cloth, plastic sheeting, tape, or a special mask derived by photography from an artwork pattern (silk screening).

The coating may then be applied by stencil or rubber squeegee. Masking is usually removed while the coating is still wet to prevent frayed edges and to ensure sharpness.

1.4.2.2 Operating Characteristics

Emissions of hydrocarbons occur in surface-coating operations because of the evaporation of the paint vehicles, thinners, and solvents used to facilitate the application of the coatings. The major factor affecting these emissions is the amount of volatile matter contained in the coating. The volatile portion of most common surface coatings averages approximately 50 percent, and most, if not all, of this is emitted during the application and drying of the coating. The compounds released include aliphatic and aromatic hydrocarbons, alcohols, ketones, esters, alkyl and aryl hydrocarbon solvents, and mineral spirits.

Solvent concentrations in spray booth effluents vary from 100 to 200 ppm for manual operations. Solvent emissions from spray booth stacks vary with the extent of the operation, from less than 1 pound to more than 3,000 pounds per day. No definitive data are available for automatic spray booths.

Virtually all solvents evaporate in the course of the coating sequence, each at its own rate. For measuring purposes, this evaporation is viewed in terms of "flash-off," defined as the quantity of solvent evaporated under either ambient or forced conditions from the surface of a coated object during a specific time. The graph in Figure 1-15 shows flash-off times for various coating types[11] applied by spraying and is useful for determining potential emissions from different coating systems. Percent of total solvent evaporated is plotted against time for various coating systems. For a given system, the flash-off at a specific time can be easily read from the graph. The total emission load, however, is significantly affected by the size, shape, and number of pieces being coated.

Generally, emission levels are increased by overspray, i.e., material that misses the surface to be coated. Table 1-44 gives overspray percentages for various spray techniques.[11]

1.4.2.3 Emission Characterization

Emissions of solvent vapor vary not only with the coating formula, but also with each individual processing step.

1.4.2.3.1 Spraying

Paint-spray booths generally have one side open; ventilation of the booth is necessary to ensure both operator and plant safety. Normal spray-booth ventilation velocities of from 100 to 150 feet per minute per square foot of booth opening are adequate for manual operations.

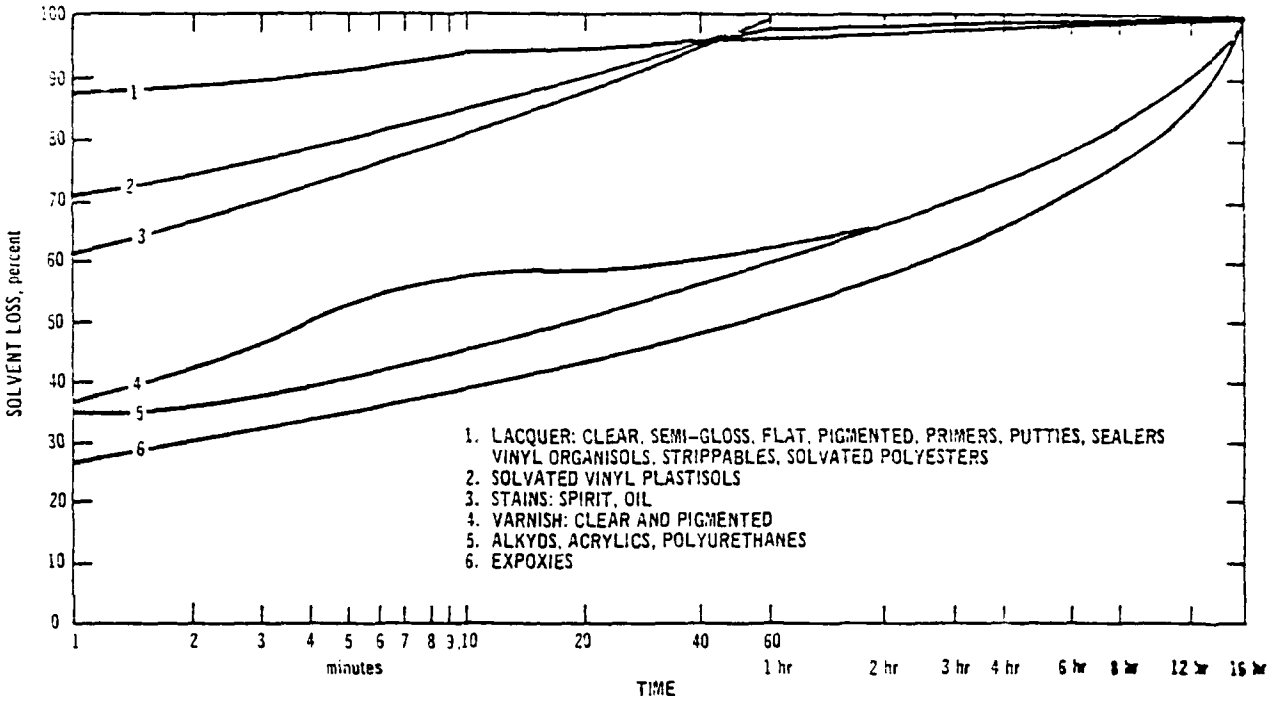


FIGURE 1-15
EVAPORATION OF VARIOUS FORMULAS[11]

TABLE 1-44

OVERSPRAY PERCENTAGES AS A FUNCTION OF SPRAYING METHODS AND SURFACES SPRAYED[11]

Method of Spraying	Flat Surfaces (%)	Table Leg Surface (%)	Bird Cage Surface (%)
Air atomization	50	85	90
Airless	20-25	90	90
Electrostatic			
Disc	5	5-10	5-10
Airless	20	30	30
Air Atomized	25	35	35

OSHA standards require that the average velocity over the open face of the booth be not less than about 1.5 feet per second. All fumes should be vented through a fume hood.

Discharge from a paint-spray booth consists of particulate matter and organic solvent vapors. The particulate matter consists of entrained coating material that did not adhere to the object being painted or to the inside surface of the booth. The organic vapors are generated from the evaporation of solvents, resins, diluent, or thinner.

1.4.2.3.2 Other Application Techniques

Emissions from other application techniques such as flow coating, dip coating, or coil coating differ from spray coating emissions to the extent that these methods require less coating material. However, the expected solvent emission load from these techniques can vary widely.

In fact, flow coating may not be much better from an emission standpoint than spray coating. For flow coating, the proper percentage of solids and correct viscosity must be maintained. Further, so much solvent is lost during recirculation and air blow-off of excess coating that flow coating is often done in a "tunnel" to keep the solvent-laden air in a fixed area. The result is that a well-run flow coating operation using 60,000 gallons of coating per year may use as much as 54,000 gallons of makeup solvent to compensate for "tunnel solvent" losses. This is much more wasteful than an air-atomized spray operation within 50 percent overspray.

Dip coating solvent losses are generally under 10 percent, depending on time of year and temperature in the plant. This usually represents much less solvent loss than that occurring with spraying or flow coating and does not normally require much makeup solvent.

From the standpoint of overall emissions, the single most efficient coating method is roller or coil coating, a process in which extraneous evaporation is practically negligible, since all coating supplied to the coating head is placed onto the web to be coated.

1.4.2.4 Emission Factors

Table 1-45 presents emission factors for surface coating operations.[3]

1.4.2.5 Control Systems[20]

Particulate matter can be effectively removed (50 to 98 percent) by techniques to control the particulate emissions. These techniques are described in this section.

TABLE 1-45

GASEOUS HYDROCARBON EMISSION FACTORS FOR SURFACE-COATING OPERATIONS[3]

Type of Coating	Emissions[a] (lb/ton)
Paint	1,120
Varnish and shellac	1,000
Lacquer	1,540
Enamel	840
Primer (zinc chromate)	1,320

[a] Reported as undefined hydrocarbons, usually organic solvents, both aryl and alkyl. Paints weigh 10 to 15 pounds per gallon (1.2 to 1.9 kilograms per liter); varnishes weigh about 7 pounds per gallon (0.84 kilograms per liter).

1.4.2.5.1 Dry Baffle

In the dry baffle method, the wet overspray collects on large panels called baffle plates, which catch 50 to 90 percent[20] of the particulates produced by spraying a high-solids enamel. With low-solids lacquers containing highly volatile solvents, efficiencies may be much lower due to the rapid drying of the lacquer and poor adhesion of dry particles to the baffle.

1.4.2.5.2 Paint Arrestor

Filter pads used in the paint arrestor method can remove up to 98 percent of paint particulates.[20] Filtering velocities should be less than 4 ft/sec.

1.4.2.5.3 Water Wash

Water curtains and sprays are 95 percent effective in removing paint particulates.[20] A water circulation rate of 0.1 to 5 gallons per cubic foot of exhaust air is usually recommended. Surfactants may be added to the water to aid in removing paint from the circulating tank.

In order of effectiveness, the paint arrestor would be considered the best technique for removing particulates when downstream solvent vapor processes such as catalytic or other afterburners, heat exchangers, or carbon absorption beds are used. Water washing to remove particulates would be a second choice, assuming that the solvent vapor processes can tolerate some water in the vapor stream. Baffle plates would be considered the third and least effective method, although by far the cheapest.

Control of the gaseous emissions can be accomplished by the use of adsorbers (activated carbon) or afterburners. The collection efficiency of activated carbon has been reported at 90 percent or greater. Water curtains or filter pads have little or no effect on escaping solvent vapors.

1.4.3 Liquid Fuel Loading and Unloading

1.4.3.1 Process Characterization

Tank truck and rail car terminals for the loading and unloading of crude oil, petroleum products, and organic chemicals are located at the end of pipelines or near refineries, marine terminals, and chemical plants. Bulk plants are secondary distribution facilities receiving and distributing product by tank trucks. A major source of hydrocarbon and organic chemical emissions from tank truck

terminals, rail car terminals, and bulk plants is loading operations.

The volume of vapors produced during the loading operation, as well as their composition, is greatly influenced by the type of loading or filling employed. The types in use throughout the industry may be classified under two general headings, top loading (or overhead loading) and bottom loading.

1.4.3.1.1 Top Loading

Top loading is divided into open top, with and without vapor recovery, and top-tight submerged fill. Open top involves loading of products into the compartment via the manhole located on top of the tank. Gasoline can be loaded directly into the compartment through a top loading head (splash fill). Attachment of a fixed or extensible downspout to the loading head provides a means of introducing the product near the bottom of the tank (submerged fill). A deflector at the outlet of the downspout provides for uniform spreading of the product and eliminates both static buildup and product splash.

In addition to submerged fill, the top loading head can be designed for vapor recovery. This top loading vapor head must be compatible with the truck hatch opening, and a vapor-tight seal is required between the head and the hatch to minimize vapor leakage during transfer of product.

These top loading systems, which require opening of the hatch (Figure 1-16), must be equipped with a separate vapor recovery system for delivery to customers with vapor balance systems.

Provisions for top-tight submerged fill have been installed in tank trucks primarily in Texas. This installation permits the loading of product through a vapor-tight loading adaptor mounted on top of each compartment (Figure 1-17) and attached to a submerged fill pipe. For vapor recovery, the vapor spaces of each compartment are manifolded to the overturn rail or to a vapor return line.

One advantage of this permanently affixed top-tight submerged fill is that the hatch/dome covers remain closed at all times except for cleanup and repair. No vapor loss is, therefore, attributable to opening of the hatch. The top-tight and vapor-head system can collect vapors expelled from the tank during product loading.

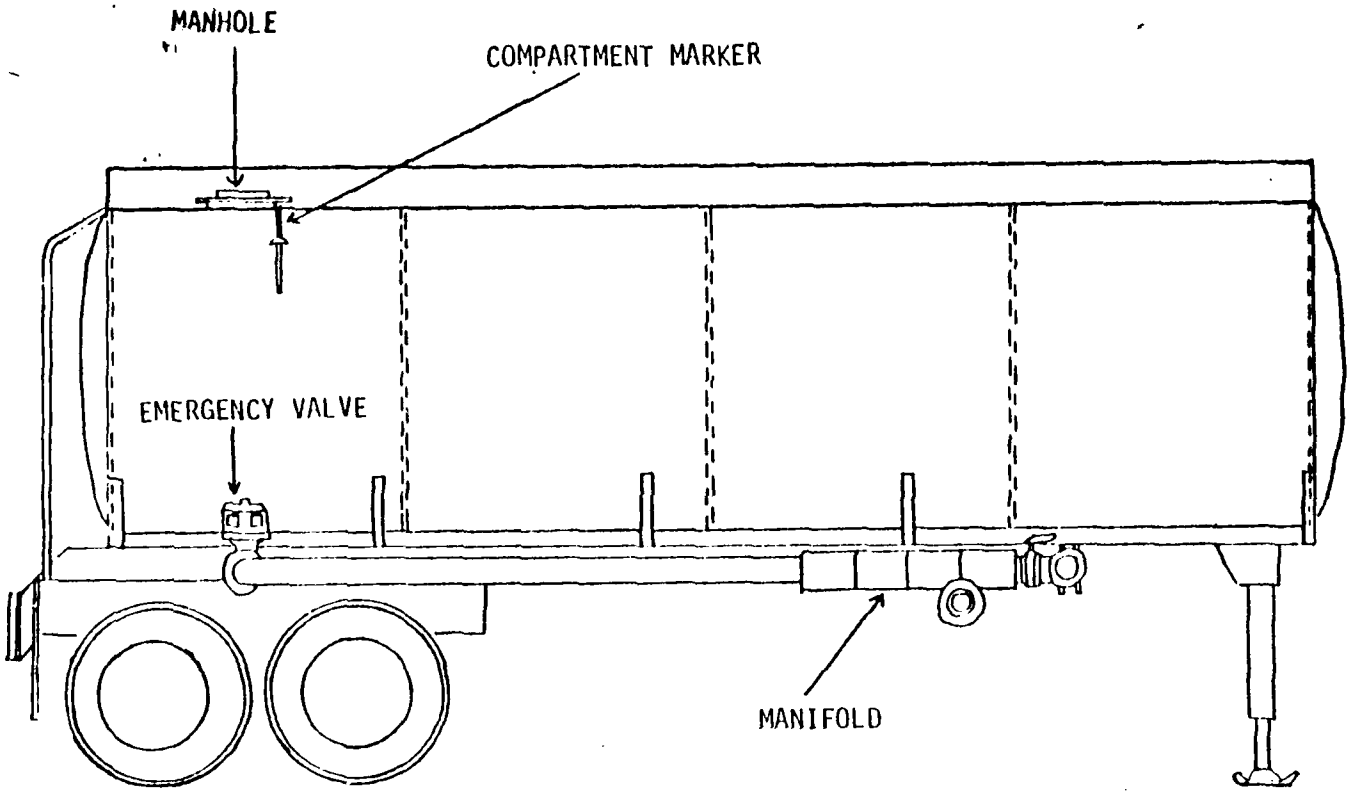
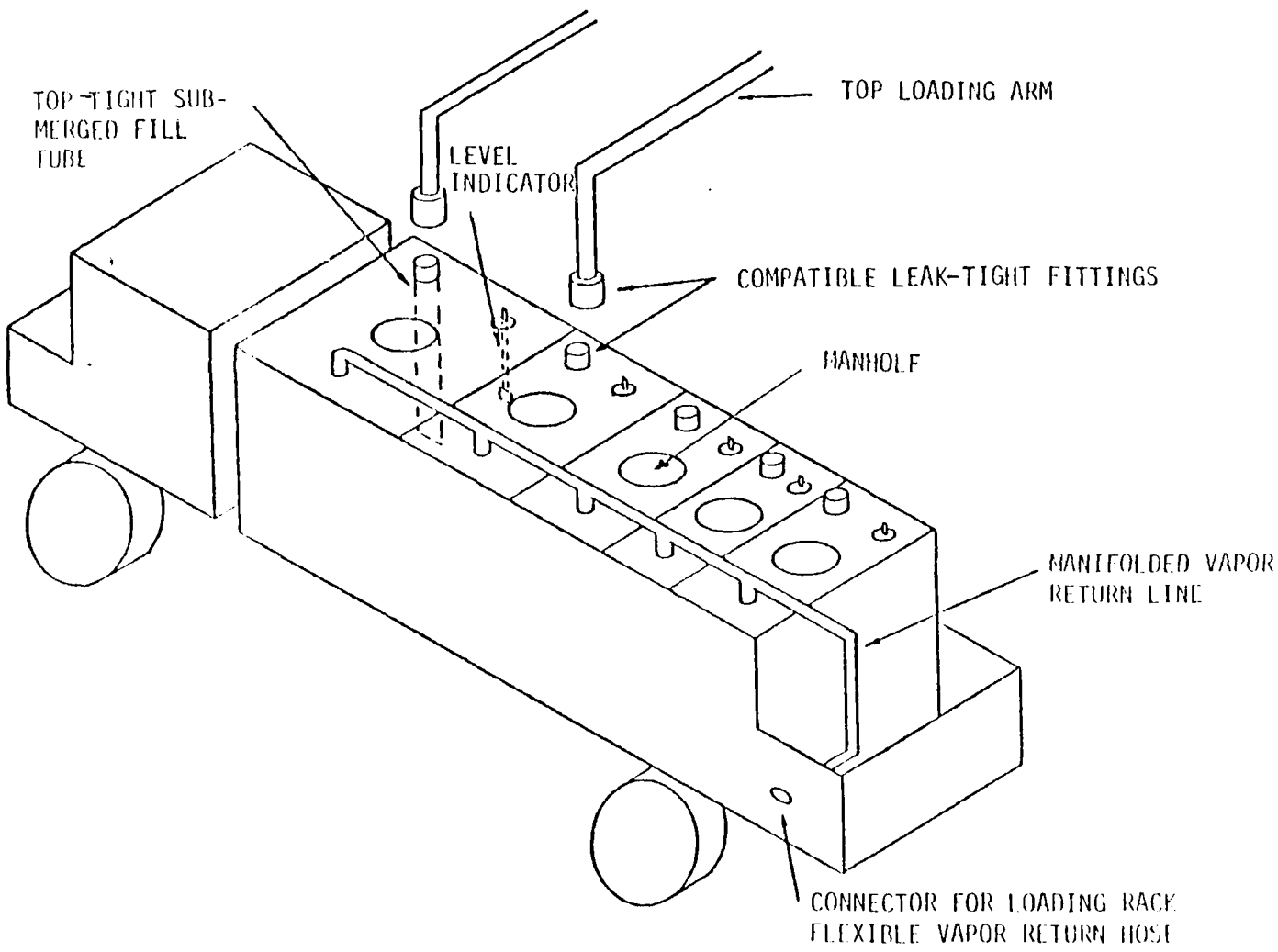


FIGURE 1-16

TOP-LOADING TANK

FROM: "SURVEY OF GASOLINE TANK TRUCKS AND RAIL CARS", U.S. EPA-450/3-79-004 MARCH 1979.


TOP-TIGHT SUB-
MERGED FILL
TUBE
LEVEL
INDICATOR
TOP LOADING ARM
COMPATIBLE LEAK-TIGHT FITTINGS
MANHOLE
MANIFOLDED VAPOR
RETURN LINE
CONNECTOR FOR LOADING RACK
FLEXIBLE VAPOR RETURN HOSE

1.4.3.1.2 Bottom Loading

Bottom loading permits the operator to stand on the ground for loading the tank through connectors at the side of the tank. Since the discharge opening of the tank is used for loading, submerged fill occurs naturally. Some of the advantages cited for bottom loading are (1) improved safety, (2) faster loading, and (3) reduced labor costs. Off-loading and on-loading adaptors, a single valve for both in- and out-service or Y-valves, must be provided for individual compartment loading and unloading (Figure 1-18). In addition, an emergency or internal valve is required as well as a vent valve when gasoline is being loaded. A means to prevent gasoline spraying into the tank must be included, and a liquid high level sensor is necessary for secondary automatic shutoff purposes. Tanks with bottom loading provisions can normally be top loaded from an open hatch.

1.4.3.2 Operating Characteristics

When a compartment of a tank vehicle or tanker is filled through an open overhead hatch or bottom connections, the incoming liquid displaces the vapors in the compartment to the atmosphere. These vapors consist of a mixture of air and hydrocarbons depending upon the product being loaded, the temperature of the product, and the type of loading. Ordinarily, but not always, when gasoline is loaded, the hydrocarbon concentration of the vapors is from 30 to 50 percent by volume and consists of gasoline fractions ranging from methane through hexane. Table 1-46 shows a typical analysis of the vapors emitted during the loading of motor gasoline into tank vehicles[11]. On the basis of a typical 50 percent splash filling operation, vapor losses from the overhead filling of tank vehicles with gasoline have been determined empirically to amount to 0.1 to 0.3 percent of the volume loaded.

In addition to the emissions resulting from the displacement of hydrocarbon vapors from the tank vehicles, additional emissions during loading result from evaporation of spillage, drainage, and leakage of product.

1.4.3.3 Emission Characterization

Emissions from loading tank trucks and rail cars can be estimated within 30 percent using the expression:

$$L_L = K \frac{SPM}{T}$$

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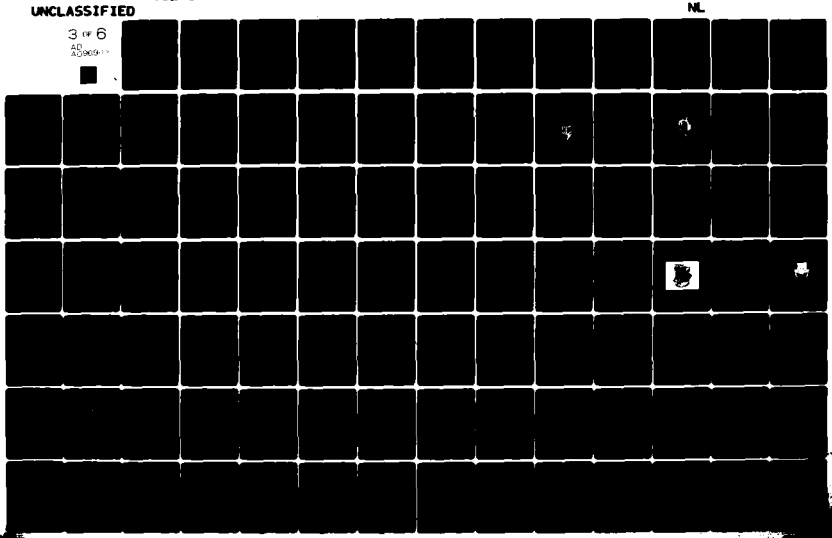

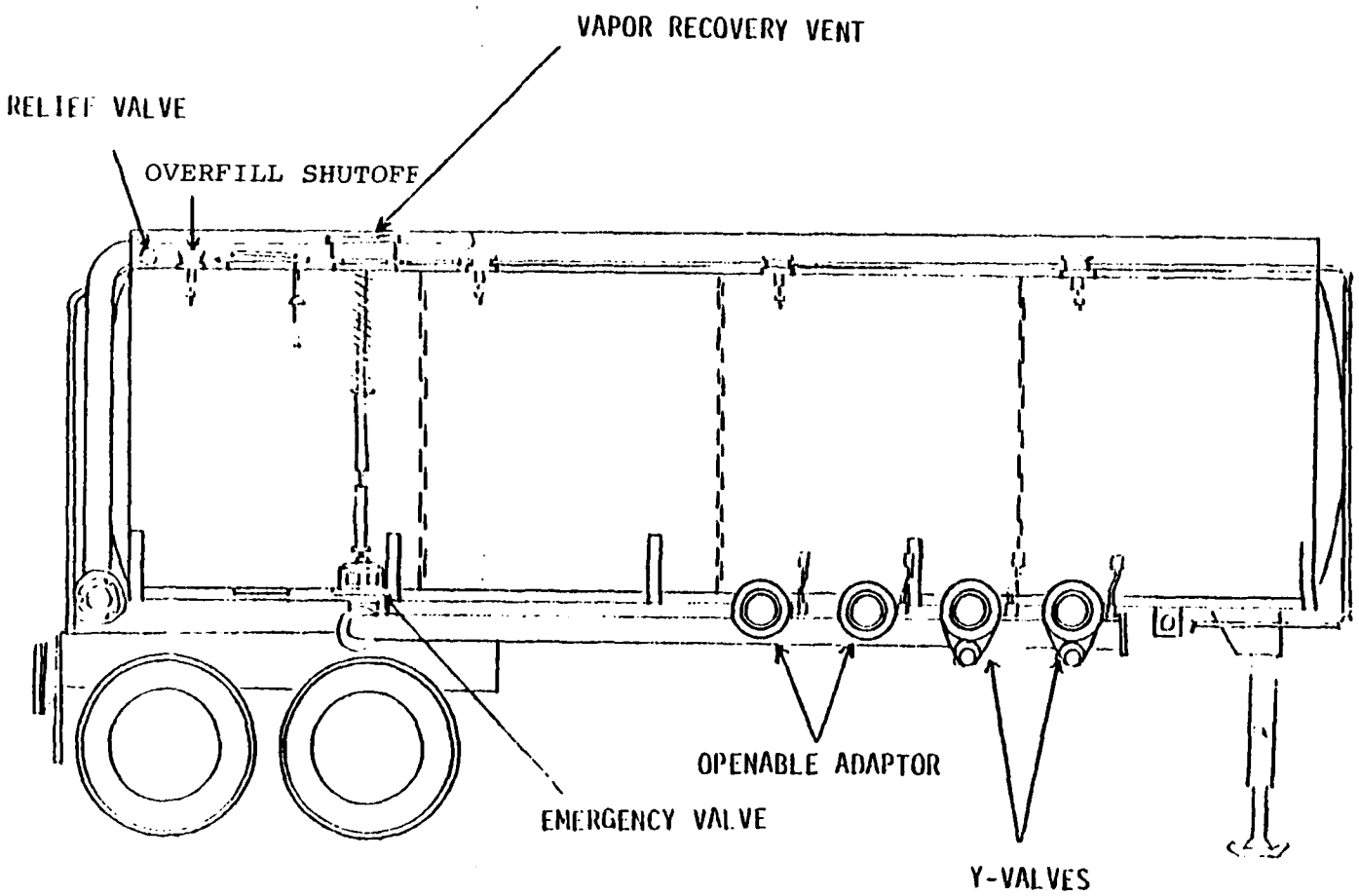

FIGURE 1-18

BOTTOM-LOADING TANK

FROM: "SURVEY OF GASOLINE TANK TRUCKS AND RAIL CARS", U.S. EPA-450/3-79-004, MARCH 1979.

TABLE 1-46

TYPICAL ANALYSIS OF VAPORS FROM BULK LOADING OF GASOLINE INTO TANK TRUCKS[11]

Fraction	Vol %		Wt %	
Air		58.1		37.6
Hydrocarbon				
Propane	0.6		0.6	
Iso-Butane	2.9		3.8	
Butene	3.2		4.0	
N-Butane	17.4	41.9	22.5	62.4
Iso-Pentane	7.7		12.4	
Pentene	5.1		8.0	
N-Pentane	2.0		3.1	
Hexane	3.0		8.0	
		100.0		100.0

where

L_L = loading loss, kg/m^3 of liquid loaded ($lb/10^3$ gal)

K = constant, 12.04×10^{-5} kg-mole, °K/Pa-m^3 (12.46 lb-mole, °R/psia-10^3 gal)

M = molecular weight of vapors, kg/kg-mole (lb/lb-mole)

P = true vapor pressure of liquid loaded, Pa (psia)

T = bulk temperature of liquid loaded, °K (°R)

S = a saturation factor.

The saturation factor (S) represents the expelled vapor's fractional approach to saturation and accounts for the variations observed in emission rates for different loading methods.

Table 1-47 lists suggested saturation factors for tank truck and rail car loading.[22] The emission factor for hydrocarbon emissions generated during submerged-fill (top or bottom) gasoline loading operations is 600 mg hydrocarbons emitted per liter of gasoline loaded (5 $lb/10^3$ gal). This figure represents 40 to 50 percent hydrocarbon saturation of the air in the tank trucks.

Processing Unit Emissions

Figure 1-19 is a schematic diagram of gasoline loading with vapor processing (control) unit and vapor return line. Hydrocarbon mass emissions from such a system can be determined directly using flow meters and hydrocarbon analyzers, as presented below:

i. Volume of air-hydrocarbon mixture exhausted from the processing unit:

$$V_e = V_{ef} - V_{ei} \ (m^3)$$

where V_e = totalized volume from flow rate and time records

subscript f = final

subscript i = initial.

TABLE 1-47

S FACTORS FOR CALCULATING TANK TRUCK AND RAIL CAR LOADING LOSSES[22]

Cargo Carrier	Mode of Operation	S Factor
Tank Trucks and Tank Cars	Submerged loading of a clean cargo tank	0.50
	Splash loading of a clean cargo tank	1.45
	Submerged loading: normal dedicated gasoline service	0.60
	Splash loading: normal dedicated gasoline service	1.45
	Submerged loading: dedicated gasoline balance service	1.00
	Splash loading: dedicated gasoline balance service	1.00

Notes: (i) A cargo carrier in normal dedicated gasoline service transports only gasoline. Tanks are not cleaned or vented between trips and retain a significant concentration of vapors generated by evaporation of residual gasoline product.

(ii) A cargo carrier in dedicated gasoline balance service picks up vapors displaced during unloading operations and transports these vapors in the cargo tank back to the loading terminal.

(iii) Splash loading and submerged loading are described in Section 1.4.3.1.

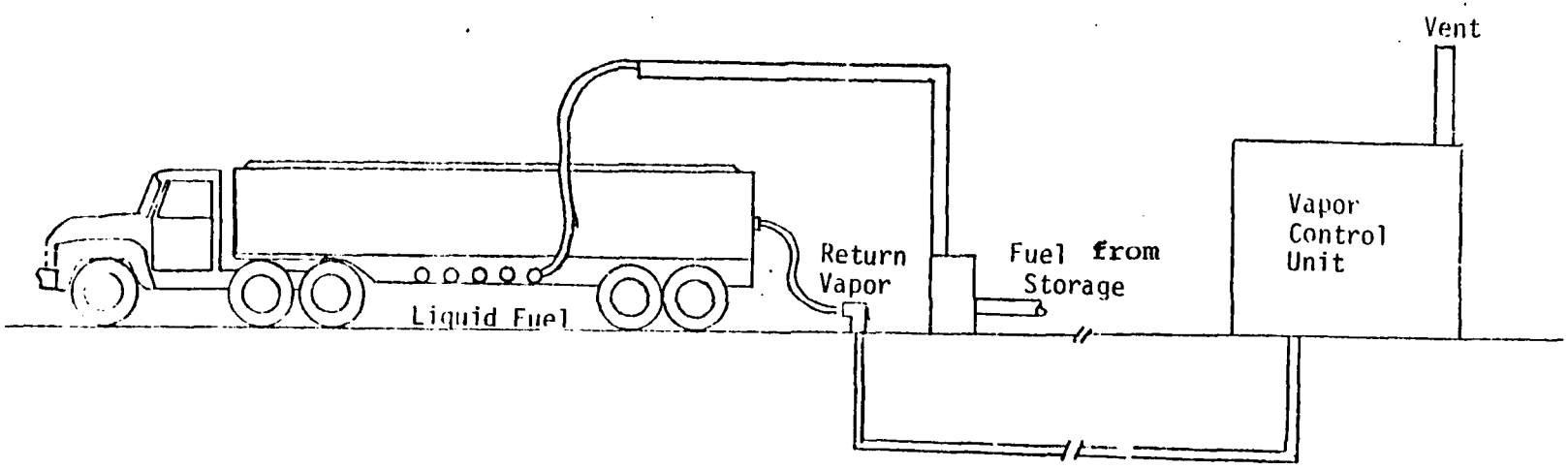

FIGURE 1-19

TANK TRUCK GASOLINE LOADING VAPOR CONTROL SCHEMATIC

From: "CONTROL OF HYDROCARBONS FROM TANK TRUCK GASOLINE LOADING TERMINALS", U.S. EPA-450/2-77-026, October 1977.

ii. Normalized volume of exhausted mixture:

$$V_{es} = \frac{(0.3858\ ^\circ K/mm\ Hg)\ V_e P_e}{(T_e + 273.2)}$$

where V_{es} = normalized volume of air-hydrocarbon mixture exhausted, NM^3 at 20°C, 76°mm Hg

P_e = pressure at processing unit exhaust, mm Hg abs

T_e = temperature at processing unit exhaust, °C.

iii. Mass of hydrocarbons exhausted from the processing unit:

$$M_e = (1.833 \times 10^6 \frac{mg\ C_3H_8)}{NM^3\ C_3H_8} \times V_{es} C_e\ (mg)$$

where C_e = volume fraction of hydrocarbons in exhausted mixture (volume % as $C_3H_8/100$, corrected for methane content if required.)

iv. Average processing unit emissions:

$$(M/L)_e = \frac{\Sigma M_e}{L_t}\ (mg/liter)$$

where $(M/L)_e$ = mass of hydrocarbons exhausted from the processing unit per volume of liquid loaded, mg/l

L_t = total volume of liquid dispensed from all controlled racks during the test period, liter.

Note: This method is applicable only to motor tank truck and trailer loading terminals employing vapor balance collection systems (discussed in Section 1.4.3.4) and either continuous or intermittent vapor processing devices.

1.4.3.4 Control Systems

Emission control technology for tank truck and rail car loading includes the use of modified loading techniques, vapor recovery units, and the balance system. A 40 to 60 percent[22] reduction in emissions can be achieved by the conversion of loading procedures from splash loading to bottom loading. This conversion requires moderate piping modifications to both the cargo carrier and the loading rack.

If bottom loading is practiced in conjunction with the application of a vapor recovery system, the emissions from tank truck and rail car loading operations can be reduced 90 to 98 percent.[22] In a properly operating vapor recovery system, vapors displaced from the cargo tanks during product loading are collected in a vapor header on a cargo carrier and conveyed to a vapor recovery unit. Through processes such as refrigeration, condensation, adsorption, or absorption, the vapors are recovered as liquid product. Occasionally, incineration and catalytic combustion systems are used to dispose of loading vapors.

The vapor balance system is an additional vapor control technique applicable only to facilities such as bulk plants that also receive their products by tank or rail car. In the vapor system, vapors displaced from the cargo tanks during product loading are collected in a vapor header on the cargo carrier and conveyed to a vapor recovery unit. The recovered liquid product is returned to storage. This "balanced" exchange occurs because the volume of displaced vapors is approximately equal to the volume of liquid cargo transferred.

The control efficiency of the balance system has been demonstrated to range from 90 to 100 percent.[22]

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CHAPTER 2

CONTROL EQUIPMENT FOR PARTICULATE POLLUTANTS

Air pollution control equipment may be classified into three groups: (1) equipment controlling particulate matter, (2) equipment controlling gaseous emissions, and (3) equipment controlling both gaseous and particulate emissions. From an air pollution viewpoint, particulate matter is any material that exists as a solid or liquid at standard conditions. Some examples of particulates are smoke, dusts, fumes, mists, and sprays.

Devices for control of particulate matter are available in a wide variety of designs using various principles of operation and having a wide latitude in collection efficiency, initial cost, operating and maintenance costs, space, arrangement, and materials of construction. In selecting the optimum device for a specific job, it is necessary to consider many factors. Consider the following factors significant:

1. Particulate characteristics, such as particle size range, particle shape, particle density, and physiochemical properties such as agglomeration tendencies, stickiness, inflammability, toxicity, and electrical conductivity.

2. Carrier gas characteristics, such as temperature, pressure, humidity, density, viscosity, dew points of condensable components, electrical conductivity, corrosiveness, inflammability, and toxicity.

3. Process factors, such as volumetric gas rate, particulate concentration, variability of material flow rates, and collection efficiency requirements.

4. Operational factors, including structural limitations such as head room, and floor space, and equipment material limitations such as pressure, temperature, and corrosion service requirements.

In this chapter, devices for control of particulate matter have been grouped into four classes:

(1) centrifugal separators
(2) wet collection devices
(3) electrostatic precipitators
(4) fabric filters.

2.1 STEP-BY-STEP DESIGN REVIEW PROCEDURE

As mentioned earlier, the particulate control equipment have been classified into four groups. For each of these groups, design guidelines have been developed, and are presented as a logical sequence of steps in this section. Each step refers to sections, figures and/or equations within this chapter, which should be used to implement that particular step. The detailed design of any equipment for a given application can be conveniently carried out by following the step-by-step procedure recommended in this section.

2.1.1 Step-By-Step Design Review Procedure for Centrifugal Separators

		Reference
1.	Estimate or calculate diameter of cyclone, D_c, for a given flow rate.	Fig. 2-6,2-7, 2-8
2.	Determine cyclone inlet width, B_c, from Step 1 and calculate cut size particle diameter, D_{pc}, or particle collected at 50% efficiency.	Sect. 2.2.3 Eq. 2.1
3.	Apply to D_{pc} corrections for velocity, viscosity, width/diameter, or number of turns, if necessary.	Fig. 2-10, 2-11
4.	Determine fractional efficiency of proposed cyclone for average particle diameter, dp, for each size range of inlet particle size distribution.	Fig. 2-13
5.	Using information from Step 4, calculate overall collection efficiency of cyclone.	Sect. 2.2.3
6.	Compare overall collection efficiency with desired efficiency; if η is too low, return to Step 1 and re-estimate a smaller diameter D_c. If collection efficiency equals or exceeds desired η, proceed to Step 7.	
7.	Calculate remaining cyclone dimensions based on D_c from Step 1. Calculate cyclone inlet area for cost evaluation.	Fig. 2-6,2-7 2-8
8.	Calculate pressure drop of cyclone.	Sect. 2.2.3
9.	Estimate capital and operating costs of cyclone.	Sect. 4.5.4 and 4.6.1.1

Note: If dimensions of cyclone are known, use Steps 2 through 8 to determine efficiency and pressure drop.

2.1.2 Step-by-Step Design Review Procedures for Wet Scrubbers

		Reference
1.	Calculate required number of transfer units, N_t, for desired efficiency, η.	Eq. 2.19
2.	Determine characteristic parameters α and β for specific application.	Table 2-3
3.	Calculate total power required, P_T, for desired efficiency.	Eq. 2.20
4.	Select liquid to gas ratio, L/G, and water pressure into scrubber, p_L.	Consult with manufacturer
5.	Calculate contacting power based on liquid stream energy input, P_L.	Eq. 2.17
6.	Calculate contacting power based on gas stream energy input, P_G.	Eq. 2.18
7.	Calculate scrubber pressure drop, Δp.	Eq. 2.16
8.	Estimate capital and operating costs of wet scrubber.	Sect. 4.5.3, 4.5.9, and 4.6.1.4

2.1.3 Step-by-Step Design Review Procedure for Electrostatic Precipitators

		Reference
1.	Characterize gas stream: coal sulfur content, resistivity, temperature, flow rate, etc.	Engineering Evaluation
2.	Select precipitation rate parameter.	Sect. 2.4.3.1
3.	Determine the desired collection efficiency.	Sect. S.3
4.	Determine required specific collection area (SCA) using either:	
	(a) Deutsch equation	Sect. 2.4.3.1
	(b) Collection efficiency vs. SCA and precipitation rate parameter	Fig. 2-26
	(c) Collection efficiency vs. SCA and sulfur content (boilers)	Fig. 2-27
	(d) Precipitation rate parameter vs. corona power density.	Fig. 2-28
5.	Calculate power required to energize precipitator.	Fig. 2-29
6.	Determine degree of sectionalization or number of bus sections.	Sect. 2.4.3.4
7.	Remaining design parameters, such as number of rappers, rapping frequency, and rapping force, are usually specified on the basis of past experience.	Consult the manufacturer
8.	Estimate capital and operating costs of precipitator.	Sect. 4.5.1 and 4.6.1.2

2.1.4 Step-by-Step Design Review Procedure for Fabric Filters

		Reference
1.	Determine continuous or intermittent operation/pressurized or suction baghouse.	Sect. 2.5.3.1 and 2.5.3.2
2.	Select most suitable type of fabric and method of cleaning for proposed application.	Sect. 2.5.2 and 2.5.3.10
3.	Select appropriate air-to-cloth ratios.	Sect. 2.5.3.9
4.	Calculate required fabric area to adequately treat design flow without exceeding recommended filtering velocity.	Sect. 2.5.3.11
5.	Determine degree of compartmentalization.	Sect.2.5.3.11 and 2.5.3.13
6.	Maximum operating pressure drop across baghouse and time interval between cleanings are optimized after system is operational.	Sect. 2.5.3.8
7.	Estimate capital and operating costs of fabric filter system.	Sect. 4.5.2 and 4.6.1.3

2.2 CENTRIFUGAL SEPARATORS

2.2.1 Introduction

Centrifugal separators, commonly referred to as cyclones, are widely used in industry for the removal of solid and liquid (aerosol) matter (hereafter referred to as particles or particulates) from gas streams. Typical applications are found in mining and metallurgical operations, the cement and plastics industries, pulp and paper mill operations, chemical and pharmaceutical processes, petroleum production (cat-cracking cyclones), and combustion operations (fly-ash collection).

Particles suspended in a moving gas stream possess inertia and momentum and are acted upon by gravity. Should the gas stream be forced to change direction, these properties can be utilized to promote centrifugal forces to act on the particles. In the conventional unit (see Figure 2.2), the entire mass of the gas stream with the entrained particles tangentially enters the unit and is forced into a constrained vortex in the cylindrical portion of the cyclone. Upon entering the unit, a particle develops an angular velocity. Due to its greater inertia, it tends to move outward across the gas stream lines in a tangential rather than a rotary direction; thus, it attains a net outward radial velocity. By virtue of its rotation with the carrier gas around the axis of the tube (main vortex) and its higher density with respect to the gas, the entrained particle is forced toward the wall of the unit. Eventually the particle may reach this outer wall where it is carried by gravity and assisted by the downward movement of the outer vortex and/or secondary eddies toward the dust outlet at the bottom of the unit. The flow vortex is reversed in the lower (conical) portion of the unit, leaving most of the entrained particles behind. The cleaned gas then passes up through the center of the unit (inner vortex) and out of the collector. The spiral motion of both vortices is in the same direction. The tangential velocity (how fast the gases are swirling) is lowest near the wall and at the center of the cyclone; it reaches a maximum at a point approximately 60 to 70 percent of the way from the wall to the center. In addition to the tangential velocity variation, there are also vertical eddies and what is called inward drift. The inward drift is a radial gas flow that moves toward the center of the cyclone, opposing the movement of particles. While vertical eddies can exist in the cone, the most troublesome are those in the annular region near the gas inlet.

The eddies, which are caused by the vortices, can carry particles directly from the gas inlet to the gas outlet with a consequent detrimental effect on collection efficiency. Eddy currents in the annular region require the gas outlet to extend into the cyclone in order to prevent excessive amounts of dust from passing directly from the inlet to the outlet. Usually, this extension ends just below the bottom of the inlet.

2.2.2 Available Equipment

Cyclones may generally be classified into four categories (see Figure 2-1) depending on how the gas stream enters the unit and how the collected particles leave the unit.

The four cyclone categories are:

1. tangential inlet and axial dust discharge

    a. Conventional (large diameter, greater than 10 inches)
    b. High efficiency (small diameter, less than 10 inches)

2. tangential inlet and peripheral dust discharge

3. axial inlet and axial dust discharge

4. axial inlet and peripheral dust discharge.

Types 1 and 3 are the most widely used. Large diameter cyclones, with body diameters three to five times the diameter of the inlet duct, are useful where large gas handling capacity and moderate particle collection efficiency are required. The ratio of gas volume to capital investment dollar is greater than that for most cleaning devices. As can be deduced from the definition of "high-efficiency" cyclones, decreasing the body diameter will increase the efficiency. This is due to increased separation forces caused by the smaller vortex radius. Individual high-efficiency, small diameter cyclones have a small capacity, and they must be operated in parallel to handle typical gas volumes. They generally have a common gas inlet, dust hopper, and gas outlet and can be arranged in banks of up to several hundred cyclones each. Typical performance of conventional and high-efficiency cyclones[15] for different particle sizes is shown in Table 2-1. Conventional cyclone performance often experienced in various applications[15] is shown in Table 2-2.

Several equipments using the principle of centrifugal separation are available; some of them are described here.

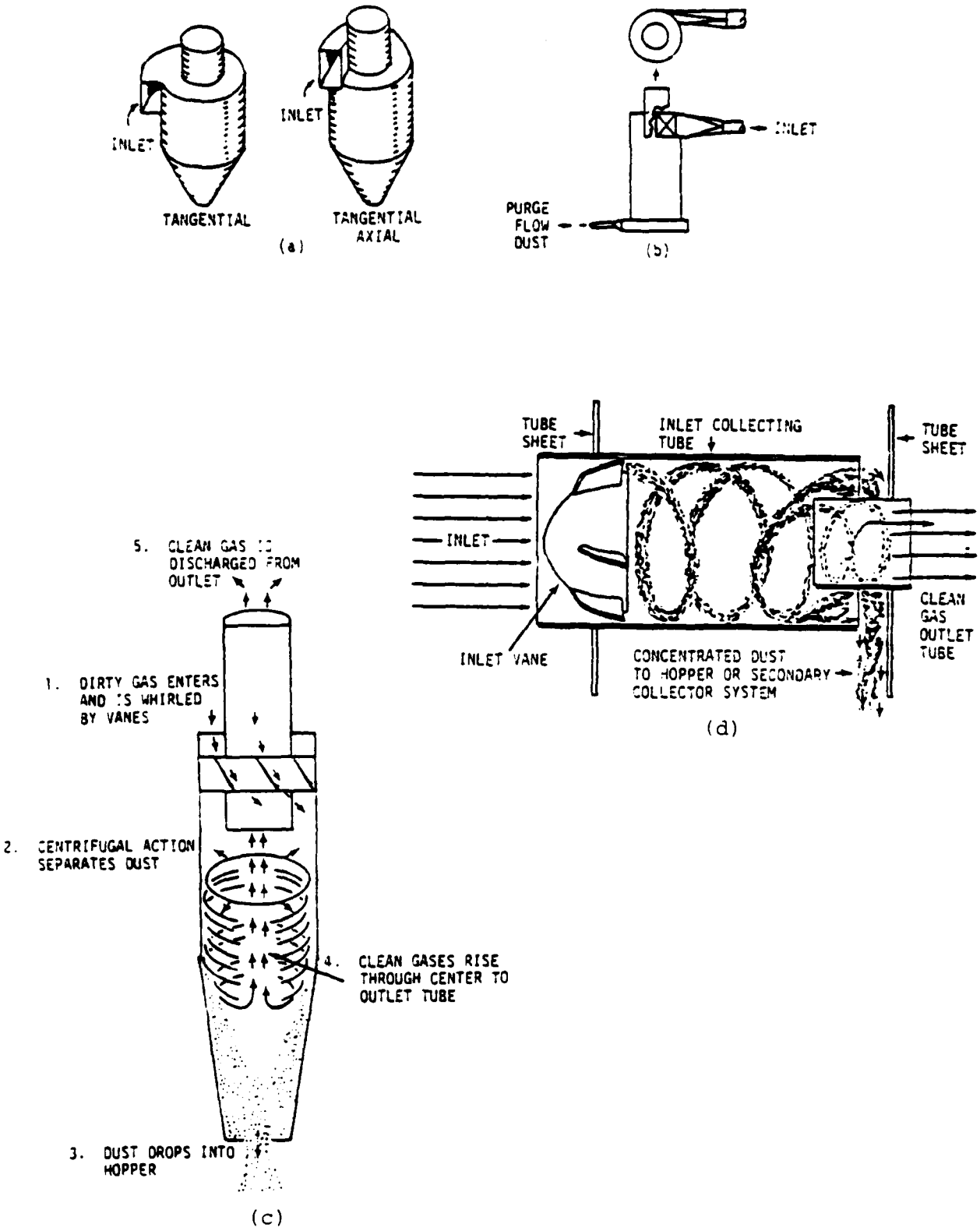

FIGURE 2-1

GENERAL CYCLONE CATEGORIES[15]

(a) Tangential inlet and axial dust discharge. (b) Tangential inlet and peripheral dust discharge. (c) Axial inlet and axial dust discharge. (d) Axial inlet and peripheral dust discharge.

TABLE 2-1

CYCLONE COLLECTION EFFICIENCY[15]

Particle Size (m)	Efficiency (% by Weight) Conventional Cyclone (%)	High Efficiency Cyclone (%)
5	-	50-80
5-20	50-80	80-90
15-50	80-95	90-99
40	95-99	95-99

TABLE 2-2

CONVENTIONAL CYCLONE PERFORMANCE FOR VARIOUS APPLICATIONS[15]

Efficiency vs. Particle Size	% of Particles Below 10 m in Size	Efficiency Range
Fly Ash (power):		
Spreader Stoker-fired Boilers	20	90-95
PC-fired Boilers	42	75-90
Cyclone-fired Boilers	65	55-65
Nonmetallic Minerals (when collector is part of process and collector catch is reusable)		
Cement (kilns and process)	40	70-85
Asphalt Plant	10	80-95
Lightweight Aggregate (kiln)	30-40	80-90
Refractory Clays (kiln)	40-50	70-80
Lime (kiln)	40-50	75-80
Fertilizer Plant (process equipment)	40	80-85
Steel (ore benefication)		
Pelletizing (vertical shaft and rotary kiln)	10-40	80-95
Foundry (general)	10-40	80-95
Chemical Process (drying, calcining)	10-40	80-95
Incinerators (municipal)	20-40	65-75
Coal Processing (thermal drying)	10	90-97
Petroleum (catalytic cracking process)	0.6	99+
General Industrial Application (in plant)	10-60	65-95

2.2.2.1 Conventional Cyclones

2.2.2.1.1 Description

The cyclone collector is an inertial separator. It consists of a cylinder, a tangential gas inlet, a cone to deliver the collected dust to a central disposal point, and an axial gas outlet. The dirty gas enters tangentially at the top and spirals downward in an outer vortex. Near the bottom of the cone, the gases reverse the direction and begin to move upward in an inner vortex. The spiraling action of the gases produces sufficient centrifugal force to drive the suspended particulate to the collector wall. These particles then move along the wall towards the dust discharge by the force of gravity and the downward movement of the outer vortex (Figure 2-2).

2.2.2.1.2 Typical Operating Characteristics

1. Conventional cyclones are of medium efficiency (60 to 80 percent) and are capable of handling high throughput at relatively low pressure losses, typically in the range of 2 to 4 inches of water.

2. Conventional cyclones usually have body diameters ranging from 4 to 12 ft.

3. High-efficiency (80 to 95 percent), single cyclone units are generally long and narrow. Body diameters seldom exceed 36 inches and are most often in the range of 12 to 24 inches.

4. High-efficiency, single cyclone units have pressure drops most frequently in the range of 3 to 6 inches of water.

2.2.2.2 Multiple Cyclones

2.2.2.2.1 Description

Multiple-cyclone collectors usually consist of a number of small-diameter cyclones operating in parallel with a common gas inlet and outlet. The flow pattern differs from a conventional cyclone in that instead of bringing the gas tangentially from the side to initiate the swirling action, the gas is brought in axially at the top of the collection tube and swirling action is imparted by a stationary vane positioned in the path of the incoming gas. Figure 2-3 shows a typical multiple cyclone.

2.2.2.2.2 Typical Operating Characteristics

1. Cyclone tube diameters typically range from 6 to 12 inches.

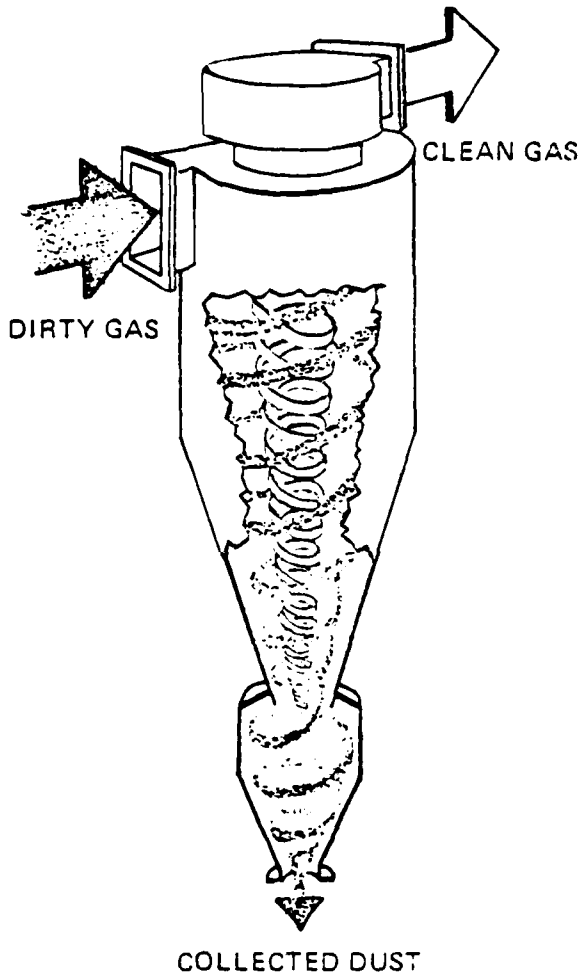

FIGURE 2-2

CONVENTIONAL CENTRIFUGAL CYCLONE[15]

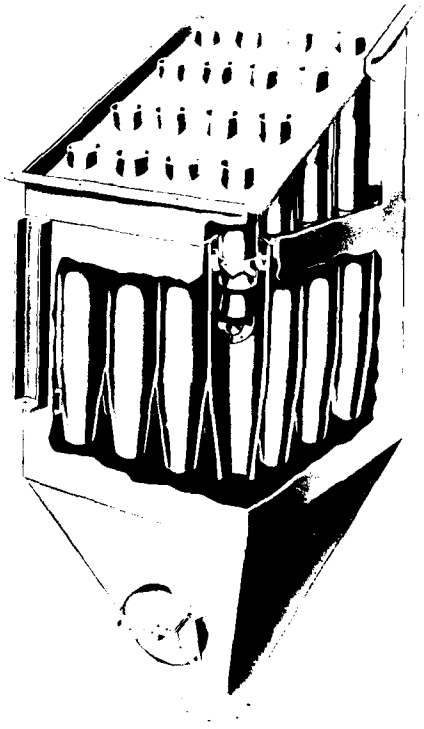

FIGURE 2-3
MULTIPLE CYCLONE[15]

2. Pressure drops are usually in the range of 2 to 6 inches of water.

2.2.2.3 Mechanical Centrifugal Collectors

2.2.2.3.1 Description

Centrifugal force may also be supplied by a rotating vane. The unit serves both as an exhaust fan and a dust collector. In operation, the rotating fan blade exerts a large centrifugal force on the particles, ejecting them from the tip of the blades to a skimmer bypass leading into a dust hopper. Efficiencies of these systems are somewhat higher than those obtained with conventional cyclones. A typcial collector is shown in Figure 2-4.

2.2.2.3.2 Typical Operating Characteristics

1. Handles gas flows up to 20,000 cfm and temperatures up to 750°F.

2. Pressure drop is about 0.5 inches of water.

2.2.2.4 Cyclone Spray Chambers

2.2.2.4.1 Description

There are many other systems capitalizing on centrifugal forces. Many of these systems utilize water to assist in the collection mechanism, and these are discussed in greater detail in Section 2.2. In cyclonic spray chambers, such as illustrated in Figure 2-5, the dust-laden gas enters tangentially at the bottom and spirals up through a spray of high-velocity fine water droplets. The dust particles are collected on the fine spray droplets, which are then hurled against the chamber wall by centrifugal action. Other units utilize water to wet and entrap the particles separated from the gas stream by centrifugal action.

2.2.2.4.2 Operating Characteristics

1. Gas flow ranges from 500 to 25,000 cfm.

2. Gas velocity into cyclone can be up to 200 fps.

3. Pressure loss ranges from 2 to 6 inches of water.

4. Water requirement is 3 to 10 gpm/1,000 cubic feet of gas cleaned.

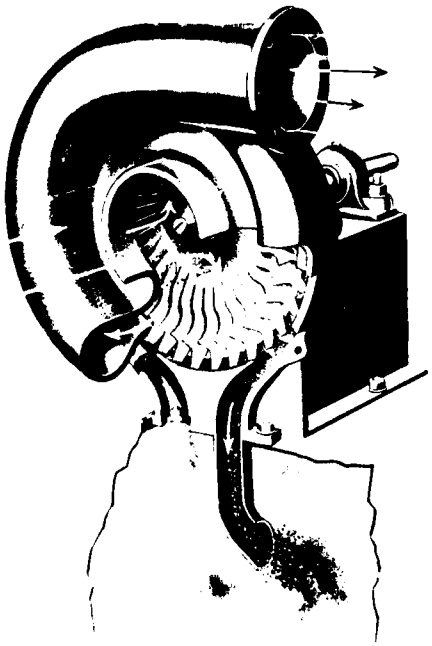

FIGURE 2-4
MECHANICAL CENTRIFUGAL COLLECTOR[15]

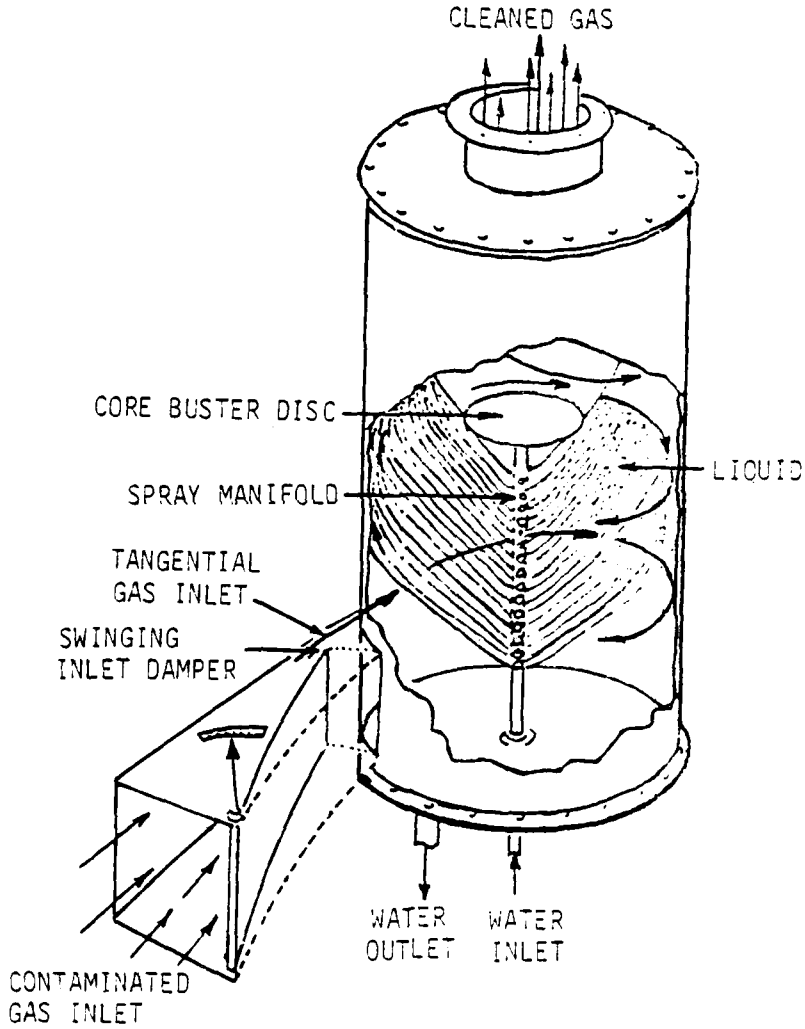

FIGURE 2-5

CYCLONIC SPRAY CHAMBER [15]

2.2.3 Design of Centrifugal Separators

The prime considerations in cyclone design are the pressure drop and overall particle collection efficiency. To establish a consistent nomenclature for the dimensions and capacity of a cyclone, Figure 2-6 gives the flow rate calculation and typical proportions used in conventional cyclone design. Note that all variables are relative to the cyclone body diameter, Dc. Cyclones are by no means limited, however, to the proportions specified in Figure 2-6. Figure 2-7 depicts the typical dimensions and flow rate calculation for a high-efficiency, medium throughput single cyclone while Figure 2-8 gives the same information for a medium-efficiency, high throughput single cyclone.

The performance of a cyclone is usually specified in terms of a cut size, D_{pc}, which is the size of the particle collected with 50 percent efficiency. The cut size depends on the gas and particle properties, the cyclone size, and the operating conditions. It may be calculated[15] from

$$D_{pc} = \left[\frac{9\mu \ B_c}{2N_t V_i (\rho_p - \rho_G)\ \pi} \right]^{0.5} \qquad (2.1)$$

where

D_{pc} = cut-size particle diameter (particle collected at 50 percent efficiency), ft

μ = gas viscosity, lb/(ft)(sec)

B_c = width of gas inlet, ft

N_t = effective number of turns the gas stream makes in the cyclone

v_i = inlet velocity, ft/sec

ρ_p = particle density, lb/ft^3

ρ_G = gas density, lb/ft^3

Lapple[23] provides a convenient graphic solution to Equation 2.1 for typical cyclones (as shown in Figure 2-6) having an inlet velocity of 50 ft/sec, a gas viscosity of 0.02 (cp), an effective number of turns equal to five, and a cyclone inlet width of $D_c/4$. This graphic solution is shown[15] in Figure 2-9. The cut size may be approximated knowing only the cyclone diameter and the true particle specific gravity. Corrections for viscosity, inlet gas velocity, effective number of turns, and inlet width, different from those assumed, may be found graphically[15] from Figures 2-10 and 2-11.

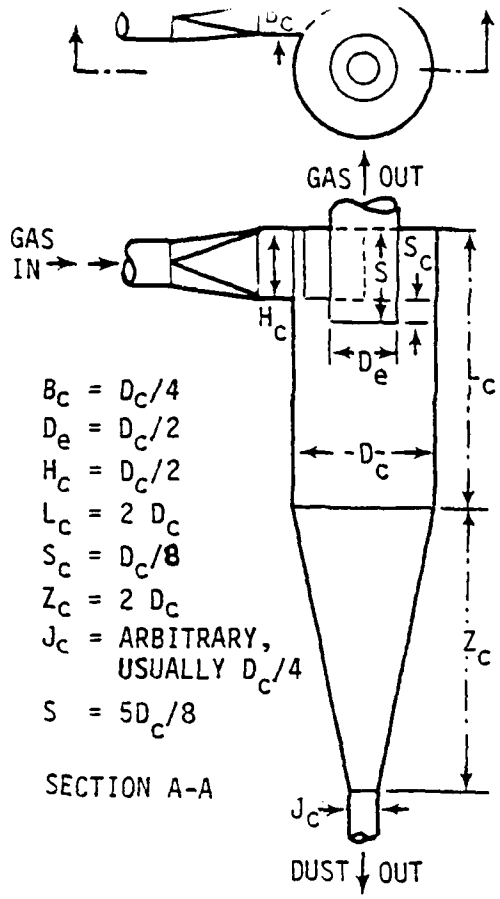

FIGURE 2-6. Typical dimensions of conventional cyclone (Normal flow rate = 500 D_c^2 ft^3/min).[15]

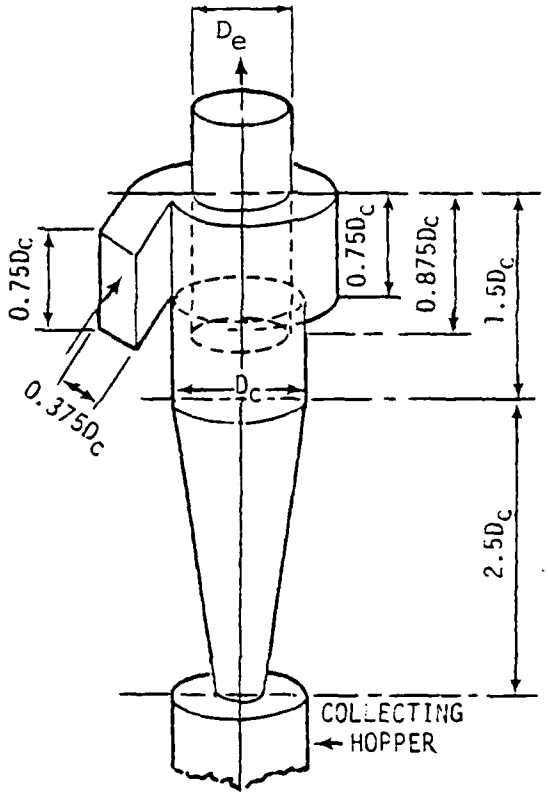

FIGURE 2.8. Typical dimensions of a medium-efficiency, high-throughput single cyclone (normal flow rate = 900 D_c^2 ft^3/min).[15]

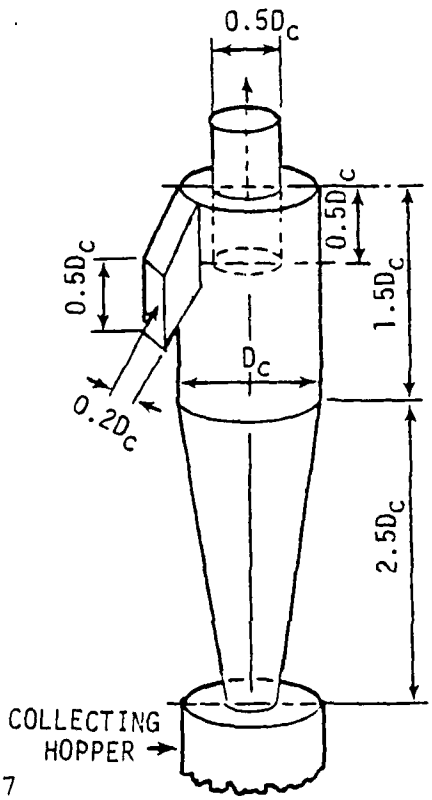

FIGURE 2-7

TYPICAL DIMENSTIONS OF A HIGH-EFFICIENCY, MEDIUM-THROUGHPUT SINGLE CYCLONE (NORMAL FLOW RATE = 300 D_c^2 ft^3/min)[15]

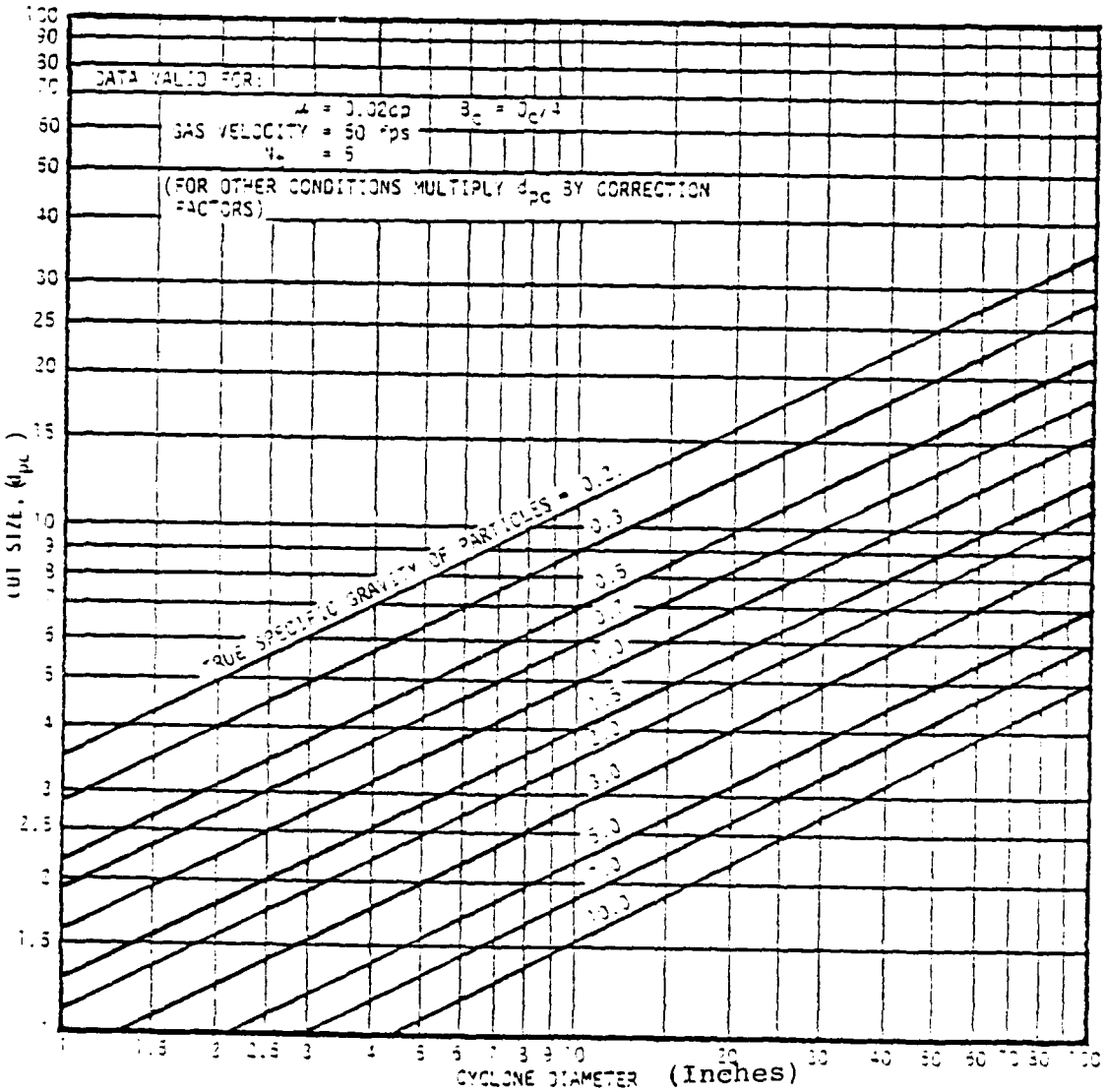

Figure 2-9

CUT SIZE IN MICRONS FOR CYCLONES OF CONVENTIONAL TYPE [15]

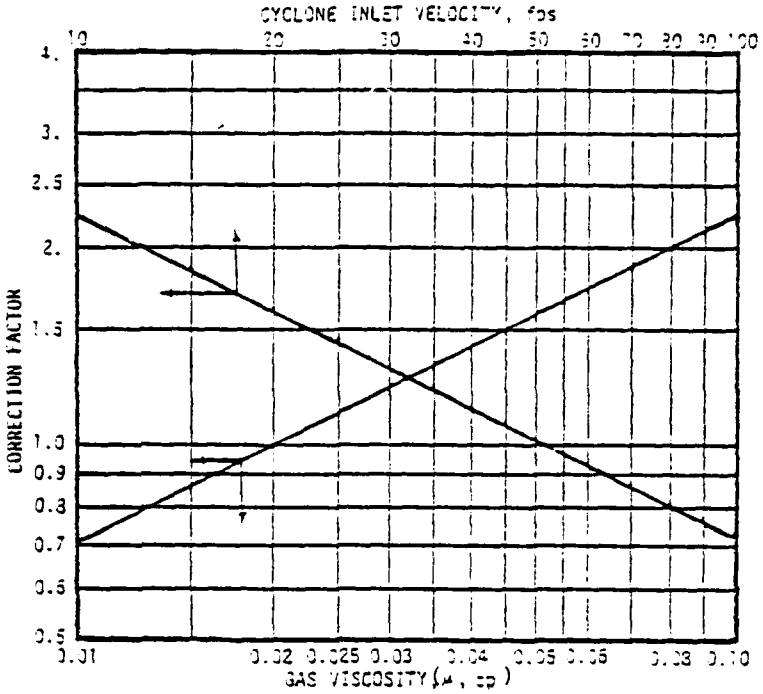

FIGURE 2-10

VISCOSITY AND VELOCITY CORRECTION FACTORS FOR CUT SIZE PARTICLE OF CONVENTIONAL CYCLONES [15]

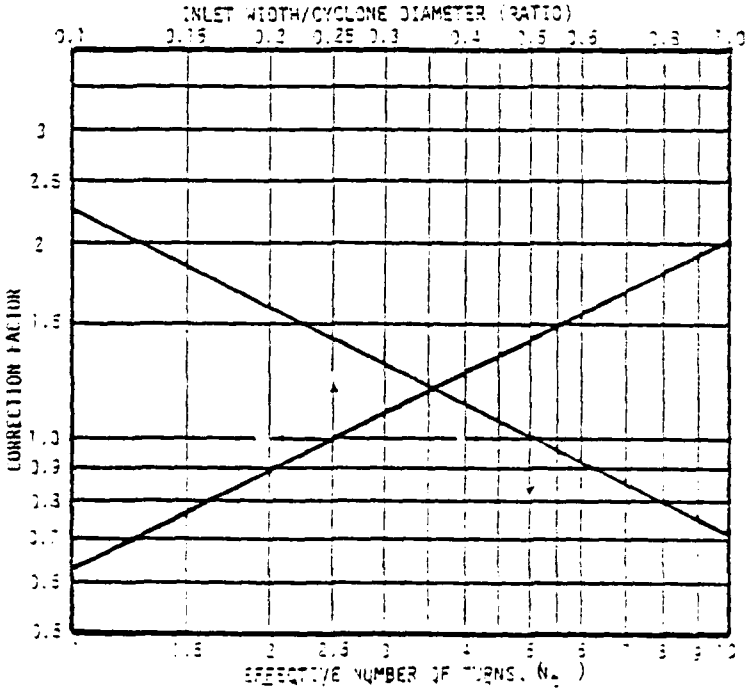

FIGURE 2-11

INLET WIDTH/CYCLONE DIAMETER AND EFFECTIVE NUMBER OF TURNS CORRECTION FACTORS FOR CUT SIZE PARTICLE OF CONVENTIONAL CYCLONES [15]

Although the effective number of turns, N_t, in a conventional cyclone with the proportions shown in Figure 2-6 has been found to be about 5, this value typically ranges from 3 to 10, but may be as low as 0.5 in some designs. From a theoretical viewpoint,[15]

$$N_t = \frac{(t_r\ v_i)}{\pi\ D_c} = \frac{(V/Q)\ v_i}{\pi\ D_c} \qquad (2.2)$$

where

t_r = residence time of the gas stream, sec
V = volume of cyclone, ft^3
Q = volumetric flow rate, ft^3/sec
v_i = inlet velocity, ft/sec.

The effective volume, V, of a cyclone with the dimensions given in Figure 2-6 is[15]

$$V = \frac{\pi}{4}\left[\left(\frac{Zc}{Dc-Jc}\right)\left(\frac{Dc^3-Jc^3}{3}\right) + Dc^2Lc - D_e^2\ (Hc + Sc)\right] \qquad (2.3)$$

For a conventional cyclone designed in accordance with the proportions given in Figure 2-6, this becomes

$$V = 2.135\ D_c^3 \qquad (2.4)$$

Substituting Equation 2.4 into Equation 2.2 allows N_t to be solved for directly (assuming that the proportions in Figure 2-6 have been used).

$$N_t = \frac{(V/Q)v_i}{\pi\ D_c} = \frac{(2.135\ D_c^{\ 3}/Q)v_i}{\pi\ D_c} \qquad (2.5)$$

Since

$$Q = v_iB_cH_c = v_i(D_c/4)\ (D_c/2) = 0.125\ D_c^2v_i \qquad (2.6)$$

then

$$N_t = \frac{(2.135\ D_c^3/0.125\ D_c^2v_i)v_i}{\pi\ D_c} = 5.44$$

In actual practice for large diameter cyclones, the number of turns correlates well with the inlet velocity and may be estimated[15] using Figure 2-12.

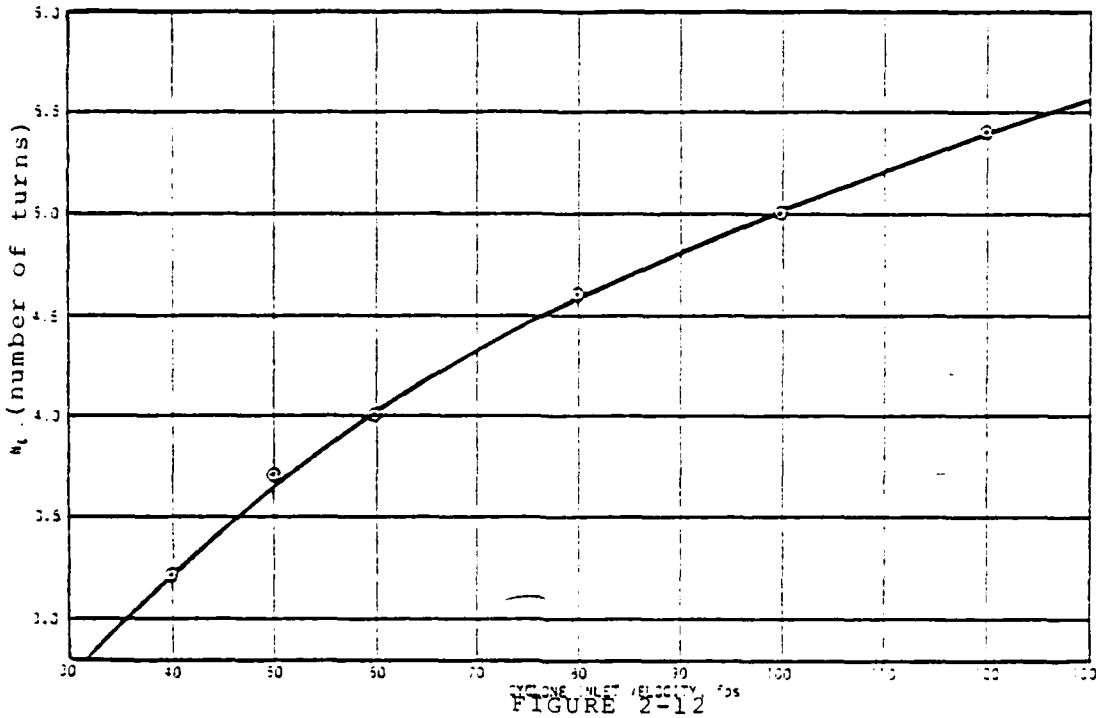

FIGURE 2-12

NUMBER OF TURNS AS A FUNCTION OF CYCLONE INLET VELOCITY [15]

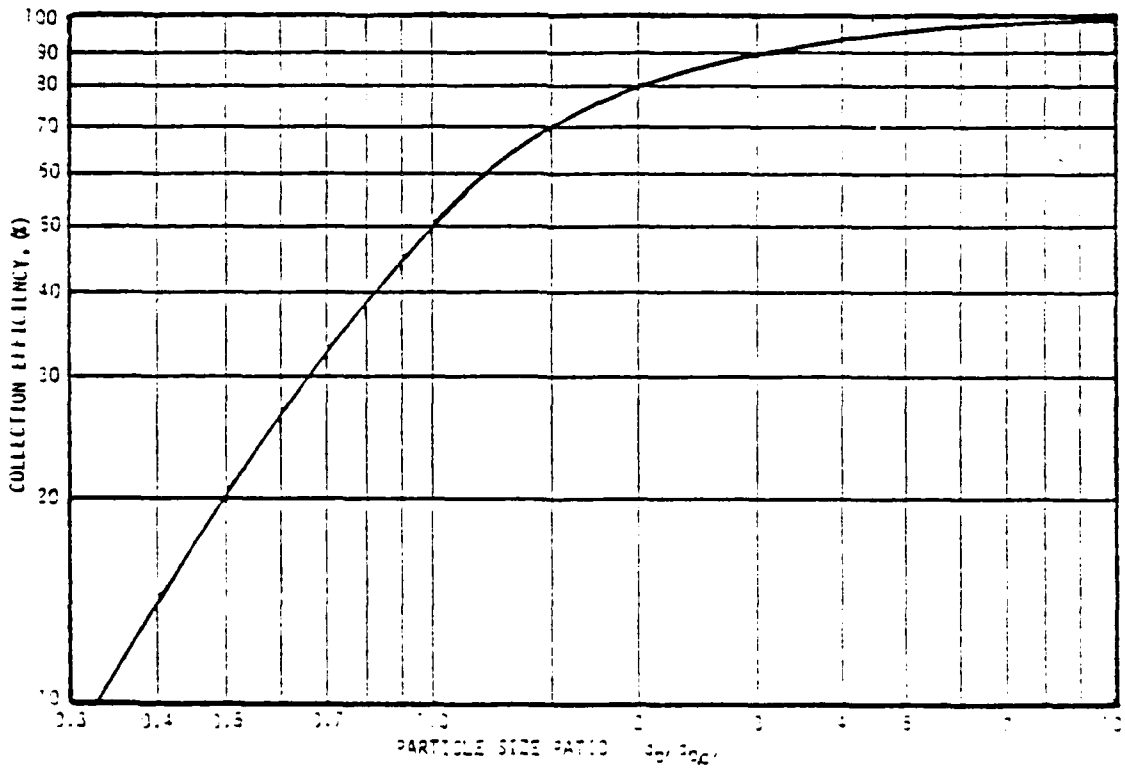

FIGURE 2-13

CYCLONE EFFICIENCY AS A FUNCTION OF PARTICLE SIZE RATIO [15]

Particles larger than the cut size d_{pc} will be collected to an extent greater than 50 percent, while smaller particles will be collected to a lesser extent. This may be represented[15] quantitatively by a curve as shown in Figure 2-13, which is essentially a generalized form of the fractional efficiency plot frequently found in commercial literature. The specific values given in Figure 2-13 apply for any cyclone of the proportions given in Figure 2-6. The calculated particle cut size used in conjunction with the general cyclone efficiency curve as shown in Figure 2-13 will determine the particle size efficiency curve for the cyclone in question. Additional experimental data were used to supplement Lapple's ratios of d_p/d_{pc}. All results compared favorably with Lapple's original curve. Typical manufacturer's efficiency curves for cyclones and multiple cyclones converted to d_p/d_{pc} curves had slightly lower efficiencies than Lapple's correlation for d_p/d_{pc} of 1.5 and 12 percent for the multiple cyclone curve at d_p/d_{pc} of 2 to 3. Apparently, Lapple's correlation is accurate enough for an engineering estimation of many cyclone applications. To determine the overall collection efficiency, the particle size distribution of the feed must be known.

A fractional efficiency curve for a geometrically similar cyclone may be constructed from a given fractional efficiency curve using the following procedure:

1. Determine d_{pc} from the fractional efficiency curve for a known cyclone (particle diameter collected at 50 percent efficiency).

2. Replot the fractional efficiency curve as efficiency vs. the ratio d_p/d_{pc}.

3. Calculate d_{pc} for the unknown cyclone from Equation 2.1 or Figures 2-9 through 2-11.

4. Assume that the efficiency vs. d_p/d_{pc} curve applies to the unknown cyclone. Using the value of d_{pc} for the unknown cyclone and the efficiency vs. d_p/d_{pc} curve, new values of d_p vs. efficiency may be calculated and plotted as the fractional efficiency curve of the unknown cyclone. In most cases, however, a range of d_{pc} for the unknown cyclone is selected instead of a single value. Then, using the maximum and minimum values for d_{pc}, two size efficiency curves can be plotted. The overall efficiencies obtained from these curves may serve as an engineering estimate of the expected cyclone performance.

The overall collection efficiency for the cyclone (and also other particulate control systems) is determined as follows:

1. Divide the particle size distribution of the dust to be collected into ranges (if this has not already been done). For example, for the distribution[15] given in Figure 2-14,

   Particle Size (μm)

d_p Range	Avg d_p		Wt %
6	3	(100.00-99.97) =	0.03
6-8	7	(99.97-99.91) =	0.06
8-10	9	(99.91-99.80) =	0.11
10-12	11	(99.80-99.65) =	0.15
12-15	13.5	etc.	0.35
15-20	17.5		0.90
20-30	25		2.80
30-40	35		3.60
40+	40+		91.90

2. From the fractional efficiency curve, either found experimentally or predicted using Lapple's technique, determine the collection efficiency for the average d_p in each size range of the inlet particle size distribution.

3. Multiply the weight fraction for each size range by the collection efficiency determined in Step 2. The summation ($\Sigma y_i \cdot \% wt_i$) gives the overall collection efficiency.

The pressure drop across a cyclone collector will generally range between 2 and 6 inches of water, and it is usually determined empirically. The method frequently used in industrial practice is to determine the pressure drop of a geometrically similar prototype. Lapple[23] suggested the relationship in Equation 2.7

$$F_{cv} = K\ (B_c H_c / D_e^2) \qquad (2.7)$$

where K varies from 7.5 to 18.4, with a value of 13 found to check with experimental data within 30 percent. The friction loss, F_{cv}, is given in units of inlet velocity heads. This inlet velocity head, expressed in inches of water, may be expressed as follows:

$$\text{one inlet velocity head} = 0.003\ \rho_G v_i^2, \text{ in } H_2O \qquad (2.8)$$

where

ρ_G = the gas density (lb/ft^3)
v_i = the inlet velocity (ft/sec).

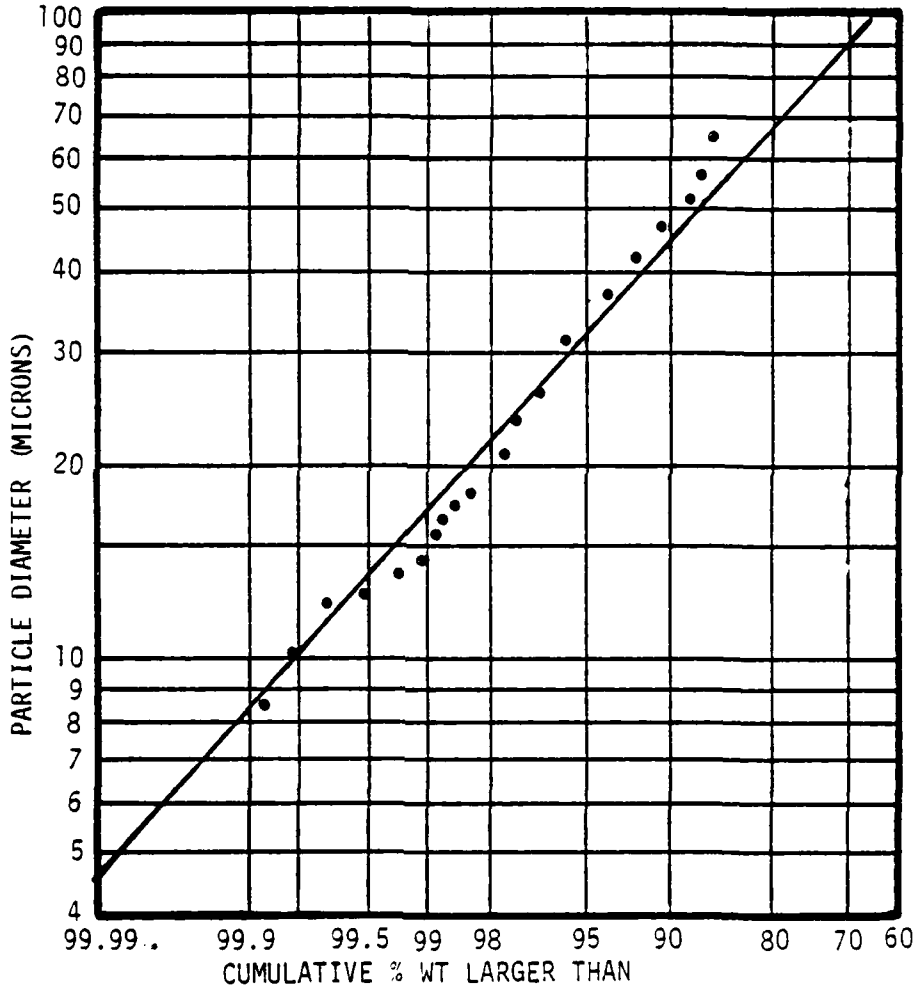

FIGURE 2-14

TYPICAL PARTICLE SIZE DISTRIBUTION[15]

The friction loss through cyclones encountered in practice may range from 1 to 20 inlet velocity heads, depending on the geometric proportions. For a cyclone of the proportions shown in Figure 2-6, the friction loss is approximately eight inlet velocity heads or, assuming an inlet velocity of 50 ft/sec (typical) and a gas density of 0.075 lb/ft^3,

$$\Delta P = 8(0.003)\ (0.075)\ (50)^2 = 4.5 \text{ in. } H_2O$$

An equation that can be used to relate the pressure drop of a cyclone operating at several different conditions or for geometrically similar cyclones is provided below.[15]

$$\Delta P = \frac{0.0027\ Q^2}{kD_e^2\ B_c H_c (L_c/D_c)^{1/3} (Z_c/D_c)^{1/3}} \qquad (2.9)$$

where

ΔP = pressure drop, in. H_2O
Q = volumetric flow rate at the inlet, ft^3/sec
D_e = diameter of gas outlet, ft
B_c = inlet width, ft
H_c = inlet height, ft
L_c = height of cylinder, ft
D_c = cyclone diameter, ft
Z_c = height of cone, ft
k = dimensionless factor descriptive of cyclone inlet vanes: 0.5 without vanes, 1 for vanes that do not expand the entering gas or touch the gas outlet wall (a in Figure 2-15), and 2.0 for vanes that expand and touch the gas outlet wall (b in Figure 2-15).

Remember that the cyclone dimensions (B_c, H_c, etc.) are the inner dimensions. For example, B_c is the inside width of the duct, not including any insulation, etc.

For three different types of cyclones, as shown in Figures 2-6, 2-7, and 2-8, Equation 2.9 reduces to the following:

(i) For the conventional cyclone shown in Figure 2-6,

$$\Delta P = \frac{0.0544Q^2}{kD_c^4} \qquad (2.10)$$

(ii) For the high-efficiency, medium throughput cyclone shown in Figure 2-7,

$$\Delta P = \frac{0.0695Q^2}{kD_c^4} \qquad (2.11)$$

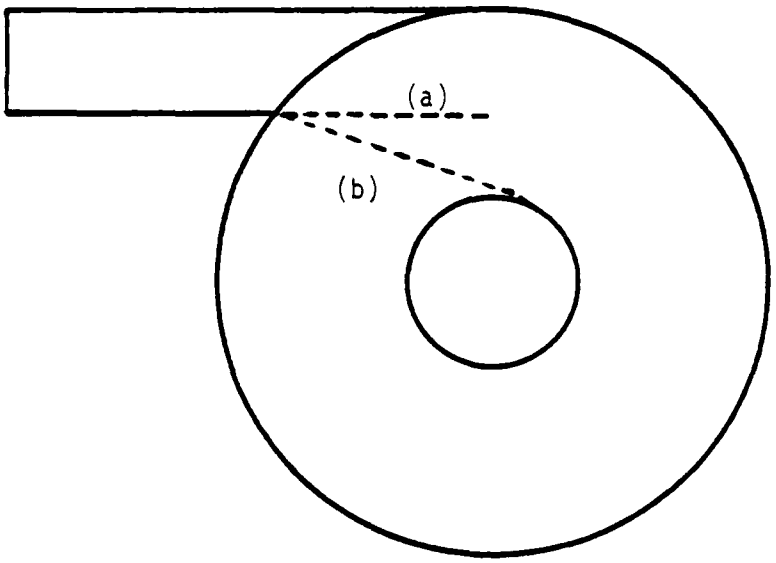

FIGURE 2-15
INLET VANES[15]

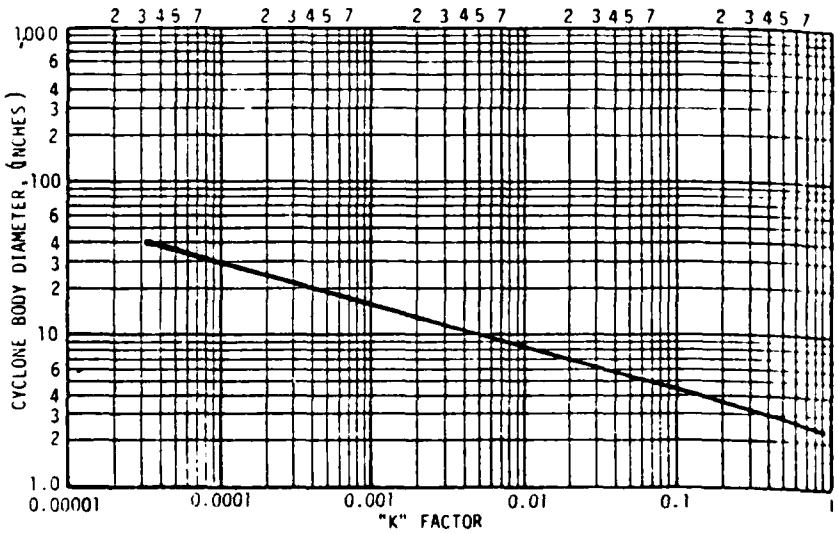

FIGURE 2-16
RELATION BETWEEN CYCLONE DIAMETER AND PRESSURE LOSS[15]

(iii) For the medium efficiency, high-throughput cyclone shown in Figure 2-8,

$$\Delta P = \frac{0.0110Q^2}{kD_c^4} \qquad (2.12)$$

An alternate pressure drop relationship, which attempts more directly to include some of the operating conditions, is given below.[15]

$$\Delta P = KQ^2 P \rho_G / T \qquad (2.12a)$$

where

P = absolute pressure, atm;
T = absolute temperature of the gas °R:
K = proportionality constant (see Figure 2-16).[15]

The design factor having the greatest effect on collection efficiency is the cyclone diameter. For a given pressure drop, the smaller the diameter of the unit, the higher the collection efficiency obtained, since centrifugal acceleration increases with decreasing radius of rotation. Centrifugal forces employed in modern designs vary from as low as 5 to as high as 2,500 times gravity, depending on the diameter of the cyclone.

For practical applications, cyclone design is usually limited by pressure drops reasonable with commercially available fans. The pressure drop will increase as a function of the inlet velocity squared. This limitation usually dictates that inlet velocities range from 20 to 70 ft/sec; however, equipment is normally designed for an inlet velocity of 50 to 60 ft/sec. Increasing the inlet velocity will increase the efficiency, although the relationship is very complex. There is also an upper limit, about 100 to 120 ft/sec, above which there is increased turbulence that, in turn, causes reentrainment of the separated particles and reduced efficiencies.

The length of the cyclone body determines the residence time during which particles are subject to the separating forces; increasing this length will increase efficiency. Also, dust entrained in the vortex core will have more time to become reseparated. Increasing the body diameter to outlet diameter ratio will also increase efficiency, although the optimum ratio is between 2 and 3.

The eddy current in the annular region requires that the gas outlet extend into the cyclone to prevent excessive amounts of dust from passing directly from the inlet to the outlet. Usually this extension ends just below the bottom of the inlet. Devices that permit the gases to leave the gas outlet tube tangentially have been successful in reducing the pressure loss without sacrificing the efficiency.

Since the pressure drop in a cyclone is caused by the vortex and not by wall friction, rough walls actually reduce the pressure drop due to suppression of vortex formation. However, they also greatly reduce the collection efficiency due to increased turbulence and reentrainment. It should also be mentioned that particle size variations have a negligible effect on the pressure drop.

2.2.4 Advantages and Disadvantages of Cyclones

Because of numerous misapplications, lack of understanding, and stricter environmental demands on particulate collection, the cyclone has lost much of its attraction and is often looked upon as something from the past. Often, a cyclone may have been installed when the only objective was to collect a portion of escaping dust that could be profitably returned to the process, while leaving the finer dust to escape. With time and the advent of environmental restrictions, even some of the finer dust must now be collected, but since the cyclone as originally designed is not capable of meeting the new demands, it is replaced with a more rigorous (and more costly) collector such as a fabric filter or wet scrubber.

Many such applications could be satisfied by either replacing the original cyclone with a properly designed, more efficient cyclone or placing a better cyclone in series with the existing unit. In many instances, the new performance requirements can be met while retaining a relatively simple piece of equipment requiring low maintenance.

Cyclones offer the following advantages:

1. low cost of construction

2. relatively simple equipment with few maintenance problems

3. relatively low operating pressure drops (for degree of particulate removal obtained) in the range of approximately 2 to 6 inches of water

4. temperature and pressure limitations imposed only by the materials of construction used

5. dry collection and disposal

6. relatively small space requirements.

The disadvantages include:

1. relatively low overall particulate collection efficiencies, especially on particulates less than 10 microns in size

2. inability to handle tacky materials.

Cyclones are favored for applications where the collected dry dust is valuable. Instances occur in the food industry where the desired product is a powder, and contamination with the minute fibers of cloth filter cannot be tolerated, or where retention of the collected powder on walls or in the cloth filter-material would generate a health hazard.

Cyclones are constantly used as the lowest-cost collector in all those applications where only a small portion of the dust to be collected is below 5 microns. In groups, or as multi-cyclones, these devices are often used as a first-stage collector in large modern plants. In case the dust concentration proves too large for cloth filters, electrostatic precipitators, or even wet collectors, cyclones are provided for precollection. It may also be necessary to protect expensive wet collectors from excessive abrasion, and in such instances cyclones offer low-cost protection.

A special area of application for cyclones is the cleaning of very hot gases that have high dust loads. The units are built from heat-resistant materials and are enclosed in a refractory-lined vessel. For cyclones of larger diameter, it is possible to line internal surfaces with layers of insulating, abrasion-resistant refractory.

One such application is the use of cyclones in separating valuable catalyst from gas streams during fluid catalytic cracking. In order to reach the rather high collection efficiencies commonly required for such plants, where initial dust concentrations can reach 7,000 grains per cubic foot of gas or more, cyclones are strung together in two or three stages.

Other similar applications include recovery of usable products in operations such as drying and calcining collection of iron-oxide dust in ore benefication processes and various gas-cleaning operations connected with fluid-bed incinerators.

For high-pressure gas applications, conventionally designed cyclones can be enclosed by a pressure vessel, with the cyclone being exposed only to the operating differential pressure drop plus some safety margin. Cyclones have been designed for removing sand and grit from very high-pressure natural-gas wellheads, and operate quite effectively enclosed in a pressure vessel. Collected particulate is removed through letdown valves operating on a continuous basis. Other high pressure applications where cyclones seem to be the only suitable collectors occur in various coal-gasification systems.

Cyclone dust collectors can have a medium to high collection efficiency and handle gases containing high dust concentrations. As such, they have become the workhorses of a number of industries, including the chemical industry.

While the cyclone dust collector will rarely meet air pollution control codes when used alone, it should always be evaluated as a precleaner to be followed by other collectors.

Using a cyclone precleaner will not necessarily reduce the size of the secondary dust collector, but it will very often both simplify the design and increase the reliability of the secondary collector and accessory equipment.

2.2.5 Illustrative Examples

2.2.5.1 Fractional Efficiency Determination

A particle size analysis was run on a cyclone with the following results:

Size Range (microns)	Hopper (%)	Outlet Gas (%)
0- 5	12.2	53.3
5-10	36.6	41.9
10-20	13.2	3.8
20-30	15.1	0.7
30+	22.9	0.3

The overall efficiency was found to be 77.5 percent.

a. Determine the fractional efficiencies
b. Construct a fractional efficiency curve for the cyclone (see Figure 2-16A).

Solution:

Cyclone loss = 100 - 77.5 = 22.5%

$$\text{Fractional efficiency} = \frac{a}{a+b}$$

where a = (overall collection)(hopper size range fraction)
b = (cyclone loss)(outlet size range fraction).

For 0 to 5 micron size range:

a = (0.775) (0.122) = 0.0895
b = (0.225) (0.533) = 0.1199

$$\text{Fractional efficiency} = \frac{0.0895}{0.0895 + 0.1199} = 0.414$$

Size Range (microns)	Fractional Efficiency
0- 5	0.414
5-10	0.740
10-20	0.924
20-30	0.984
30+	0.996

2.2.5.2 Overall Collection Efficiency Using Lapple's Method

The particle size distribution of a dust from a cement kiln is provided below:

Particle Size (microns)	% Weight
1	3
5	20
10	15
20	20
30	16
40	10
50	6
60	3
60	7

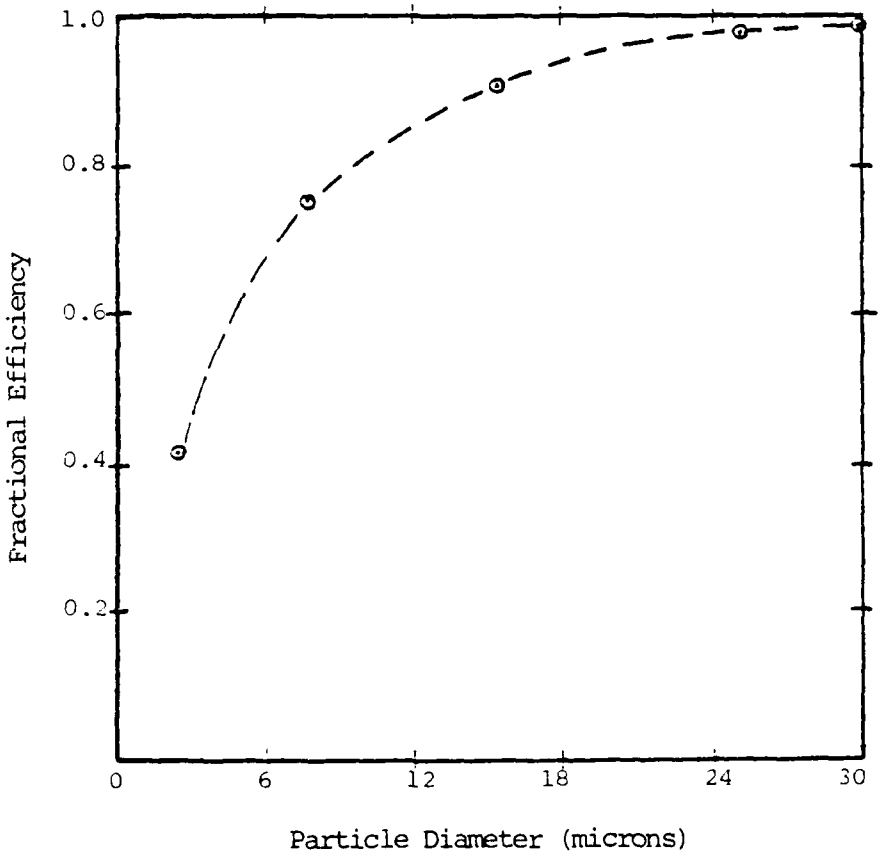

FIGURE 2-16A

FRACTIONAL EFFICIENCY CURVE

The following information is also known:

Gas viscosity	0.02 centipoise (cp)
Particle specific gravity	2.9
Inlet gas velocity to cyclone	50 ft/sec
Effective number of turns Within cyclone	5
Cyclone diameter	10 ft
Cyclone inlet width	2.5 ft

a. Determine the cut-size particle diameter, i.e., diameter of particle collected at 50 percent efficiency, and estimate the overall collection efficiency using Lapple's method.

b. If the same cyclone is used, but the inlet gas velocity is increased to 60 ft/sec and the gas viscosity changes to 0.018 cp (all else remaining the same), find the new cut size particle diameter and determine the new overall collection efficiency using Lapple's method.

Solution:

(a) Step 1.

Cyclone diameter, D_c = 10 ft (given)

Step 2.

Cyclone inlet width, B_c = 2.5 ft (given)

The cut size particle diameter, given by Eq. 2.1,

$$D_{pc} = \left[\frac{9 \mu B_c}{2 N_t V_i (\rho_p - \rho_G) g} \right]^{0.5}$$

Gas viscosity, = 0.02 cp (given)
= 0.02 x 6.72 x 10^{-4} lb/ft-sec
= 1.344 x 10^{-5} lb/ft-sec

Effective number of turns within cyclone, N_t = 5 (given)

Inlet gas velocity, V_i = 50 ft/sec (given)

Particle specific gravity, SG_p = 2.9 (given)
therefore, density, ρ_p = 2.9 x 62.4 lb/ft^3
= 180.96 lb/ft^3

Neglecting the density of air compared to the particle density,

$$D_{pc} = \left[\frac{9\ (1.344 \times 10^{-5}\ \text{lb/ft-sec})\ 2.5\ \text{ft})}{2(5)\ (50\ \text{ft/sec})\ (180.96\ \text{lb/ft}^3)\ (\pi)}\right]^{0.5}$$

$$= 3.26 \times 10^{-5}\ \text{ft}$$

$$= 3.26 \times 10^{-5}\ \text{ft} \times 30.48 \times 10^{4}\ \frac{\text{microns}}{\text{ft}}$$

$$= 9.94\ \text{microns}$$

Use D_{pc} = 10.0 microns

Step 3.

From Figures 2-10 and 2-11, all the four correction factors, viz., velocity, viscosity, width/diameter,and number of turns, are unity. The value of the cut size particle diameter, therefore, remains the same as determined in Step 2.

Step 4.

Fractional collection efficiency is determined using the given particle size distribution to the cyclone and Figure 2-13.

d_p (microns)	Weight %	d_p/D_{pc}	(%)*
1	3	0.10	nil
5	20	0.50	20
10	15	1.00	50
20	20	2.0	80
30	16	3.0	90
40	10	4.0	95
50	6	5.0	98
60	3	6.0	99
60	7	-	100

*From Figure 2-13, collection efficiency vs. d_p/D_{pc}

Step 5.

Overall collection efficiency is given by the summation of the product of weight fraction and fractional collection efficiency (columns 2 and 4 of the above table) for all the size intervals.

$$\eta = (0.03)\ (0.0) + (0.20)\ (0.20) + (0.15)\ (0.50) + \ldots\ldots + (0.03)\ (0.99) + (0.07)\ (1.00)$$
$$= 0.671$$

Thus, overall collection efficiency = 67.1 percent.

(b) Steps 1 and 2 are same as in (a).

Step 3.

Since the inlet conditions have varied, it is necessary to apply correction factors to the previous particle cut size diameter, D_{pc}. Apply correction factors found in Figures 2-10 and/or 2-11.

Inlet velocity correction (from Fig. 2-10) = 0.92
Viscosity correction (from Fig. 2-10) = 0.95

Corrected cut size = D_{pc} (correction factors)
= 10.0 (0.92) (0.95)

= 8.74 microns

Step 4.

Fractional collection efficiency is determined as in (a).

d_p (microns)	Weight %	d_p/D_{pc}	η (%)
1	3	.11	nil
5	20	.57	24
10	15	1.14	55
20	20	2.29	83
30	16	3.43	92
40	10	4.58	95
50	6	5.72	96
60	3	6.86	98
60	7	-	98

Step 5.

$$\text{Overall efficiency} = \sum_{\substack{\text{All}\\ \text{Size}\\ \text{Intervals}}} (\text{weight fraction})(\text{fractional efficiency})$$

= 0.696
Overall efficiency = 69.6%

2.3 WET SCRUBBERS

2.3.1 Introduction

Wet scrubbers have found widespread use in cleaning contaminated gas streams because of their ability to remove effectively both particulate and gaseous pollutants. Specifically, wet scrubbing describes the technique of bringing a contaminated gas stream into intimate contact with a liquid.

The particulate collection mechanisms involved in the wet scrubbing operation may include some or all of the following:

- Inertial impaction
- Direct interception
- Diffusion (Brownian movement)
- Electrostatic forces
- Gravitational forces
- Condensation
- Thermal gradients

Inertial impaction occurs when an object (the droplet), placed in the path of a particulate-laden gas stream, causes the gas to diverge and flow around it. Larger particles, however, tend to continue in a straight path because of their inertia; they may impinge on the obstacle and be collected (Figure 2-17.a).[15] Direct interception also depends on inertia and is merely a secondary form of impaction. A collision occurs due to direct interception if the dust particle's center misses the target object by some dimension less than the particle's radius (Figure 2-17.b).[15] Direct interception is, therefore, not a separate principle, but only an extension of inertial impaction. Diffusion is another extension of the impaction principle. Very small particles (submicron) suspended in a gas stream have an individual oscillatory motion known as Brownian movement (Figure 2-17.c).[15] In this case, particle and target collide as a result of relative motion within limited space. As in all diffusional processes, the rate of diffusion is favored by large areas for diffusion, thereby necessitating small liquid droplets with high surface-to-volume ratios for high collection efficiencies. While collision or impaction may be the result of either inertia or Brownian movement, the results are the same.

Gravitational forces can also cause a particle, as it passes an obstacle, to fall from the streamline and settle

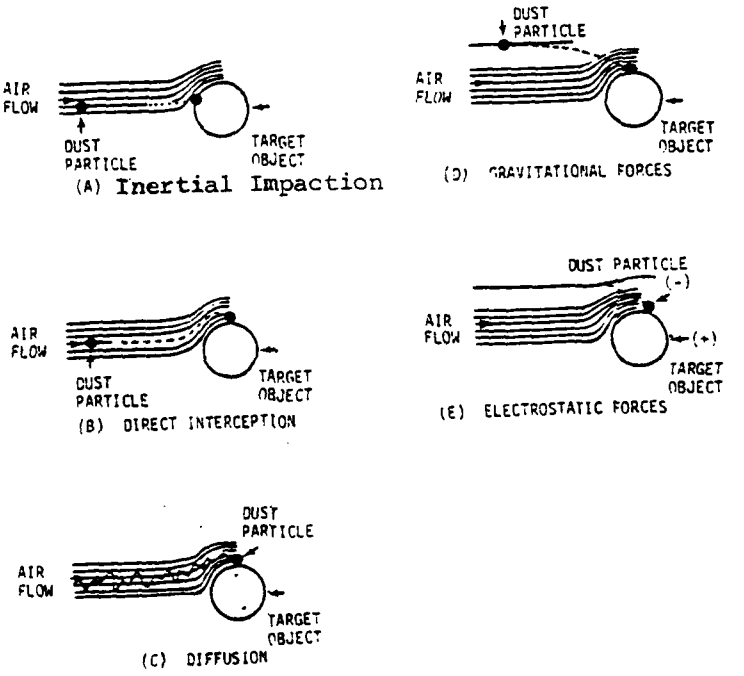

FIGURE 2-17

PRINCIPAL COLLECTION MECHANISMS IN WET SCRUBBERS[15]

on the surface of the obstacle (Figure 2-17.d).[15] Such forces come into play, however, when dealing with the larger size particles (usually greater than 40 μm). Electrostatic forces result when particles and liquid droplets become electrically charged (Figure 2-17.e).[15] In addition, when only the particle or obstacle is charged, a charge may be induced on the uncharged component, resulting in a polarization force that can also effect particle removal. The electrostatic charge may be acquired, for example, by liquid droplets during their formation. An electrical charge also may be induced by flame ionization, friction, or the presence of charged matter. The effect of the electrostatic mechanism may be significant when the charge on the particle or obstacle is high and when gas velocity is low. Condensation effects may also come into play. Condensation occurs if the gas or air is rapidly cooled below its dew point. When moisture is condensed out of the gas stream, fogging occurs, and the dust particles can serve as condensation nuclei. The dust particles can become larger as a result of the condensed liquid, and the probability of removal by impaction is increased. Particle collection may also be affected by thermal gradients. Such forces can drive particles from hotter to colder regions. The motion can be caused by unequal gas molecular collision energy on the surfaces of the hot and cold sides of the particle; it is directly proportional to the temperature gradient.

2.3.2 Available Equipment

Wet scrubbers are constructed with such a multiplicity of designs that no single type can be considered representative of the category as a whole. Some units simply consist of an existing dry-type collector modified by the introduction of a liquid phase to assist in particulate removal and to prevent particulate reentrainment; other units are specifically designed to operate as wet collectors. It is difficult to generalize about relationships among operating parameters such as pressure drop and liquid flow rate. It is also difficult to classify wet scrubbers according to particle collection mechanisms since the usual case finds many of the collection mechanisms working simultaneously in each type of scrubber. However, in a very general sense, wet scrubbers may be loosely categorized by pressure drop (or energy consumption). Low-energy scrubbers are those with typical pressure drops less than 5 inches of water; medium-energy scrubbers are those with typical pressure drops from 5 to 15 inches of water; and high-energy scrubbers are those with typical pressure drops greater than 15 inches of water.

Spray chambers and spray towers, for example, provide the lowest pressure drop and, correspondingly, the lowest collection efficiencies; they would be classified as low-energy scrubbers. The medium pressure drop group could include centrifugal fan wet scrubbers, atomizing impingement collectors, and certain packed-bed scrubbers. The most familiar of the high-pressure-drop (high energy) group is the venturi-type collector. Since many scrubbers can conceivably be listed in more than one of the above categories, in this section each type of collector will be considered individually, with its eventual application determining whether it will be classified as a low, medium, or high-energy scrubber. Only conventional scrubber systems will be considered.

2.3.2.1 Spray Chambers

2.3.2.1.1 Description

The simplest type of scrubber is a chamber in which spray nozzles are placed (see Figure 2-18). The gas stream velocity decreases as it enters the chamber, and the wetted particles settle and are collected at the bottom of the chamber. Its efficiency as a dust collector is low except for coarse dust. Efficiency can be improved by baffle plates upon which particles can be impinged.

2.3.2.1.2 Typical Operating Characteristics

1. Low pressure drop, typically 1 to 2 inches of water exclusive of mist eliminator and distribution plate.

2. Liquid requirements range from 3 to 20 gal/1000 acfm of gas.

3. Can handle large gas volumes; 800 to 2,500 lb/hr-ft^2 is typical.

4. Often used as precoolers to reduce gas stream temperatures.

5. Efficiencies often exceed 70 percent for particles larger than 10 um.

6. Improved efficiencies are possible by adding high-pressure sprays ranging from 100 to 400 psig.

2.3.2.2 Moving Bed Scrubbers

2.3.2.2.1 Description

Moving-bed (fluid-bed) scrubbers incorporate a zone of movable packing where gas and liquid can intimately mix. The system shown in Figure 2-19 uses packing

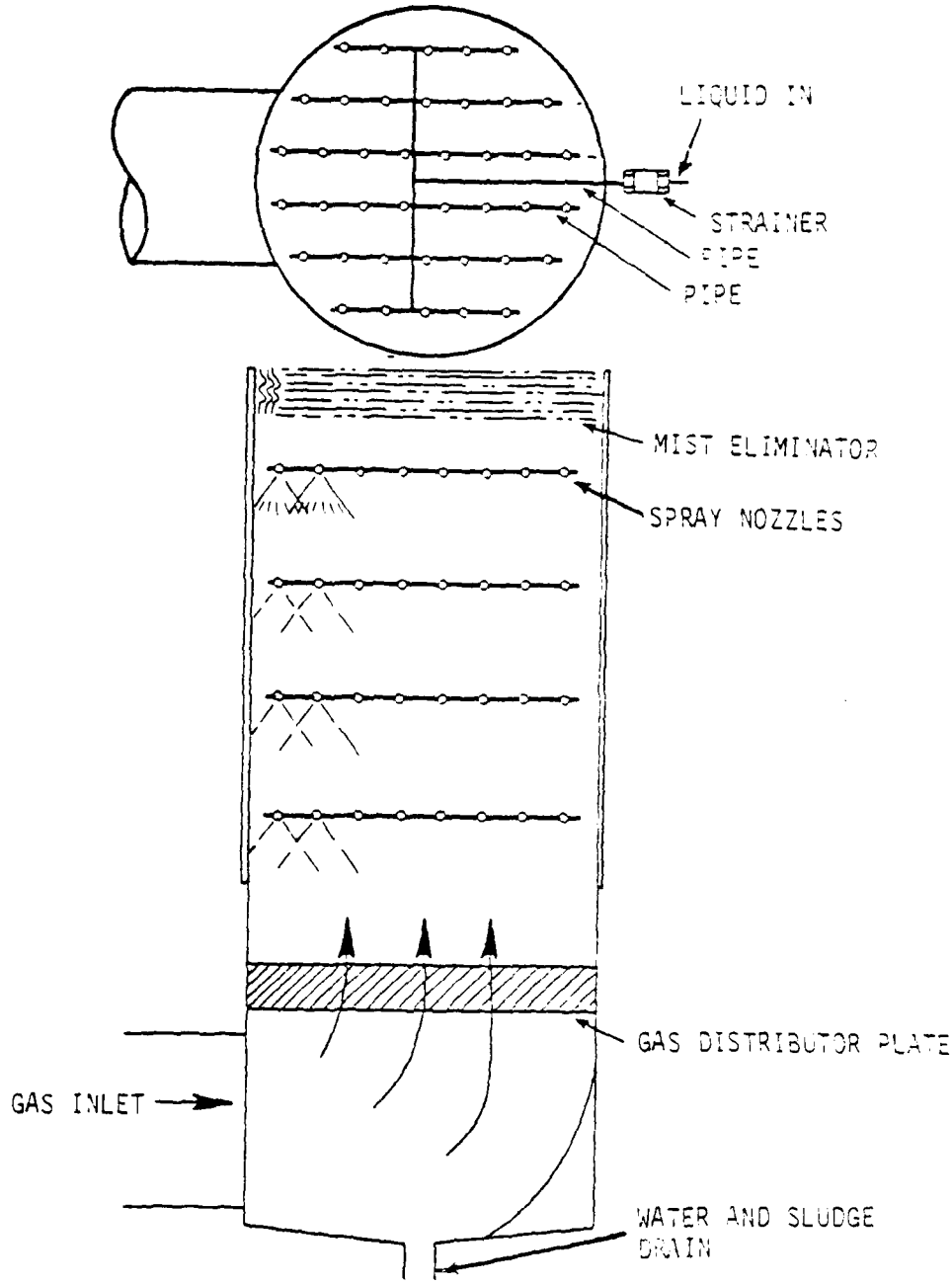

FIGURE 2-18

SPRAY CHAMBER [15]

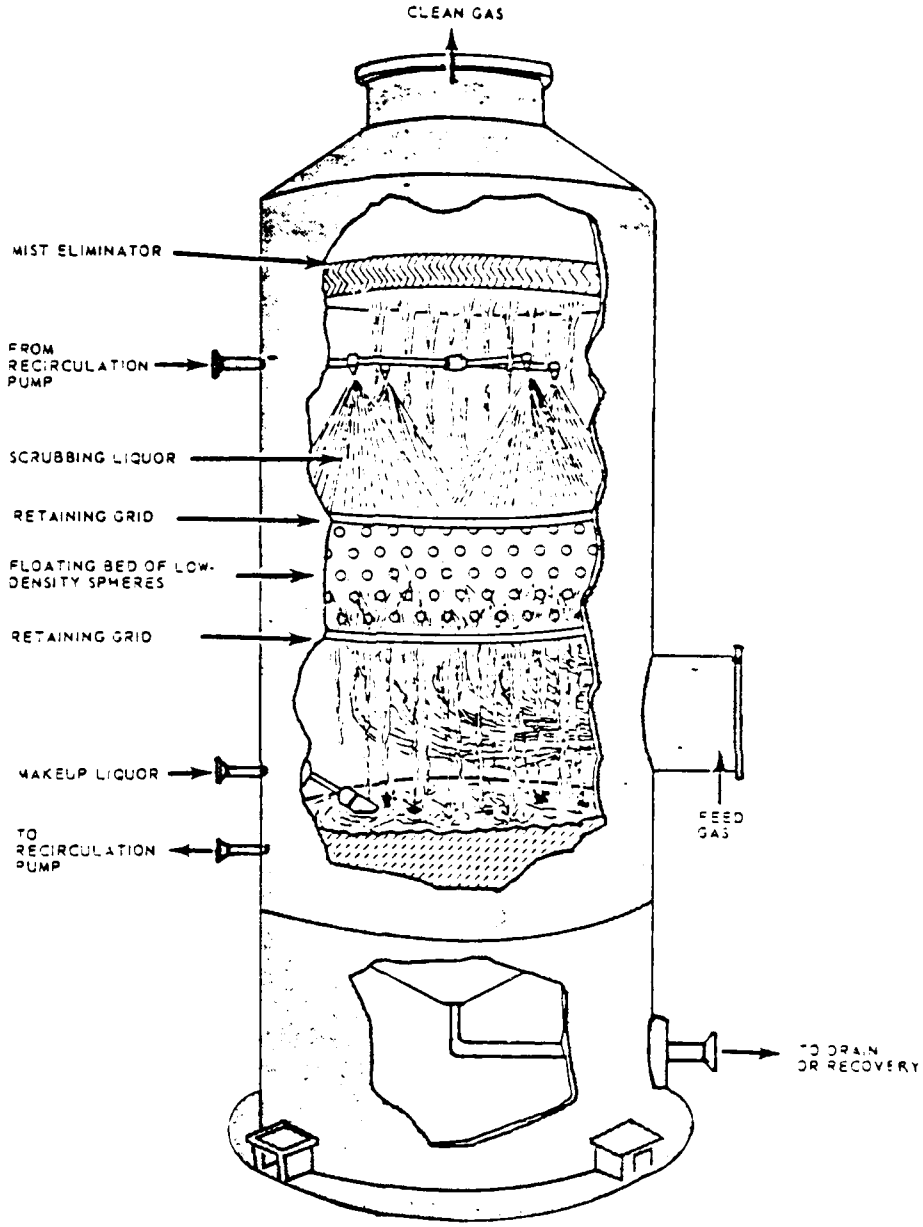

FIGURE 2-19

MOVING BED SCRUBBER[15]

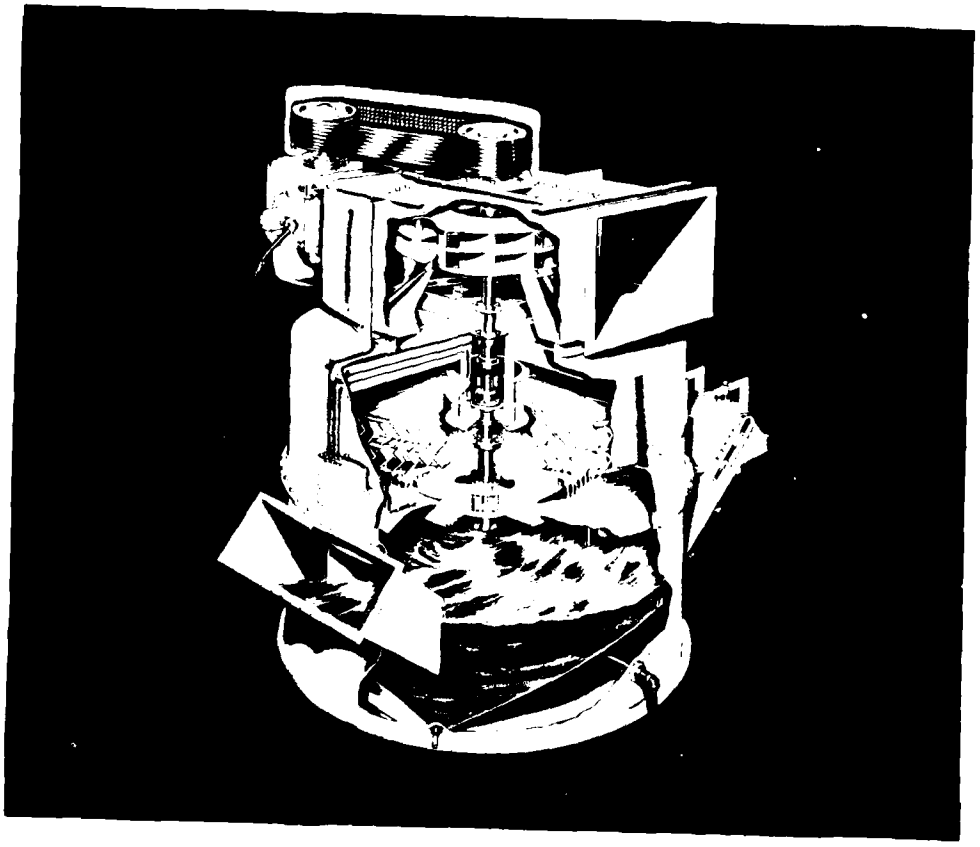

FIGURE 2-20
CENTER SPRAY HIGH VELOCITY SCRUBBER[15]

consisting of low-density polyethylene or polypropylene spheres about 1.5 inches in diameter; these are kept in continuous motion between the upper and lower retaining grids. Such action keeps the spheres continually cleaned and considerably reduces any tendency for the bed to plug.

2.3.2.2.2 Typical Operating Characteristics

1. Pressure drops typically range from 3 to 5 inches of water per stage.

2. Collection efficiencies are often in excess of 99 percent for particles down to 2 um.

3. Collection of particles may be enhanced by using several moving bed stages in series.

2.3.2.3 Atomizing Mechanical Scrubbers

2.3.2.3.1 Description

In mechanically induced scrubbers, high-velocity sprays are generated at right angles to the direction of gas flow by a partially submerged rotor (see Figure 2-20). The dirty gas stream passes through the area of the collector that contains the mechanically produced droplets. Scrubbing is achieved by impaction because of both high radial droplet velocity and vertical gas velocity. Liquid atomization occurs at the rotor and the outer wall. Recirculation rates and degree of dispersion vary widely with the different types of rotating elements.

2.3.2.3.2 Typical Operating Characteristics

1. Power requirements typcially range form 3 to 10 hp/1,000 acfm.

2. Liquid requirements range from 4 to 6 gal/1,000 acfm depending on particle size and desired collection efficiency.

2.3.2.4 Orifice-Type Wet Scrubbers

2.3.2.4.1 Description

In orifice-type wet scrubbers (sometimes referred to as self-induced spray scrubbers), the gas stream comes into contact with a pool of liquid at the entrance to a constriction (submerged orifice). Liquid is entrained and carried into restriction where greater liquid-particulate interaction occurs, resulting in a high frequency of particulate impaction on the droplets (see Figure 2-21). Upon leaving the restriction, most of the water droplets (those large enough) are separated by gravity since the gas velocity is reduced from what it was in the restriction. Smaller droplets are subsequently removed by centrifugal force and impingement on baffles located in the upper part of the unit.

FIGURE 2-21
ORIFICE SCRUBBER[15]

2.3.2.4.2 Typical Operating Characteristics

1. Pressure drop typically ranges from 3 to 10 inches of water.

2. Water requirements range from 1 to 3 gal/1,000 acfm.

2.3.2.5 Venturi Scrubbers

2.3.2.5.1 Description

The venturi scrubber is a type of gas-atomized spray scrubber. It has a converging section, a throat, and a diverging section in series (Figure 2-22). It may have a circular or a rectangular cross-section. The gas loaded with the particulates enters the converging section. Usually liquid enters the venturi upstream of the throat through nozzles. It is shattered into very small droplets by high velocity gas. These drops are then accelerated until they reach the gas velocity. Because of the difference between the velocity of gas and that of liquid drops, the gas flows past these drops. The particles suspended in the gas and moving at the gas velocity strike the liquid drops due to various mechanisms, most importantly, inertial impaction. They become attached to the drops and are removed from the gas stream. Venturi scrubbers thus provide a small droplet diameter and high relative velocity, the conditions required to achieve high collection efficiency for the particulates by impaction.

2.3.2.5.2 Typical Operating Characteristics

1. Gas velocities through venturi typically range from 12,000 to 24,000 ft/min.

2. Venturi must be followed by a separating section for the elimination of entrained droplets.

3. Water requirements most frequently range for 6 to 10 gal/1,000 acfm of gas treated.

4. High collection efficiencies are achievable with pressure drops ranging from 6 to 70+ inches of water.

5. Collection efficiency is directly related to pressure drop.

6. Variable-throat venturi scrubbers have been introduced to maintain pressure drop with varying gas flows.

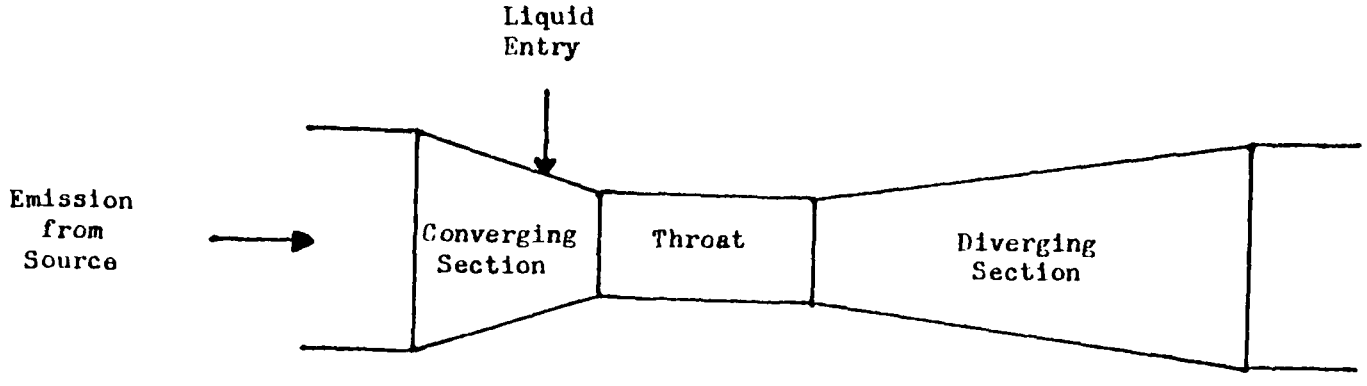

FIGURE 2-22

VENTURI SCRUBBER

2.3.3 Design of Wet Scrubbers

The design of wet scrubbers usually focuses on those parameters affecting collection efficiency and pressure drop. In most cases, the scrubber must be designed to guarantee a specified collection efficiency, which, in turn, is strongly dependent upon pressure drop (among other parameters). The system pressure drop also dictates the power requirements and the size of auxiliary equipment such as the fans.

Collection efficiency equations as a function of particle diameter typically are of the form:[15]

$$\eta = 1 - \exp\left[-k\,(L/G)\sqrt{\Psi_I}\right] \qquad (2.13)$$

where

η = collection efficiency for particle of diameter d_p
L/G = liquid-to-gas ratio, gallons/1,000 acfm
I = inertial impaction parameter
= $C\rho_p V_G\, d_p^2 / 18\, d_o\, \mu_g$
C = Cunningham correction factor (see Eq. 2.14)
ρ_p = particle density, lb/ft^3
V_g = gas velocity at venturi throat, ft/sec
d_p = particle diameter, ft
d_o = droplet diameter, ft
= $(16{,}400/V_G) + 1.45\,(L/G)^{1.5}$ for air-water system in a venturi scrubber and d_o in microns
μ_G = gas viscosity, lb/(ft) (sec)
k = correlation coefficient whose value depends upon system geometry and operating conditions, typically 0.1 to 0.2

$$C = 1 + (2\,A\,\lambda/d_p) \qquad (2.14)$$

where

λ = mean free path of gas molecules (2.14×10^{-7} ft in ambient air)
A = $1.257 + 0.40 \exp(-1.10 d_p/2\lambda)$.

Pressure drop equations typically take the form[15]

$$\Delta P = k'\, V_G^2\, (L/G) \qquad (2.15)$$

where

ΔP = pressure drop, inches of water
k' = correlation coefficient for particular scrubber design, typically 0.00005.

The practical design of a scrubber system can also be developed from the contact power theory. Contact power theory relates particulate collection efficiency in scrubbers to the pressure drop for the gas plus any power expended in atomizing the liquid. The total pressure loss, P_T, is assumed to be composed of two parts: the pressure drop of the gas passing through the scrubber, P_G, and the pressure drop of the spray liquid during atomization, P_L. These two terms can be estimated by

$$P_G = 0.157\,\Delta P \qquad (2.16)$$

where

P_G = contacting power based on gas stream energy input, hp/1,000 acfm;
ΔP = pressure drop across the scrubber, inches of water

and

$$P_L = 0.583 p_L\,(L/G) \qquad (2.17)$$

where

P_L = contacting power based on liquid stream energy input, hp/1,000 acfm
p_L = liquid inlet pressure, psi
L = liquid feed rate, gal/min
G = gas flow rate, ft^3/min.

Then,

$$P_T = P_G + P_L \qquad (2.18)$$

To correlate contacting power with scrubber collection efficiency, the efficiency is often expressed as the number of transfer units, defined by,

$$N_t = \ln[1/(1-\eta)] \qquad (2.19)$$

where

N_t = number of transfer units.

The relationship between the number of transfer units and collection efficiency is by no means unique. The number of transfer units for a given value of contacting power (hp/1,000 acfm) or vice versa varies over nearly an order of magnitude. For example, at 2.5 transfer units (η = 0.918), the contacting power can range from approximately 0.8 to 10.0 hp/1,000 acfm, depending on the scrubber and the particulate.

TABLE 2-3[24]

PARAMETERS FOR EQUATION 2.20

Aerosol	Scrubber Type	α	β
Raw gas (lime dust and soda fume)	Venturi and cyclonic spray	1.47	1.05
Prewashed gas (soda fume)	Venturi, pipe line, and cyclonic spray	0.915	1.05
Talc dust	Venturi	2.97	0.362
	Orifice and pipe line	2.70	0.362
Black liquor recovery furnace fume	Venturi and cyclonic spray	1.75	0.620
Cold scrubbing water humid gases			
Hot fume solution for scrubbing (humid gases)	Venturi, pipe line, and cyclonic spray	0.740	0.861
Hot black liquor for scrubbing (dry gases)	Venturi evaporator	0.522	0.861
Phosphoric acid mist	Venturi	1.33	0.647
Foundry cupola dust	Venturi	1.35	0.621
Open-hearth steel furnace fume	Venturi	1.26	0.569
Talc dust	Cyclone	1.16	0.655
Copper sulfate	Solivore (a) with mechanical spray generator	0.390	1.14
	(b) with hydraulic nozzles	0.562	1.06
Ferrosilicon furnace fume	Venturi and cyclonic spray	0.870	0.459
Odorous mist	Venturi	0.363	1.41

TABLE 2-4

FACTORS INVOLVED IN SCRUBBER SELECTION[15]

Environmental Factors

1. Equipment location
2. Space available
3. Ambient conditions
4. Availability of adequate water and power utilities and sludge disposal facilities
5. Maximum allowable emission (air pollution codes)
6. Visible water vapor or steam plume
7. Equipment noise levels (i.e., in case of high-energy scrubbers, the noise level of high top speed fans may be objectionable)

Engineering Factors

1. Characteristics of the dust, fume, mist, or fog to be collected
   a. Particulate-size distribution
   b. Concentration or loading
   c. Chemical reactivity
   d. Physical and chemical properties (e.g., density, solubility in scrubbing liquid, agglomeration tendencies, shape, explosiveness, stickiness).
   e. Corrosiveness and abrasiveness
   f. Toxicity
2. Characteristics of the gas stream (e.g., temperature, pressure, humidity, volume composition).
3. Characteristics of the scrubbing liquid (e.g., density, viscosity, corrosiveness, foaming tendencies).
4. Design characteristics of the scrubber
   a. Size and weight
   b. Fractional efficiency curve (i.e., collection efficiency vs. particle size)
   c. Pressure drop
   d. Reliability and dependability
   e. Method of disposal
   f. Materials of construction
   g. Effect of air volume changes (efficiency, pressure drop)
   h. Power requirements other than fan
   i. Utility requirements
   j. Temperature limitations
   k. Maintenance requirements

Economic Factors

1. First cost
   a. Equipment
   b. Installation
2. Operating cost
   a. Utilities
   b. Maintenance
   c. Savings (when recovering valuable products) or disposal costs
3. Expected equipment lifetime

For a given scrubber and particulate properties, there will usually be a very distinct linear relationship on a log-log plot between the number of transfer units and the contacting power.

$$N_t = \alpha P_T^{\beta} \qquad (2.20)$$

where

$\alpha, \beta =$ characteristic parameters for the type of particulates being collected (see Table 2-3).[24]

While the power function relationship can represent what has been observed, it should not be used to predict what will happen--except for identical conditions.

2.3.4 Scrubber Selection

Some of the more important conditions that indicate a potential scrubber application are presented below.[15]

1. Introduction of liquid to the gas is permissible to the process.

2. The liquid can be purged from the process without causing a water pollution problem. Water quality requirements of the receiving water must be considered, and a satisfactory effluent treatment system must be provided.

3. The gas must be cooled in any event.

4. Combustible particles or gases must be treated with minimum risk.

5. Vapors or gaseous matter and particulates must be removed from the gas.

There are a number of additional factors to consider in selecting a scrubber. In general, they can be grouped into three categories: economic, environmental and engineering[15]. These are outlined in Table 2-4. Proper selection of the particular type of wet scrubber for a particular application can often be difficult. It is in the best interest of the prospective user to review the literature, request performance information available from the scrubber manufacturers and, if possible, visit an installation(s) with a similar type of application. In the final analysis, one should rely on previous experience.

2.3.5 Advantages and Disadvantages of Wet Scrubbers

Wet scrubbers offer the following advantages:

1. No secondary dust sources.

2. Relatively small space requirements.

3. Ability to collect gases as well as particulates (especially "sticky" ones).

4. Ability to handle high-temperature, high-humidity gas streams.

5. Capital cost is low (if wastewater treatment system not required).

6. For some processes, the gas stream is already at high pressures (so the pressure drop may not be significant).

7. Ability to achieve high collection efficiencies on fine particulates (however, at the expense of pressure drop).

The disadvantages include:

1. May create water disposal problem.

2. Product is collected wet.

3. Corrosion problems are more severe than with dry systems.

4. Steam plume opacity and/or droplet entrainment may be objectionable.

5. Pressure drop and horsepower requirements may be high.

6. Solids build up at the wet-dry interface may be a problem.

7. Relatively high maintenance costs.

2.3.6 Illustrative Examples

2.3.6.1 Example No. 1

A vendor proposes to use a spray tower on a lime kiln operation to reduce the discharge of solids to the atmosphere. The inlet loading of the gas stream from the kiln is 5.0 grains/ft^3 and is to be reduced to 0.05 grains/ft^3 in order to meet state regulations. The vendor's design calls for a liquid pressure drop of 80 psi and a pressure drop across the tower of 5 inches of water. A gas flow rate of 10,000 acfm and water rate of 50 gal/min is proposed. Assume the contact power theory to apply.

A. Will the spray tower meet regulations?

B. What total pressure loss is required to meet regulations?

Solution:

(a) Step 1.

$$\text{Desired efficiency (\%)} = \frac{(5.0 \text{ grains/ft}^3 - 0.05 \text{ grains/ft}^3)}{5.0 \text{ grains/ft}^3} \times 100$$

$$= 99\%$$

From Eq. 2.19, the Number of transfer units,

$$N_t = \ln \frac{1}{(1-\eta)} = \ln \frac{1}{(1-0.99)} = 4.61$$

Step 2.

For a lime kiln dust and/or fume,

$\alpha = 1.47$ and $\beta = 1.05$ (Table 2-3)

Step 3.

Now, $N_t = \alpha P_T^{\beta}$ (Eq. 2.20)

$4.61 = 1.47\ P_T^{1.05}$

$P_T = 2.97$ hp/1,000 acfm

Step 4.

Liquid to gas ratio, $\frac{L}{G} = \frac{50 \text{ gal/min}}{1{,}000 \text{ ft}^3/\text{min}}$ (given)

$= 0.005 \text{ gal/ft}^3$

Water pressure into scrubber, $p_L = 80$ psi (given)

Step 5.

Contacting power based on liquid stream energy input,

$P_L = 0.583\ p_L\ (L/G)$
$= 0.583\ (80 \text{ psi})\ (0.005 \text{ gal/ft}^3)$
$P_L = 0.23$ hp/1,000 acfm

Step 6.

$P_T = P_L + P_G$ (Eq. 2.18)
$2.97 = 0.23 + P_G$
$P_G = 2.74$ hp/1,000 acfm

Step 7.

Required pressure drop, ΔP can be found as follows:

$P_G = 0.157\ \Delta P$ (Eq. 2.16)
2.74 hp/1,000 acfm = 0.157 ΔP
ΔP = 17.45 inches of water

Since the proposed pressure drop (5 inches of water) is less than the required pressure drop (17.45 inches of water), the spray tower will not meet the regulations.

(b) Total power requirement = 2.97 hp/1,000 acfm (Step 2 of a).

Total required pressure drop across the tower = 17.45 inches of water (Step 7 of a).

2.3.6.2 Example No. 2

The installation of a venturi scrubber is proposed to reduce the discharge of particulates from an open-hearth steel furnace operation. Preliminary design information suggests a water and gas pressure drop across the scrubber of 5.0 psi and 36 inches of water, respectively. A liquid to gas ratio of 6.0 gal/min/1,000 acfm is usually employed in this application. Estimate the collection efficiency of the proposed venturi scrubber. Assume contact power theory to apply.

Solution:

This example involves the evaluation of a venturi scrubber. For a given pressure drop, it is necessary to find the collection efficiency. Hence, the procedure will have to be reversed. Also, due to the low pressure drop, it will be assumed that

$$P_G >>> P_L \quad \text{and} \quad P_T = P_G$$

Step 7.

Pressure drop across the venturi, ΔP = 36 inches of water (given)

Step 6.

Contacting power based on gas stream energy input,

$$P_G = 0.157\ \Delta P \qquad \text{(Eq. 2.16)}$$
$$= 0.157\ (36 \text{ inches of water})$$
$$= 5.65 \text{ hp/1,000 acfm}$$

Step 5.

Contacting power based on liquid stream energy input, i.e., P_L, is neglected compared to P_G as stated earlier.

Step 4.

Information on liquid to gas ratio and water pressure into the scrubber is not required.

Step 3.

Total power required,

$$P_T = P_G = 5.65 \text{ hp/1,000 acfm}$$

Step 2.

For open-hearth steel furnace fume,

$$\alpha = 1.26 \text{ and } \beta = 0.57 \qquad \text{(Table 2-3)}$$

Step 1.

Eq. 2.20 will determine the number of transfer units.

$$\begin{aligned} N_t &= \alpha P_T^{\beta} \\ &= 1.26 \ (5.65 \text{ hp/1,000 acfm})^{0.57} \\ &= 3.38 \end{aligned}$$

The collection efficiency is computed as follows:

$$\begin{aligned} N_t &= \ln[1/(1-\eta)] \\ 3.38 &= \ln[1/(1-\eta)] \\ \eta &= 0.966 \end{aligned}$$

Collection efficiency = 96.6 percent.

2.4 ELECTROSTATIC PRECIPITATORS

2.4.1 Introduction

The process of electrostatic precipitation consists of corona formation around a high-tension wire, with particle charging by ionized gas molecules formed in the localized region of electrical breakdown surrounding the high-tension wire. This is followed by migration of the charged particles to the collecting electrodes. Finally, the particles captured on the collecting electrode are removed.

The corona is a gas discharge phenomenon associated with the ionization of gas molecules by electron collision in regions of high electric field strength. As the potential difference between the electrodes is raised, the gas near the more sharply curved electrode breaks down at a voltage less than the spark-breakdown value for the gap length in question. This incomplete breakdown, called corona, appears in air as a highly active region of glow (bluish white or possibly reddish in color) extending into the gas a short distance beyond the discharge electrode surface. The process of corona generation requires a non-uniform electric field, which is obtained by the use of a small-diameter wire as one electrode (discharge electrode) and a plate as the other electrode (collecting electrode). The application of a high voltage to this electrode configuration results in a high electric field near the wire. As distance from the wire surface increases, this field drops off rapidly at first and then tends to level out. The corona process is initiated by the presence of electrons in the high-field region near the wire. Electrons for corona initiation are supplied from natural radiation or other sources. Since they are in a region of high electric field, they are accelerated to high velocities; they may possess sufficient energy so that on impact with gas molecules in the region, they release orbital electrons from the gas molecules. The additional free electrons are also accelerated and join the ionization process. This avalanche process continues until the electric field decreases to the point where the electrons released do not acquire enough energy for ionization.

In the region where ionization is taking place, defined by the corona glow discharge, there are free electrons and positive ions resulting from electron impact ionization. The behavior of these charged particles depends on the polarity of the electrodes. The corona can be negative (if the discharge electrode is negative) or positive (if the discharge electrode is positive). In the case of a negative discharge wire, free electrons in the high-field zone near the wire gain enough energy from the field

to produce positive ions and other electrons by collision. These new electrons are, in turn, accelerated and produce further ionization, thus giving rise to the cumulative process termed an electron avalanche. The positive ions formed in this process are accelerated toward the wire. By bombarding the negative wire and giving up relatively high energies in the process, the positive ions cause the ejection of secondary electrons, necessary for maintaining the discharge, from the wire surface. In addition, high-frequency radiation originating in the excited gas molecules within the corona envelope may photoionize surrounding gas molecules, likewise contributing to the supply of secondary electrons. Electrons are attracted toward the anode; as they move into the weaker electric field away from the wire, they tend to form negative ions by attachment to neutral oxygen and nitrogen molecules. These ions form a dense unipolar cloud filling, by far, most of the inter-electrode volume; they constitute the only current in the entire space outside the region of corona glow. The effect of this space charge is to retard the further emission of negative charge from the corona, limiting the ionizing field near the wire and stabilizing the discharge. However, as the voltage is progressively raised, complete breakdown of the gas dielectric (i.e., sparkover) eventually occurs.

In positive corona, the electrons generated by the avalanche process flow toward the collection electrode. Since the positive ions are the charge carriers, they serve to provide an effective space charge, and the presence of an electronegative gas is not required.

Electrode geometry, gas composition, and gas conditions have important influences on corona generation. In general, the smaller diameter wire requires a higher electric field strength for corona initiation. For a given spacing, however, the onset of corona occurs at a lower voltage for a smaller diameter wire. Also, for a given voltage, higher currents are obtained with smaller diameter discharge electrodes. Temperature and pressure also influence corona generation by changing the gas density. In the avalanche process, the time available for accelerating an electron between collisions is a function of gas density. With increased molecular spacing, higher velocities can be achieved betwen collisions. Thus, ionizing energy can be achieved with low electric fields for low gas densities.

When gases laden with suspended particulate matter are passed through an electrostatic precipitator, the great bulk of the particles acquire an electric charge of the same polarity as that of the discharge electrodes.

This preferential charging occurs because the region of corona (i.e., the region of intensive ion-pair generation) is limited to the immediate vicinity of a discharge wire, thus occupying only a small fraction of the total cross section of the precipitator.

Two distinct particle-charging mechanisms are generally considered to be active in electrostatic precipitation: (1) bombardment of the particles by ions moving under the influence of the applied electric field (field-dependent charging) and (2) attachment of ionic charges to the particles by ion diffusion in accordance with the laws of kinetic theory (diffusion charging).

Particles in an electric field cause localized distortion of the field so that electric field lines intersect the particles. Ions present in the field tend to travel along the electric field lines. Thus, ions will be intercepted by the dust particles, resulting in a net charge flow to the particle. The ion will be held to the dust particle by an induced image charge force between the ion and dust particle. As additional ions collide with and are held to the particle, it becomes sufficiently charged to divert the electric field lines so that they do not intercept it. Under this condition, no ions contact the dust particle, and it receives no further charge. The electrostatic theory of the process shows that the saturation value of the charge on the particle is related to the magnitude of the electric field in the region where charging takes place, particle size, and particle dielectric constant. The saturation charge is proportional to the square of the particle diameter. Thus, larger particles are more easily collected than smaller ones. This mechanism of charging is called field-dependent charging.

For fine particles (diameter less than 0.2 μm), the field-dependent charging mechanism is less important, and collision between the particles and gas ions is governed primarily by thermal motion of the ion. As the charge on a particle increases, the probability of impact decreases, so that there is a decreasing charging rate associated with an increasing particle charge. This second charging process is called diffusion charging. Since the range of thermal velocities has no upper boundary, there is no saturation value associated with diffusion charging.

Field charging is the dominant mechanism for large particles with a diameter greater than about 0.5 μm, while diffusion charging predominates for small particles with diameters less than approximately 0.2 μm. In the intermediate range, both mechanisms contribute significant charge.

2.4.2 Description of Available Equipment

2.4.2.1 Cold-Side Precipitator

The electrostatic precipitator (ESP) may be classified as either a high-voltage, single-stage system or a low-voltage, two-stage system. The high-voltage type is by far the more popular; it has been used successfully to collect both solid and liquid particulate matter from many operations including smelters, steel furnaces, petroleum refineries, cement kilns, acid plants, municipal incinerators, and utility boilers. Low-voltage, two-stage precipitators, on the other hand, are limited almost exclusively to the collection of liquid particles discharged from sources such as meat smokehouses, asphalt paper saturators, pipe-coating machines, and high-speed grinding machines.

The vast majority of electrostatic precipitators installed are of the plate type. Particles are collected on flat, parallel collecting surfaces spaced 8 to 12 inches apart, with a series of discharge electrodes spaced along the center line of adjacent plates. A typical arrangement of a commercial plate-type electrostatic precipitator is shown in Figure 2-23. The gas to be cleaned passes horizontally between the plates (horizontal-flow type) or vertically up through the plates (vertical-flow type). Collected particles are usually removed by rapping (dry precipitator) or in a liquid film (wet precipitator). They are generally deposited in bins or hoppers at the base of the precipitator (dry-bottom type) or in a liquid (wet-bottom type).

The physical components of high-voltage electrostatic precipitators may be generally divided into three groups. The first group includes the discharge and collecting electrodes and their auxiliary parts. Modern precipitators use strong rigid collecting plates that are usually baffled in some manner. The baffling provides shielded air pockets that the dust falls through on its way to the dust hoppers after it is rapped loose from the collecting plates. The shielding also helps to keep the dust from being reentrained into the flue gas stream. Collecting plate auxiliaries include inlet and outlet gas ducts, electrode frames, rappers, supporting framework, dust hoppers, and protective outer shell. An enlargement of collecting plate surface area requires a proportionate expansion in these auxiliaries.

2.4.2.2 Hot Side Precipitator

The increasing use of low-sulfur coal and the accompanying high ash resistivity at normal precipitator

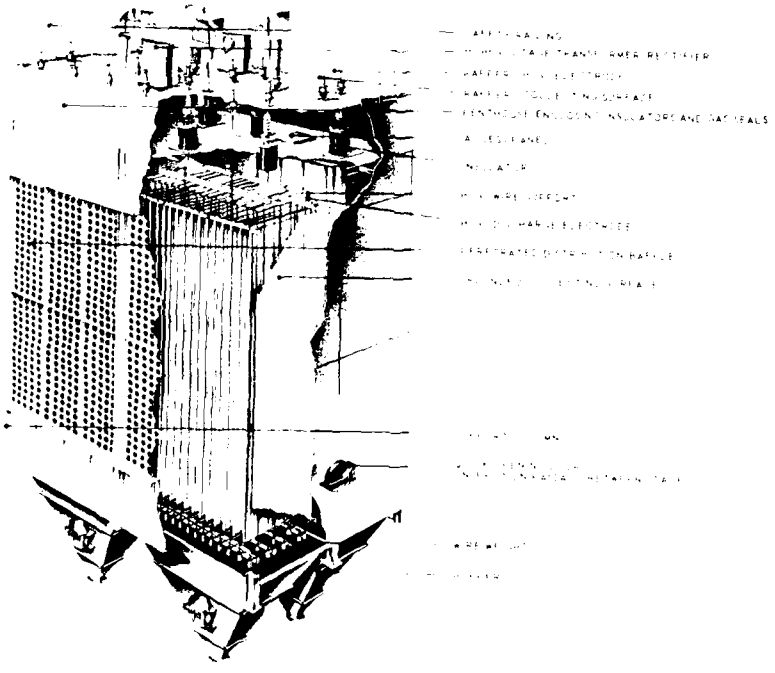

FIGURE 2-23

CUTAWAY VIEW OF A PLATE TYPE ELECTROSTATIC PRECIPITATOR[15]

operating temperatures has led to the use of so-called hot precipitators located upstream of the air heater. The flue gases ahead of the air heater are generally in the range of 600 to 900°F and the resistivity of most fly-ash is sufficiently low at these temperatures so current is not limited by fly-ash resistivity.

There are several disadvantages to the use of hot precipitators and the choice of hot-side or cold-side precipitators will generally favor cold-side operation unless the ash resistivity is exceptionally high.

The properties of the flue gas become important in hot-side precipitators because of the reduced gas density. The corona onset voltage and the sparkover voltage are both reduced and the difference between the two is also reduced. The effect is a greater problem of control of the voltage due to the steepness of the voltage-current curve. The problem may be further aggravated in the case of precipitators installed at high altitudes.

The increase in gas volume at elevated temperatures is a further disadvantage. The increased gas volume reduces the specific collection area (plate area to gas volume), which for a given size boiler results in lower efficiency of a requirement for a larger plate area. This is offset by the increase in the migration velocity. The gas viscosity is also increased at elevated temperatures and this further reduces the precipitator performance. Mechanical problems are also more severe with hot-side precipitators since expansion is greater. Provision for the increased expansion must be made to prevent buckling, warping, or failure of the shell and internal members. It is sometimes necessary to change the materials of units operating at high temperatures; the most common materials change is the use of stainless steel rather than mild steel wire.

In spite of these shortcomings the use of hot precipitators does have advantages especially when the fly-ash and flue gas composition result in a resistivity at cold-side temperatures of around 1×10^{12} ohm-cm or higher. At such resistivities, the current density would be severely reduced and the size of the precipitator required would be so large that a hot-side precipitator would be indicated. Practically, hot side precipitators are often difficult to retrofit to existing installations because of the more complex ductwork. The flue gases go from the boiler to the precipitator, back to the air heater, and then to the stack. The additional ductwork adds to the cost of a hot

system as a result of the added ducting as well as the insulation requirements.

Some difficulties have been experienced with mechanical details of hot-side precipitators due to structural failure of the shells or insulators. These difficulties can be overcome by proper attention to the requirements for thermal expansion and conventional structural design methods.

2.4.3 Design of Electrostatic Precipitators

The design of precipitators involves the determination of precipitator size and electrical energization equipment required to give a desired efficiency, the design of a gas flow system to provide acceptable gas flow quality, structural design of the precipitator shell and supports, selection of the rapping equipment, and selection of the electrode configuration.

Present design methodology is generally based upon empirical relations, the values of which have been obtained from experience with similar applications. There are several approaches to the selection of the precipitator size. In general, these methods involve the selection of a precipitation rate parameter, w, and determination of the collection plate area required from the Deutsch-Anderson equation or from design curves based upon field experience. The precipitation rate parameter varies for different applications and often varies considerably within the same application area due to variations in gas and dust properties. Selection of the precipitation rate parameter can be made on the basis of experience with similar installations or from experimentally derived curves relating precipitation rate parameters to dust properties. For many applications, the range of precipitation rate parameter variations is small (of the order of ± 10%). In such cases, the uncertainty in plate area requirements is of the same magnitude. In other instances, variations can be as high as 400 to 500 percent, so that some method for reducing the uncertainty is highly desirable. In general, some property of the effluent from the industrial process has been related to precipitation rate parameter and an empirical relationship is derived to predict the value of the precipitation rate parameter. In the case of fly-ash precipitators, sulfur content of the coal and resistivity of the dust are commonly used to establish this value. Particle size distribution is another significant variable, and curves relating precipitation rate parameter with any other variable should be modified to compensate for particle size variations if sufficient data are available.

Power requirements for a precipitator vary with collection efficiency. Selection of the power requirements is generally based on curves relating efficiency with corona power per unit volume of gas flow (watts/cfm). These curves are experimentally developed for each type of application and vary with dust properties. These curves are usually based on total delivered secondary power, and power supply capacity is selected on the basis of a power supply efficiency (from 60 to 75 percent) and the standard power supply size that will meet the efficiency requirements.

Sectionalization is also based on empirical information derived from experience. These curves must be consistent with those based on power requirements and on relationships involving collection surface area, since the same efficiencies can be achieved through the use of fewer sections and greater collection area for installations operating in a spark rate limited mode.

Design of the gas flow system is generally based on model studies with large systems, and its importance to good precipitation cannot be overemphasized.

The design of an electrostatic precipitator for a particular installation involves many parameters that can influence both cost and performance. The more significant variables involved in the design are:

1. area and type of collection electrodes
2. dimensions of the precipitator shell
3. size, spacing, and type of discharge electrodes
4. size and type of power supply units
5. degree of sectionalization
6. layout of the precipitator in accordance with physical space limitations
7. design of the gas handling system
8. size and shape of hoppers
9. type and number of electrode rappers
10. type of dust removal equipment.

There are several methods used for the selection of suitable values for these variables and each manufacturer may utilize slightly different methods in arriving at a particular design.

It should be recognized that the selection of the value of w and the curves relating power and sectionalization requirements are all interrelated. If inadequate sectionalization is used, a lower value of w would result, the precipitator could not be operated at the required power level, and the efficiency would be reduced. Consequently, curves relating to the design parameter should be internally consistent.

2.4.3.1 Deutsch-Anderson Equation and Precipitation Rate Parameter

A common approach to the selection of the area of collecting plate required is to utilize the Deutsch-Anderson equation.

$$\eta = 100 - \exp\left(-\frac{A}{Q} w\right) \qquad (2.21)$$

or

$$A = \frac{Q}{W} \ln\left(\frac{100}{100-\eta}\right)$$

where A = area of collecting surface, ft^2
Q = gas flow rate, acfm
w = precipitation rate parameter, ft/min
η = efficiency, % and
exp = base of natural logarithms.

The critical parameter in this equation is the precipitation rate w. This parameter varies with each installation depending upon resistivity and particle size of the dust, quality of gas flow, reentrainment losses, and sectionalization, among other factors. The values of w are selected by the equipment manufacturers on the basis of past experience with a particular dust, or from the composition of the dust that can be related to past experience. Each precipitator manufacturer, therefore, has a file of experience from which a precipitation rate parameter can be selected, and this file of information is kept as proprietary data.

The values of the precipitaton rate parameter, w, vary with the application as a result of variations in dust properties. Variations also occur within each application area. Table 2-5 lists the average values of precipitation rate parameters[15] for various applications and the range of values that might be expected within each application. From this table, it is apparent that the spread in the values of the precipitaton rate parameter is large in some instances, such as fly-ash precipitators, and within a reasonably narrow range within others. For the pulp and paper industry, a precipitator designed for recovery boilers would have an uncertainty of around 15 to 20 percent in precipitation rate parameter. For a precipitator designed for a 98 percent collection efficiency, the measured efficiency would range from 97 to 99.4 percent based on the range of design precipitation rate parameters.

TABLE 2-5

REPRESENTATIVE PRECIPITATION RATES FOR VARIOUS APPLICATIONS[15]

Application	Precipitation Rate, w (ft/sec) Average	Range
Utility Fly Ash	0.43	0.13-0.67
Pulp & Paper	0.25	0.21-0.31
Sulfuric Acid	0.24	0.20-0.28
Cement (wet)	0.35	0.30-0.40
Smelter	0.06	-
Open Hearth	0.16	-
Cupola	0.10	-
Blast Furnace	0.36	0.20-0.46

For fly-ash precipitators, on the other hand, the variation in precipitation rate parameters is quite large, so that a precipitator designed on the basis of a w of 0.43 for 98 percent efficiency would give an efficiency of only around 75 percent if the precipitation rate parameter were 0.13 ft/sec.

The major problem in the design of precipitators based on this approach is in the selection of the precipitation rate parameter for the specific application. Several techniques can be used to narrow the uncertainty of the value of w to be used. If the in situ resistivity of the dust is known, the precipitation rate parameter can be determined for some applications. Figure 2-24 shows the variation in w with resistivity for fly-ash precipitators[25]. If the precipitator being designed is a replacement for or an addition to an existing unit, resistivity can be measured, and the uncertainty in the value of w can be reduced. Alternatively, if a similar installation burning the same fuel is available, measurements of resistivity can be made and the value of w selected with some confidence.

The data from Figure 2-24 apply only to fly-ash or to a dust with similar properties. If the particle size differs significantly, the absolute values of w will change, although the general character of the curves would be similar.

In situ resistivity data have not been determined to the same degree in applications other than fly-ash, so that statistically reliable data relating w and resistivity are not generally available.

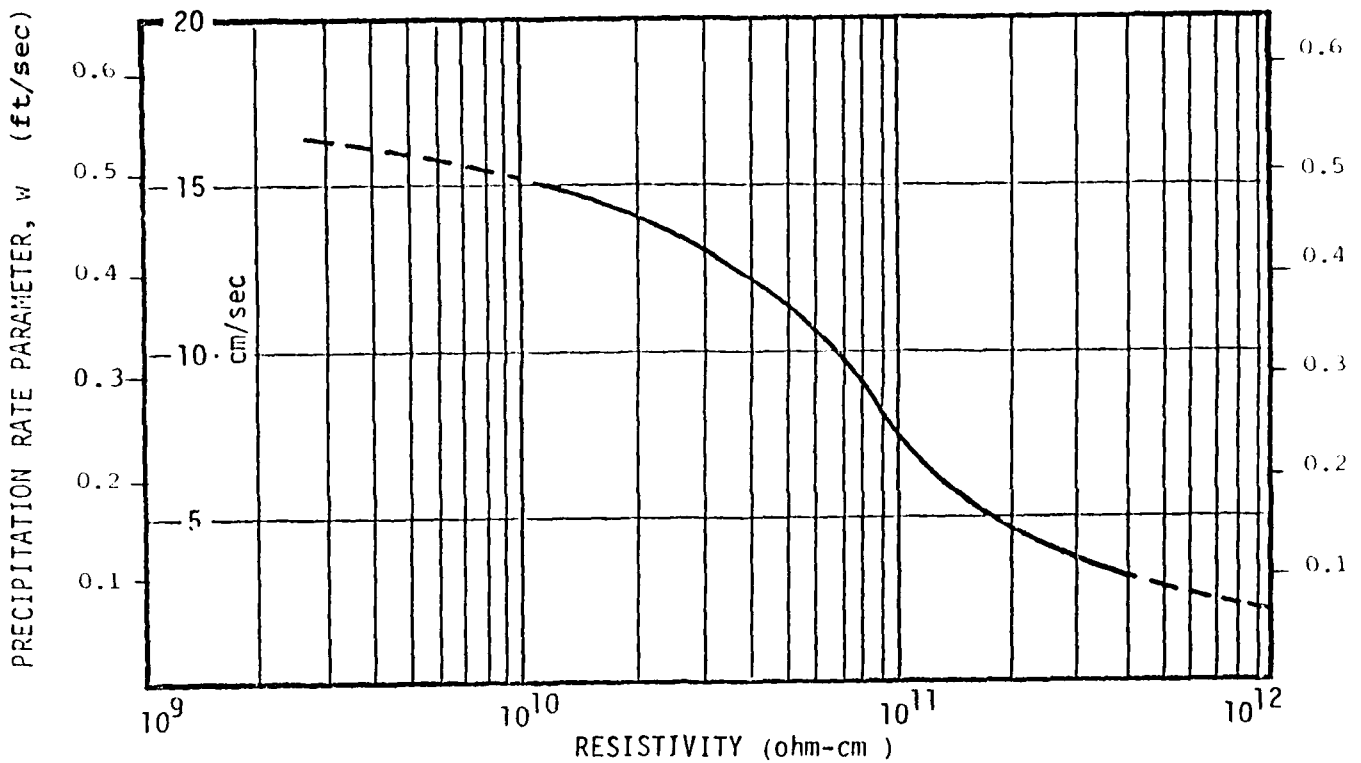

FIGURE 2-24

RELATIONSHIP BETWEEN PRECIPITATION RATE PARAMETER AND RESISTIVITY[25]

If it is impractical to select w on the basis of resistivity, other factors can often be used. In fly-ash precipitators, resistivity is influenced by the sulfur content of the fuel, and relationships have been developed between precipitation rate parameter w and percent sulfur. Figure 2-25 shows a curve developed by Ramsdell[26] for a group of fly-ash precipitators burning coals of varying sulfur contents. On a statistical basis, the precipitation rate can be predicted within reasonable accuracy. However, on an individual installation, the variations are too great to predict w with acceptable precision based on sulfur content alone. In many instances, the only information available is the sulfur content of the coal, and designs are sometimes based solely upon this parameter.

Particle size of the dust is a very important consideration in determining the value of w for design purposes. Referring to Table 2-5, the variations in w between the various application areas are due largely to particle size variations. In cement kilns, the alkali content of the raw material alters the size distribution of the dust. Metallurgical operations characteristically produce smaller size dusts from high temperature melting operations. Size of dusts from recovery boilers in pulp and paper mills can change with temperature. These factors result in variations in precipitation rate parameters between the various applications and within the same application area.

2.4.3.2 Alternate Methods of Specifying Collection Surface Area

The collection surface required for a given gas flow and efficiency is usually computed from Equation 2.21, using the parameter w. For design purposes, it is convenient to express the parameter A/Q in $ft^2/1,000$ acfm. The relation between A/Q and efficiency for a range of values of w is shown[25] graphically in Figure 2-26. Practical values of SCA (specific collection area) usually range between about 100 and 800 $ft^2/1,000$ acfm. As an example, the SCA required for 99 percent efficiency and w of 0.15 ft/sec, as read from Figure 2-26, is 510 $ft^2/1,000$ acfm.

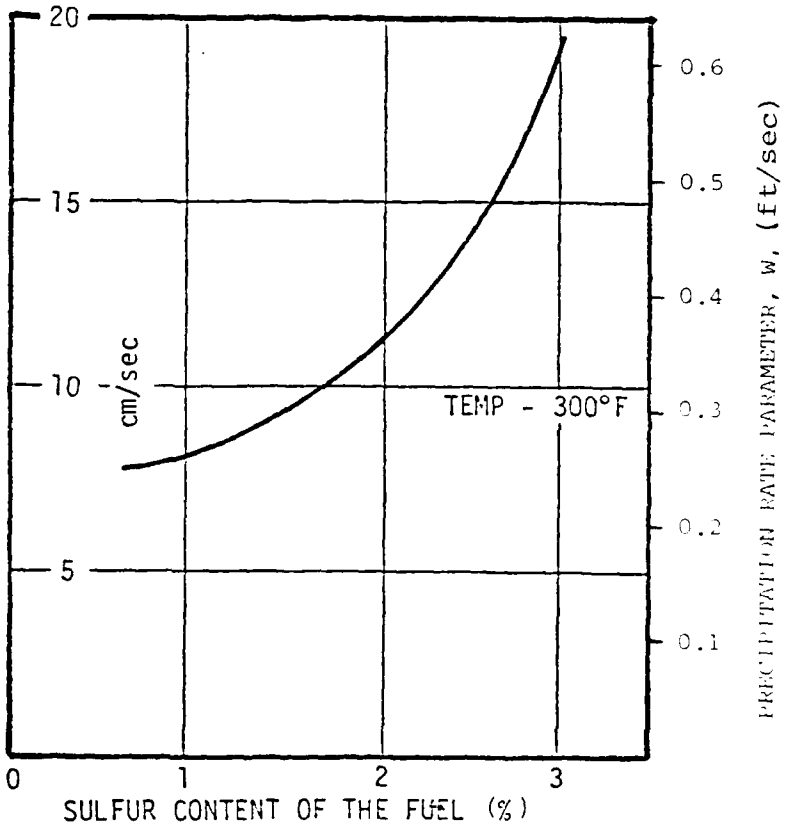

FIGURE 2-25

RELATIONSHIP BETWEEN PRECIPITATION RATE PARAMETER AND SULFUR CONTENT FOR ELECTRIC UTILITY INSTALLATIONS AT A TEMPERATURE OF 300°F [26]

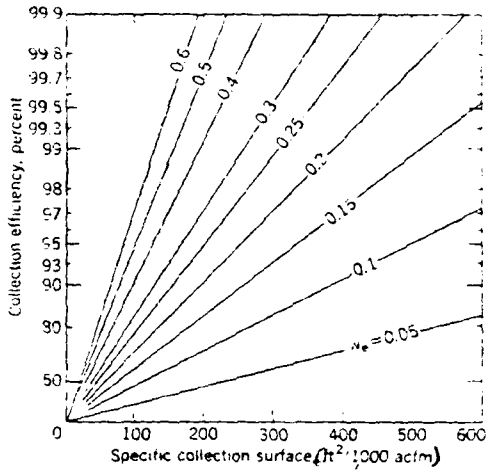

FIGURE 2-26
PRECIPITATOR EFFICIENCY VERSUS
SPECIFIC COLLECTION AREA AND
PRECIPITATION RATE, w (ft/sec)[25]

The experimentally derived relationships between collection efficiency and SCA for various coal sulfur contents can also be used to estimate necessary collection surface area. These relationships are depicted[27] in Figure 2-27.

2.4.3.3 Energy Requirements

The power required for a particular application can be determined either empirically or by the relationships between collection efficiency, current density, secondary voltage, and plate area. Input power typically ranges from 50 to 150 watts/1,000 acfm, depending upon the dust resistivity and desired collection efficiency. The relationship between the precipitation rate parameter and the power density is shown[27] in Figure 2-28 and that between collection efficiency and corona power is shown[27] in Figure 2-29.

In addition to power required for electrical energization, fan power is also required due to the pressure drop through the precipitator. Fan power can be computed from the equation

$$P = \frac{746\ Q\Delta P}{6356\ \eta} \qquad (2.22)$$

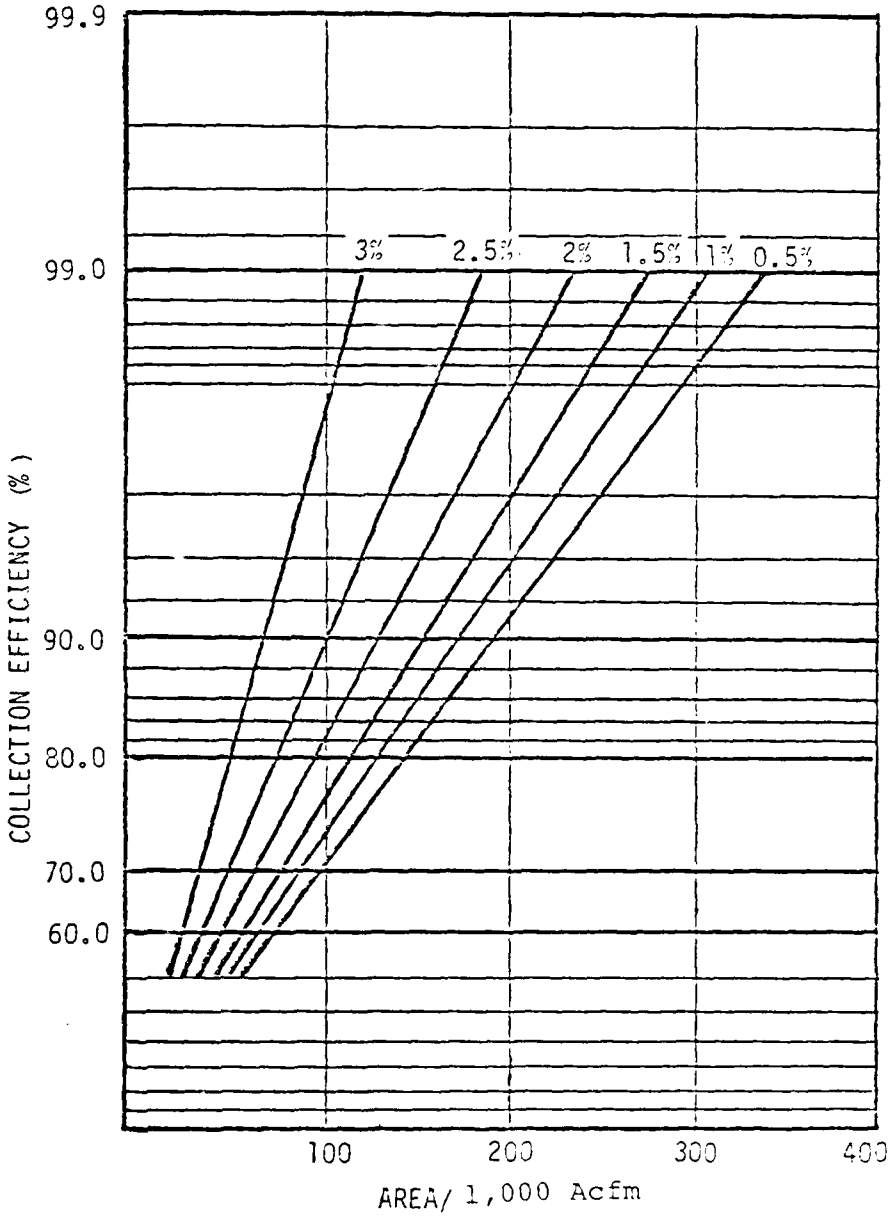

FIGURE 2-27

RELATIONSHIP BETWEEN COLLECTION EFFICIENCY AND COLLECTING SURFACE AREA TO GAS FLOW RATIO FOR VARIOUS COAL SULFUR CONTENTS [27]

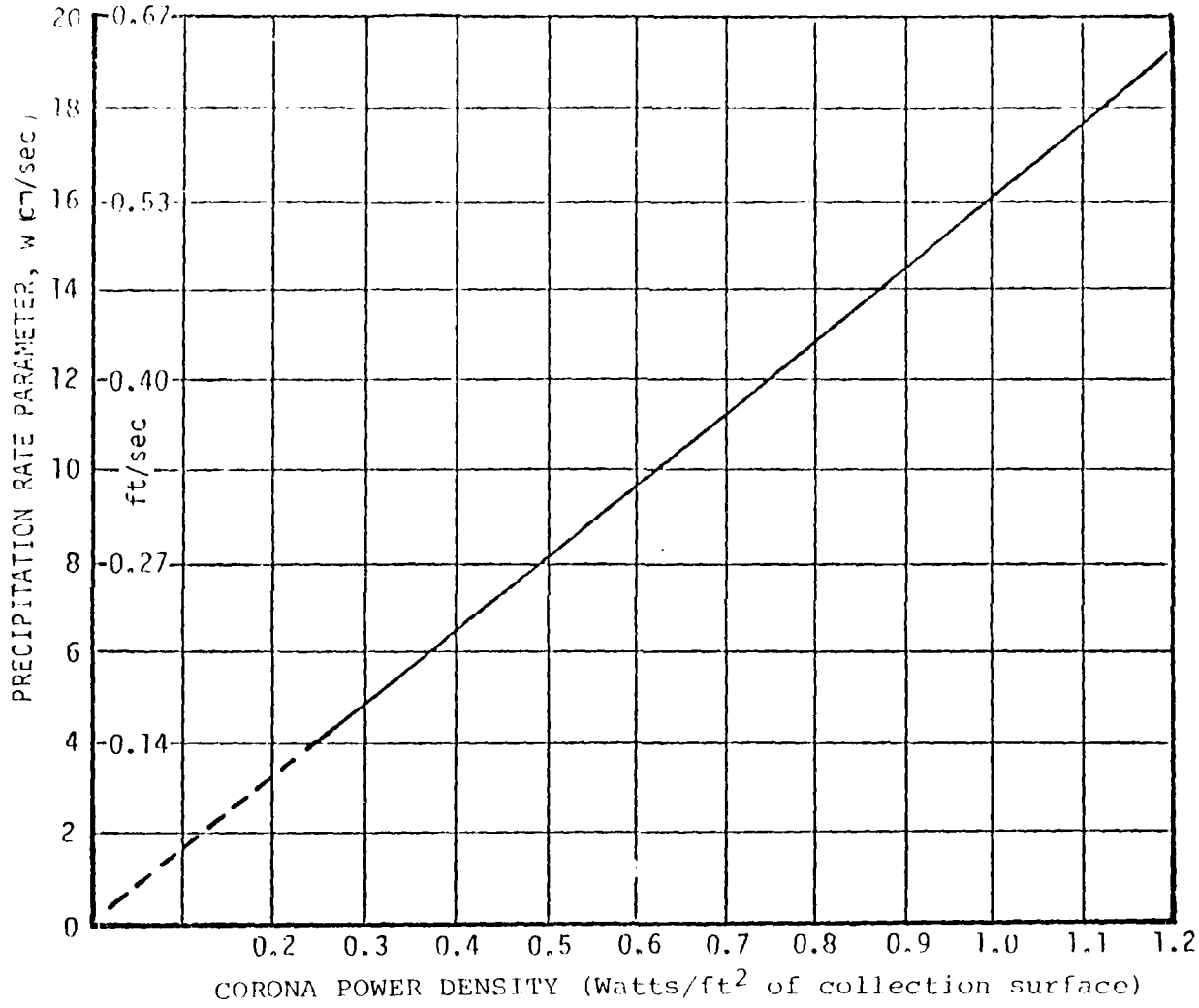

FIGURE 2-28

LINEAR RELATIONSHIP BETWEEN PRECIPITATION RATE PARAMETER AND POWER DENSITY FOR FLY ASH COLLECTORS [27]

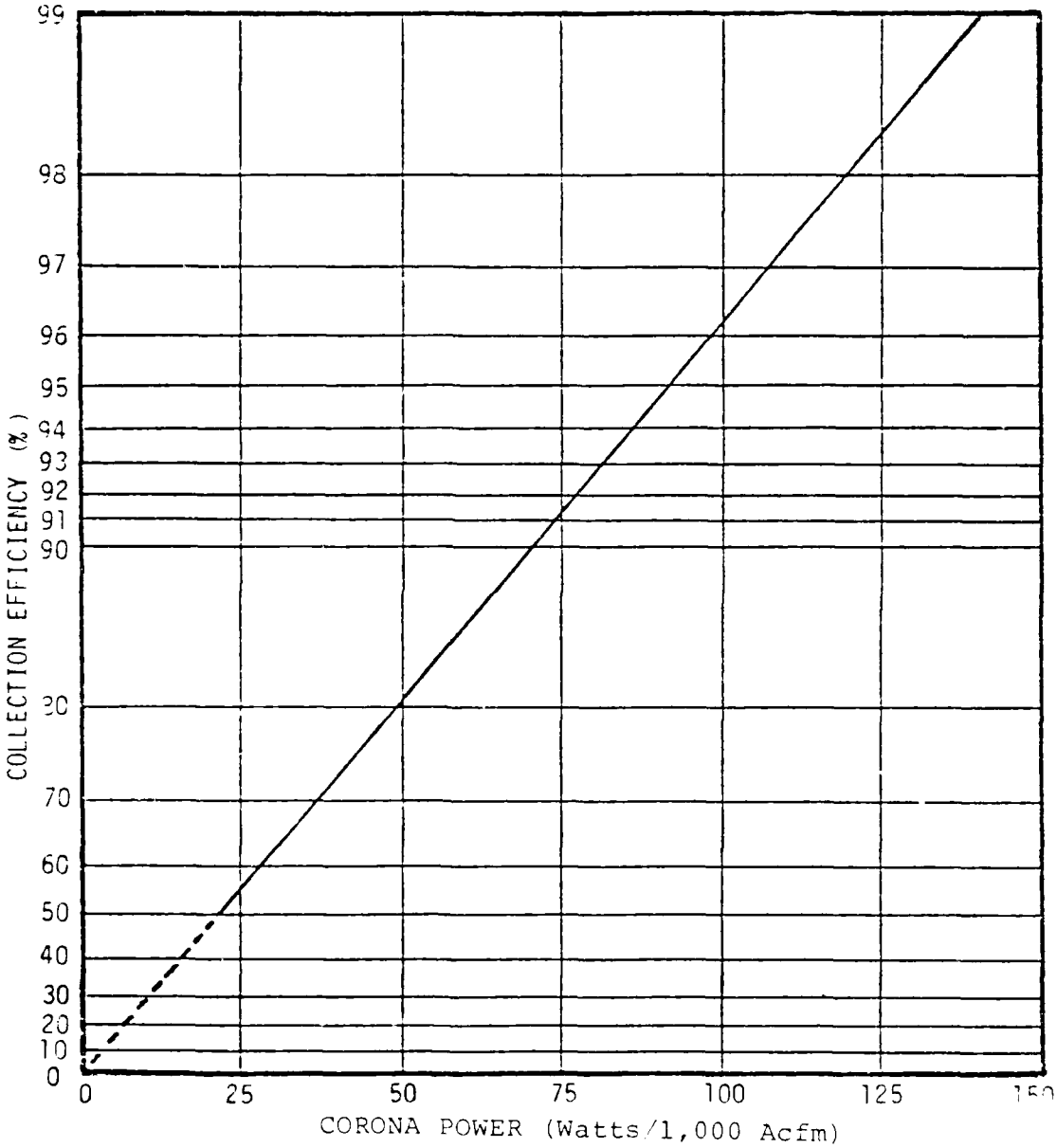

FIGURE 2-29

RELATIONSHIP BETWEEN COLLECTION EFFICIENCY AND CORONA POWER FOR FLY-ASH PRECIPITATORS [27]

where P = power, watts
Q = gas flow rate, acfm
ΔP = pressure drop, in water gauge (assumed to be 0.5)
η = combined fan and motor efficiency (assumed to be 0.40).

The sum of the power requirements for energization and for overcoming the pressure drop is usually about 0.3 percent of total plant output.

2.4.3.4 Electrical Sectionalization

The maximum voltage at which a given field can be maintained depends on the properties of the gas and the dust being collected. These parameters may vary from one point to another within the precipitator, as well as with time. In order to keep each section of the precipitator working at high efficiency, a high degree of sectionalization is recommended. This means that many separate power supplies and controls will produce better performance on a precipitator of a given size than if there are only one or two independently controlled sections. This is particularly true if high efficiencies are required.

Modern precipitators have voltage control devices that automatically limit precipitator power input. A well-designed automatic control system tends to keep the voltage level at approximately the value needed for optimum particle charging by the corona current.

The need for series sectionalization in a precipitator arises mainly because power input needs differ at various locations in a precipitator. In the inlet sections of a precipitator, concentrations of particulate matter will be relatively heavy. This requires a great deal of power input in order to generate the corona discharge required for optimal particle charging: heavy concentrations of dust particles tend to suppress corona current. On the other hand, in the downstream sections of a precipitator dust concentrations will be lighter. As a consequence, corona current will flow more freely and particle charging will tend to be limited by excessive sparking, more so here than in inlet sections of the precipitator. Hence, excessive sparking is more likely to occur first in downstream sections; if the precipitator has only a single power set, then this sparking, under spark rate-limited control, will limit power input to the entire precipitator, including the inlet sections. This will result in insufficient power being supplied to the discharge electrodes in the inlet sections, with a consequential fall in precipitator collection efficiency in the inlet sections of the precipitator.

A remedy for this situation is to divide the precipitator into a series of independently energized electric bus sections (see Figure 2-30).[15] Each bus section has its own transformer rectifier, voltage stabilization controls, and high-voltage conductors that energize the discharge electrodes within that section. This would allow greater particulate precipitation than in the previously described underpowered inlet sections.

Parallel sectionalization (see Figure 2-30) provides the means for coping with different power input needs due to uneven dust and gas distributions that usually occur across the inlet face of a precipitator. Nevertheless, the gains in collection efficiency from parallel sectionalization are likely to be very small.

Typical limits of sectionalization for utility plants are given below:

Efficiency	Number Fields Deep	Degree of Sectionalization
95	1	Low
95-99	2	Moderate
99-99.5	3	High
99.5+	4	High

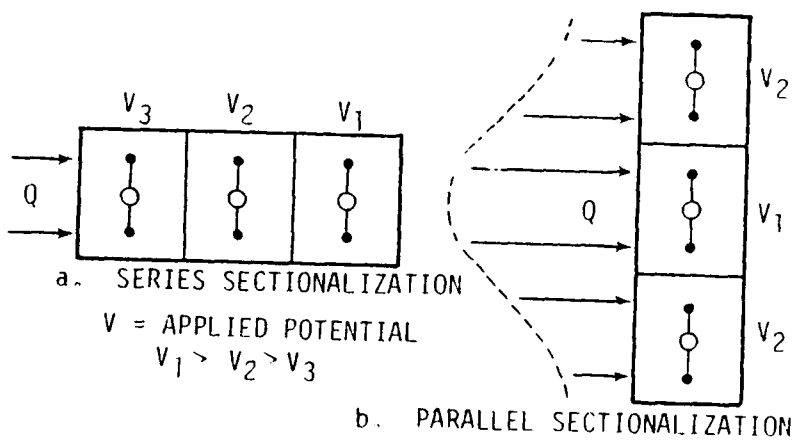

FIGURE 2-30

TYPICAL ARRANGEMENTS FOR SEPARATELY ENERGIZED BUS SECTIONS[15]

The design curves for sizing and electrical bus sectionalization of precipitators developed by Ramsdell[26] are shown in Figure 2-31.

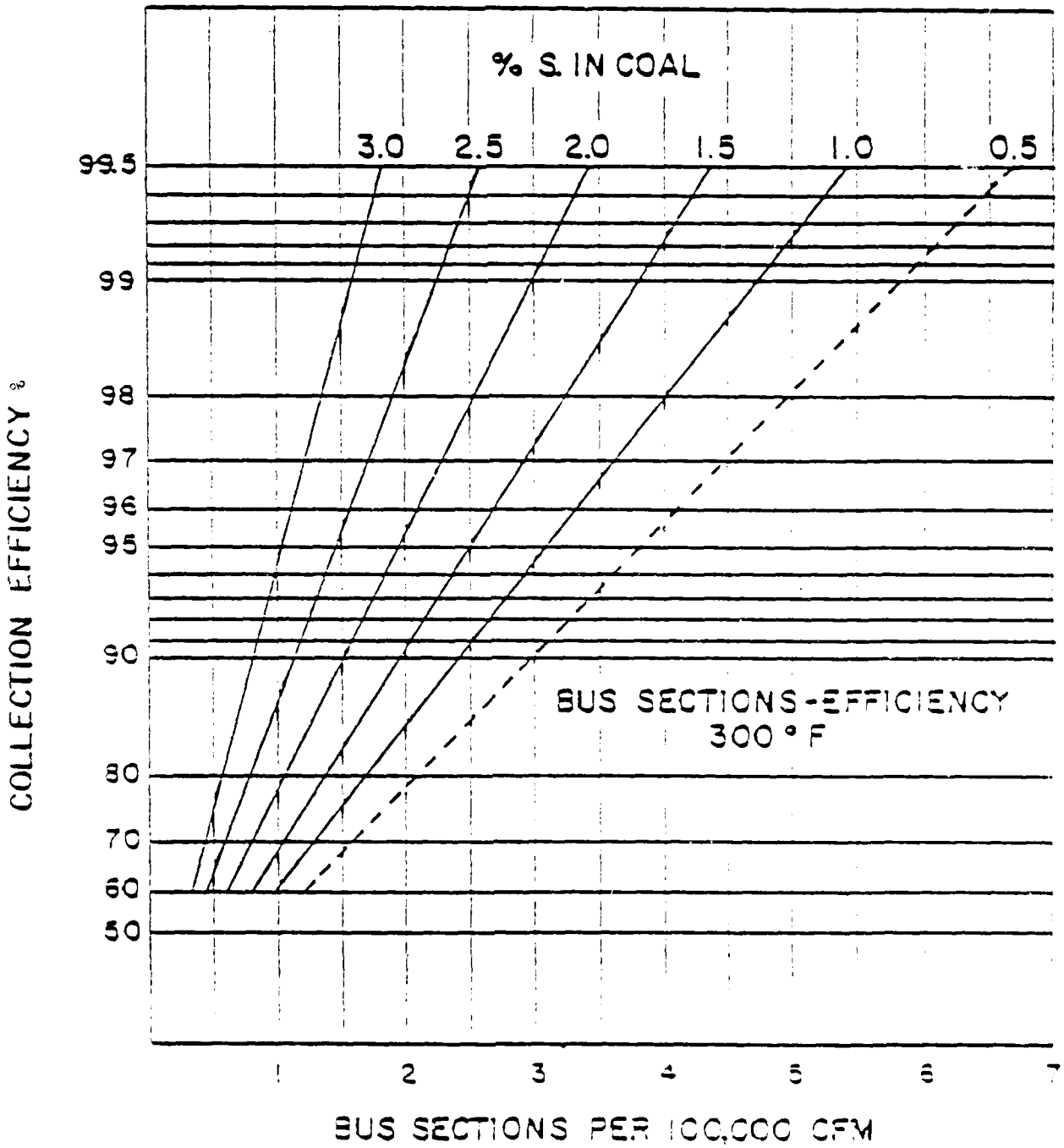

FIGURE [illegible]

COLLECTION EFFICIENCY VS BUS SECTIONS PER 100,000 CFM [26]

2.4.3.5 Gas Velocity

Although the localized gas velocities in the collection zones of a precipitator vary over a rather wide range, even for good conditions, it is convenient for design purposes to use a hypothetical average value calculated from the gas flow and the cross section of the precipitator. The cross section is taken as the open area for gas flow between the collecting plates, disregarding the plate baffles. The primary importance of this hypothetical gas velocity is its relation to rapping and reentrainment losses. Above some critical velocity, these losses tend to increase rapidly because of the aerodynamic forces on the particles. The critical velocity depends on the quality of gas flow, plate configuration, precipitator size, and other factors, but for most fly-ash precipitators, it does not exceed 6 to 8 ft/sec. This sets a design limit on gas velocity of no more than 5 or 6 ft/sec for high-efficiency fly-ash precipitators.

2.4.3.6 Aspect Ratio

Aspect ratio is defined as the ratio of the length of the ducts to the height of the ducts. It is important in precipitator design because of its effect on rapping loss. Collected dust released from the plates is carried forward by the flow of the gas. If the ducts are too short compared to their height, some of the falling dust will be carried out of the precipitator before it reaches the hoppers, thereby substantially increasing the dust loss. The time required for released dust clumps to fall from the top of a 35-ft high plate, for example, can be several seconds; this is sufficient time for some of the dust to be carried out of the precipitation zones when gas velocities exceed about 5 ft/sec and the duct length is less than 25 or 30 ft.

Aspect ratios used in practice generally range from about 0.5 to 1.5. For efficiencies of 99 percent or higher, the aspect ratio should be at least 1.0 to 1.5 to minimize carry-through of collected dust.

2.4.4 Advantages and Disadvantages of Electrostatic Precipitation

The use of electrical precipitators for the collection of air contaminants has grown because of many inherent advantages, some of which are listed below:

1. High efficiency can be attained. Efficiency may exceed 99 percent in some cases.

2. Very small particles can be collected. There is no theoretical lower limit to the size of a particle that can be collected.

3. Dusts may be collected dry for recovery of valuable material.

4. Pressure and temperature drops are small. The pressure drop through an electrical precipitator seldom exceeds 0.5 inch vertical water column.

5. Precipitators are normally designed to operate continuously with little maintenance over long periods of time.

6. There are very few, if any, moving parts, which tends to reduce the maintenance required.

7. Precipitators can be used at high temperatures. Temperatures up to about 700°F are normal. Special designs have been used for temperatures as high as 1,300°F, but ordinarily the temperature does not exceed 1,000°F.

8. Precipitators can be used to collect acid and tar mists, which are difficult, if not impossible, to collect by other methods.

9. Extremely corrosive materials can be collected with special construction.

10. Collection efficiency may be adjusted to suit the application by increasing the unit size.

11. Very large gas flow rates can be handled.

12. The power requirements for flow handled are low.

Electrical precipitators are by no means a panacea for air pollution problems. In many cases, disadvantages far outweigh the advantages. Some of the drawbacks are:

1. Initial cost is high. In most cases, the investment is greater than that required for any other form of air pollution control.

2. Precipitators are not easily adaptable to variable conditions. Automatic voltage control helps to a great extent, but precipitators are most efficient when operating conditions remain constant.

3. Some materials are extremely difficult to collect in an electrical precipitator because of extremely high or low resistivity or other causes. In some cases, this factor alone makes the use of electrical precipitation uneconomical, if not physically impossible.

4. Space requirements may sometimes be greater than those for a baghouse. In general, this is true only when high collection efficiency is required for materials difficult to collect by precipitation.

5. Electrical precipitation is not applicable to the removal of materials in the gaseous phase.

6. The use of a precleaner, generally of the cyclonic type, may be required to reduce the dust load on a precipitator.

7. Special precautions are required to safeguard personnel from the high voltage.

2.4.5 Illustrative Examples

2.4.5.1 Example No. 1

Design an electrostatic precipitator for a pulverized fuel boiler with the following given conditions:

(1) Dust resistivity: 7×10^{10} ohm-cm
(2) Gas temperature: 300°F
(3) Gas volume: 750,000 acfm
(4) Sulfur content: 1.8%

Solution:

Step 1.

The following information is given in the problem statement.

Dust resistivity = 7×10^{10} ohm-cm
Gas temperature = 300°F
Gas volume, Q = 750,000 acfm
Sulfur content = 1.8%

Step 2.

Precipitation rate parameter, w, is selected using Figure2-24.

For dust resistivity = 7×10^{10} ohm-cm
w = 0.32 ft/sec or 19.2 ft/min

Step 3.

The precipitator will be designed for 99 percent collection efficiency.

Step 4.

From the Deutsch-Anderson equation (Eq. 2-21)

$$\eta = 100 - \exp\left(-\frac{A}{Q} w\right)$$

Or rearranging,

$$A = \frac{Q}{W} \ln\left(\frac{100}{100-\eta}\right)$$

$$= \frac{750{,}000 \text{ acfm}}{19.2 \text{ ft/min}} \ln\left(\frac{100}{100-99}\right)$$

$$= 179{,}890 \text{ ft}^2$$

Step 5.

From Figure 2-29, power required for 99 percent efficiency is 140 watts/1,000 cfm.

Total power requirement

= (140 watts/1,000 cfm) (750,000 acfm)

= 105,000 watts

Step 6.

From Figure 2-31, for 99 percent collection efficiency, number of bus sections required is 3.5/100,000 cfm.

Bus sections required = (3.5/100,000 cfm) (750,000 acfm)

= 26

2.4.5.2 Example No. 2

An electrostatic precipitator is to be used for fly-ash removal from a 500,000 acfm gas stream at 300°F. The coal fired contains 1.8 percent sulfur (by weight). The fly-ash resistivity has been determined to be approximately 6×10^{10} ohm-cm. Estimate the precipitator design parameter to achieve 99 percent efficiency.

Solution:

Step 1.

The following information is given in the example:

Gas flow rate, Q = 500,000 acfm
Gas temperature = 300°F
Sulfur content = 1.8% (by weight)
Dust resistivity = 6×10^{10} ohm-cm

Step 2.

Precipitation rate parameter, w, for resistivity = 6×10^{10} ohm-cm, is

w = 0.34 ft/sec or 20.4 ft/min (From Figure 2-24)

Step 3.

Desired collection efficiency = 99% (given)

Step 4.

$$\eta = 100 - \exp\left(-\frac{A}{Q} w\right) \quad \text{(Eq. 2-21)}$$

Rearrangement gives

$$A = \frac{Q}{W} \ln\left(\frac{100}{100 - \eta}\right)$$

$$= \frac{500{,}000 \text{ acfm}}{20.4 \text{ ft/min}} \ln\left(\frac{100}{100-99}\right)$$

$$= 112{,}870 \text{ ft}^2$$

Step 5.

From Figure 2-29, power required for 99 percent efficiency is 140 watts/1,000 cfm.

Total power requirement:

= (140 watts/1,000 cfm) (500,000 cfm)
= 70,000 watts

Step 6.

From Figure 2-31, for 99 percent collection efficiency, number of bus sections required is 3.5/100,000 cfm.

Bus sections required = (3.5/100,000 cfm) (500,000 acfm)

= 18

Note that the collection surface area required could also have been obtained from Figure 2-27. For 99 percent efficiency, and 1.8 percent sulfur coal, approximately 225 ft^2/1,000 acfm is required or

(225 ft^2/1,000 acfm) (500,000 acfm) = 112,500 ft^2

Figure 2-28 could also have been used to determine collection electrode area. For a precipitation rate parameter of 0.34 ft/sec, a power density of 0.625 watts/ft^2 is selected. Hence, the area is given by:

$$(70{,}000 \text{ watts})\left(\frac{1}{0.625 \text{ watts/ft}^2}\right) = 112{,}000 \text{ ft}^2$$

Since the curves relating the various factors are mutually compatible, it makes little difference which approach is finally chosen to arrive at a design estimate.

2.5 FABRIC FILTERS

2.5.1 Introduction

The fabric filter is capable of providing high collection efficiencies for particles as small as 0.5 μm and will remove a substantial quantity of those particles as small as 0.01 μm. In its simplest form, the industrial fabric filter consists of a woven or felted fabric through which dust-laden gases are forced. A combination of factors results in the collection of particles on the fabric fibers. When woven fabrics are used, a dust cake eventually forms; this, in turn, acts predominantly as a sieving mechanism. When felted fabrics are used, this dust cake is minimal or nonexistent. Instead, the primary filtering mechanisms are a combination of inertial forces, impingement, etc., as related to individual particle collection on single fibers. These are essentially the same mechanisms that are applied to particle collection in wet scrubbers, wherein the collection media is in the form of liquid droplets rather than solid fibers.

As particles are collected, the pressure drop across the fabric filtering media increases. Due in part to fan limitations, the filter must be cleaned at predetermined intervals. Dust is removed from the fabric by gravity and/or mechanical means. The fabric filters or bags are usually tubular or flat. The structure in which the bags hang is frequently referred to as a baghouse. The number of bags in a baghouse may vary from a couple to several thousand. Quite often when great numbers of bags are involved, the baghouse is compartmentalized so that one compartment may be cleaned while others are still in service.

2.5.2 Description of Available Equipment

The basic filtration process may be conducted in many different types of fabric filters in which the physical arrangement of hardware and the method of removing collected material from the filter media will vary. The essential differences may be related, in general, to:

1. type of fabric
2. cleaning mechanism
3. equipment geometry
4. mode of operation.

Depending on the above factors, equipment will follow one of three systems, as shown[15] in Figure 2-32. Bottom-feed units are characterized by the dust-laden gas introduced through the baghouse hopper and then to the interior

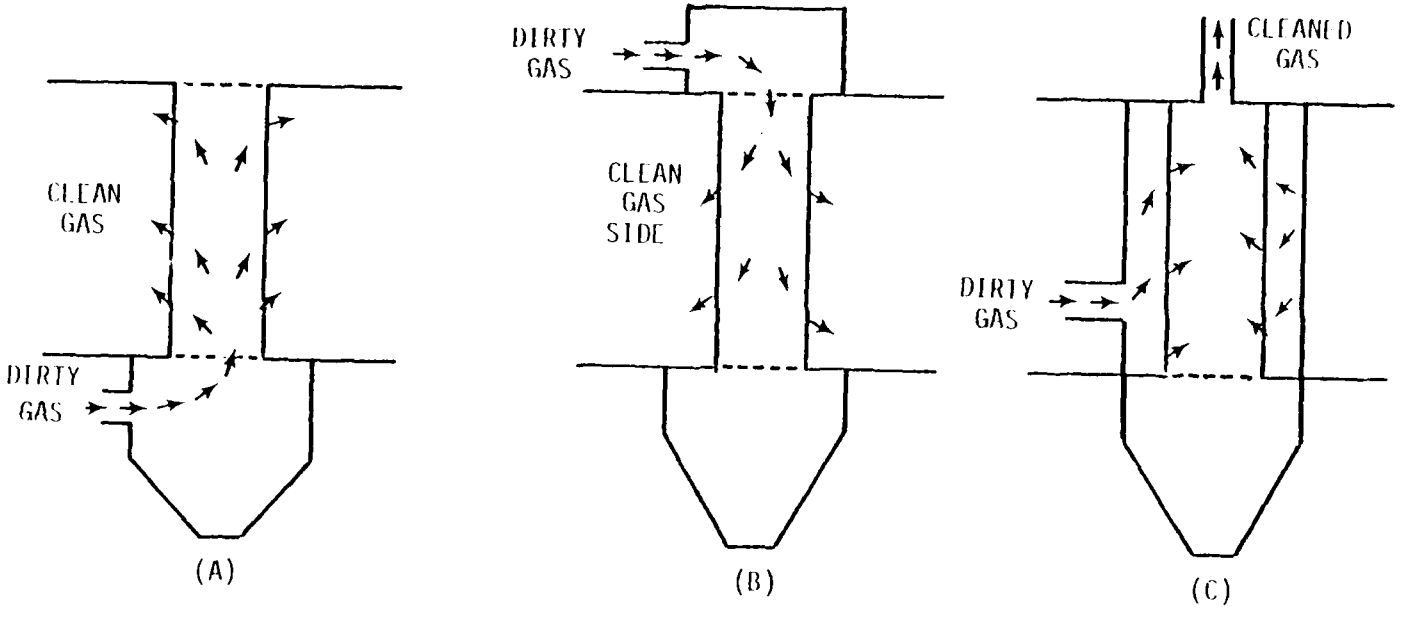

FIGURE 2-32

TYPES OF FILTERING SYSTEMS[15]

(a) bottom feed; (b) top feed; (c) exterior filtration

of the filter tube. In top-feed units, dust-laden gas enters the top of the filter tubes. In exterior filtration, the gas passes from the outside of the filters to the interior or clean-air side. When the gas flow is from inside the bag to outside, by virtue of the pressure differential, the internal area of the filter element will be open and self-supporting. The unsupported filter elements are tubular. When the filtration process is reversed, with the gas flow from outside the bag to inside, it is necessary to support the media against the developed pressures so that the degree of collapse is controlled. Supported filter elements are either of the tubular or envelope shape.

Gases to be cleaned can be either "pushed" or "pulled" through the baghouse. In the pressure system (push through), the gases may enter through the cleanout hopper in the bottom or through the top of the bags. In the suction type (pull through), the dirty gases are forced through the inside of the bag and exit through the outside. Figures 2-33 through 2-35 depict these flows.[15]

Baghouse collectors are available for either intermittent or continuous operation. Intermittent operation is employed where the operational schedule of the dust-generating source permits halting the gas cleaning function at periodic intervals (regularly defined by time or by pressure differential) for removal of collected material from the filter media (cleaning). Collectors of this type are primarily utilized for the control of small-volume operations such as grinding and polishing and for aerosols of a very coarse nature. For most air pollution control installations and major dust control problems, however, it is desirable to use collectors that allow for continuous operation. This is accomplished by arranging several filter areas in a parallel flow system and cleaning one area at a time according to some preset mode of operation (see Figures 2-36 and 2-37).[15]

Baghouses may also be characterized and identified according to the method used to remove collected material from the bags. Particle removal can be accomplished in a variety of ways, including shaking the bags, reversing the direction of air flow through the bags, blowing a jet of air on the bags from a reciprocating manifold, or rapidly expanding the bags by a pulse of compressed air. Table 2-6 lists the standard cleaning methods[28] along with a number of characteristics frequently associated with the various methods. In general, the various types of bag cleaning methods can be divided into those involving fabric flexing and those involving a reverse flow of clean air.

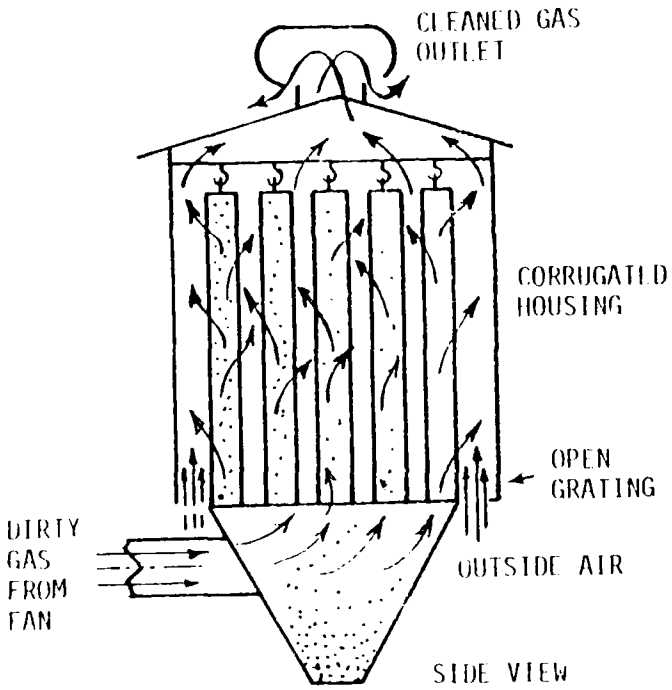

FIGURE 2-33

OPEN PRESSURE BAGHOUSE[15]

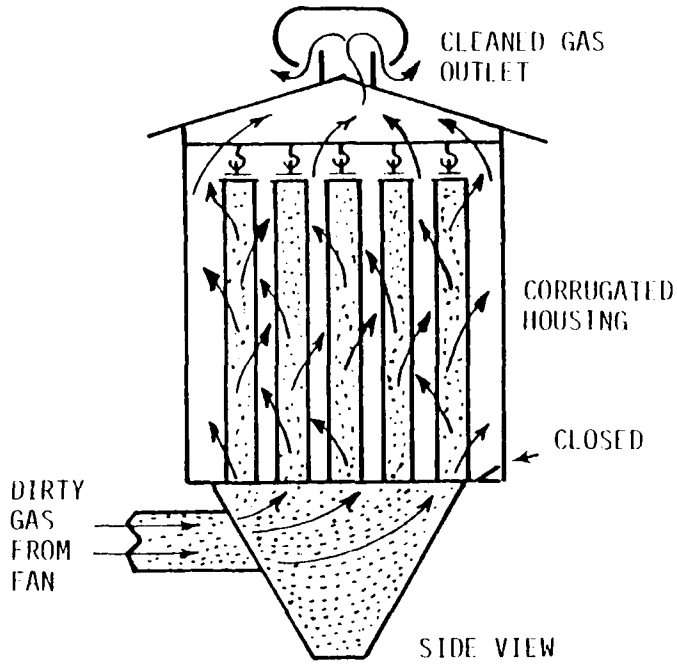

FIGURE 2-34

CLOSED PRESSURE BAGHOUSE[15]

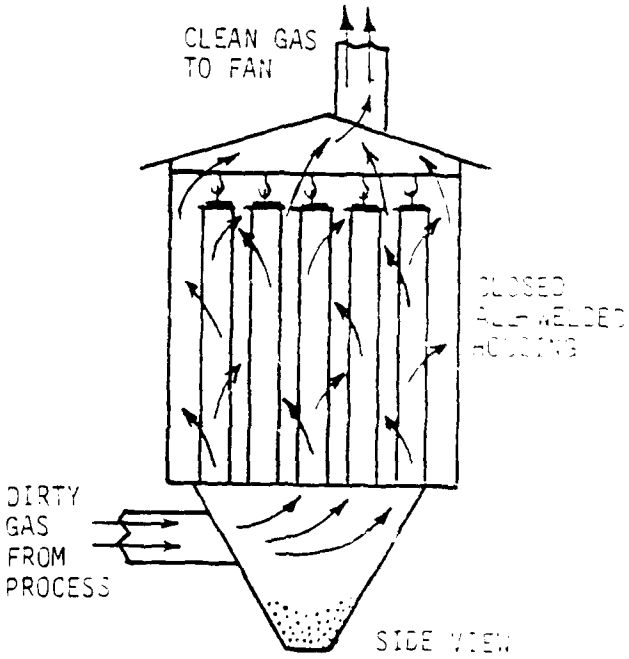

FIGURE 2-35

CLOSED SUCTION BAGHOUSE[15]

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DEVELOPMENT OF DESIGN REVIEW PROCEDURES FOR ARMY AIR POLLUTION --ETC(U)
JUL 80 A J BUONICORE, J P BILOTTI DAMD17-79-C-9051

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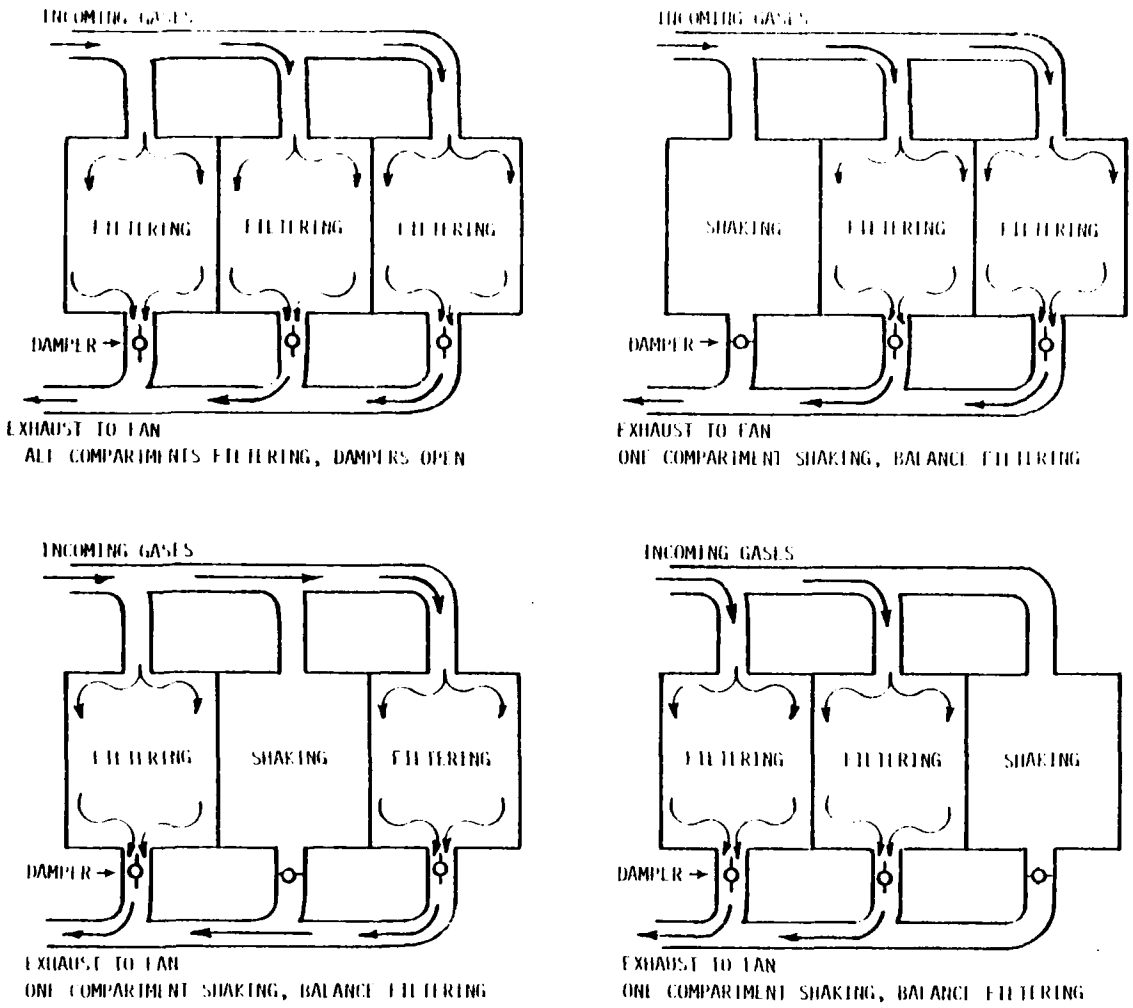

FIGURE 2-36

TYPICAL PARALLEL FLOW SYSTEM FOR A CONVENTIONAL MULTI-COMPARTMENT BAGHOUSE[15]

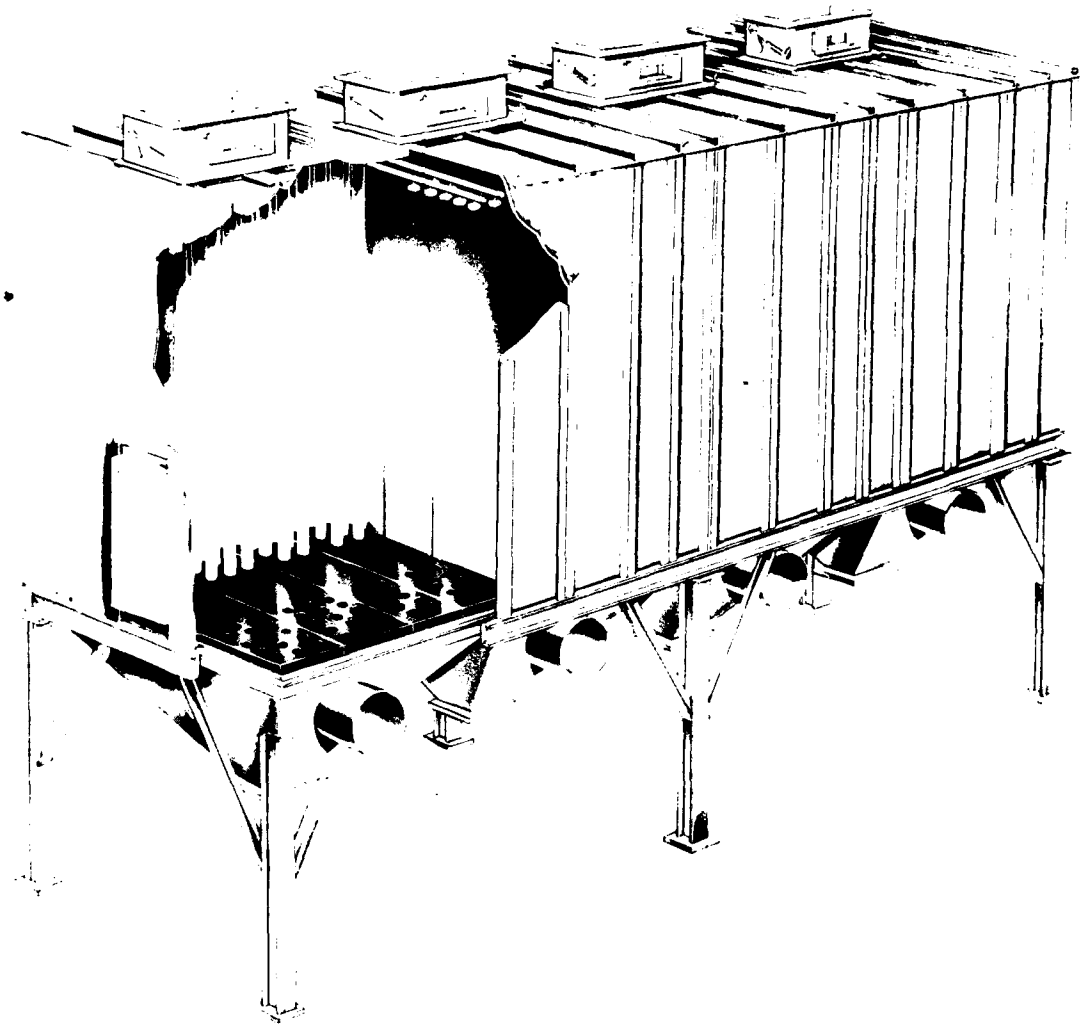

FIGURE 2-37

CONTINUOUS OPERATING COMPARTMENTALIZED BAGHOUSE COLLECTOR[15]

TABLE 2-6

COMPARISON OF BAG CLEANING METHODS [28]

Cleaning Method	Uniformity of Cleaning	Bag Attrition	Equipment Ruggedness	Type Fabric	Filter Velocity	Apparatus Cost	Power Cost	Dust Loading	Submicron Efficiency
Shake	A	A	A	Woven	A	A	L	A	G
Reverse flow, no flexing	G	L	G	Woven	A	A	M-L	A	G
Reverse flow, with collapse	A	H	G	Woven	A	A	M-L	A	G
Pulse-compartment	G	L	G	Felt,woven	H	H	M	H	H
Pulse-bags	A	A	G	Felt,woven	H	H	H	VH	H
Reverse-jet	VG	A-H	L	Felt,woven	VH	H	H	H	VH
Vibration, rapping	G	A	L	Woven	A	A	M-L	A	G
Sonic assist	A	L	L	Woven	A	A	M	-	G
Manual flexing	G	H	-	Felt-woven	A	L	-	L	G

Note: A = average; G = good; H = High; L = low; M = Medium; VG = very good; VH = very high.

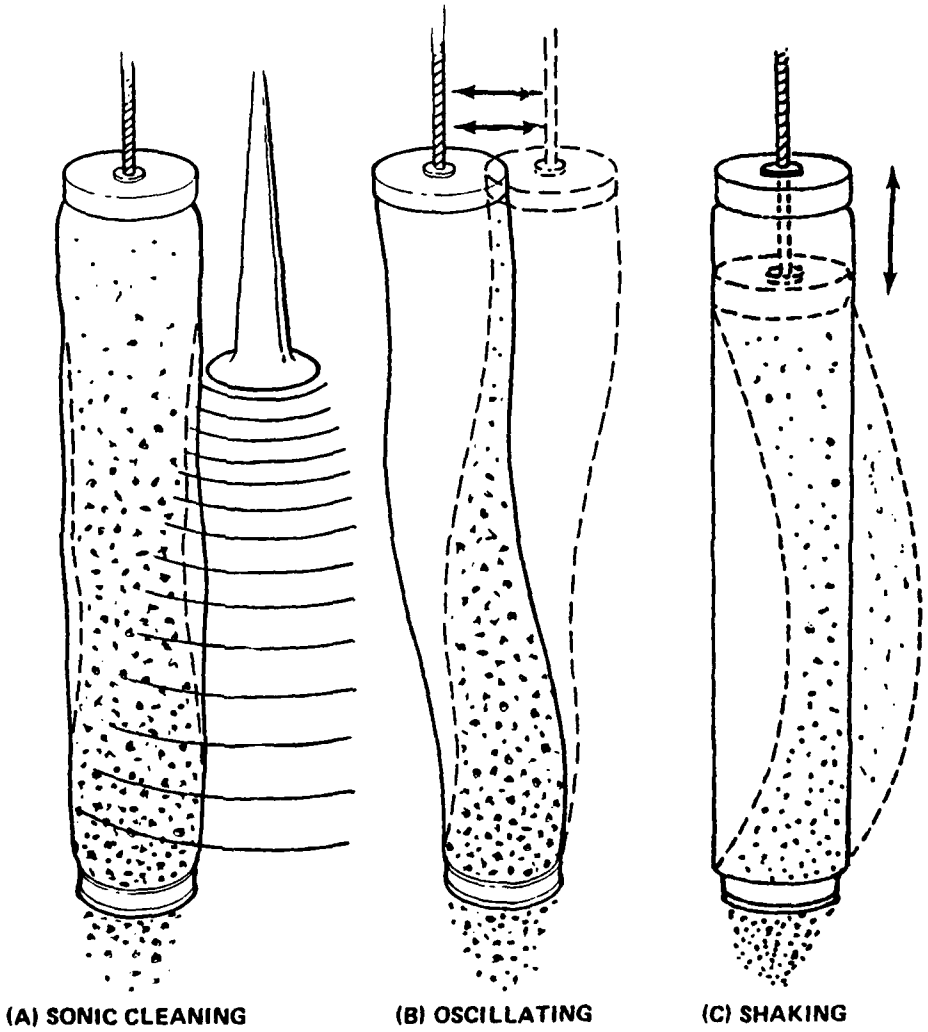

FIGURE 2-38

FLEXING BAG CLEANING METHODS [15]

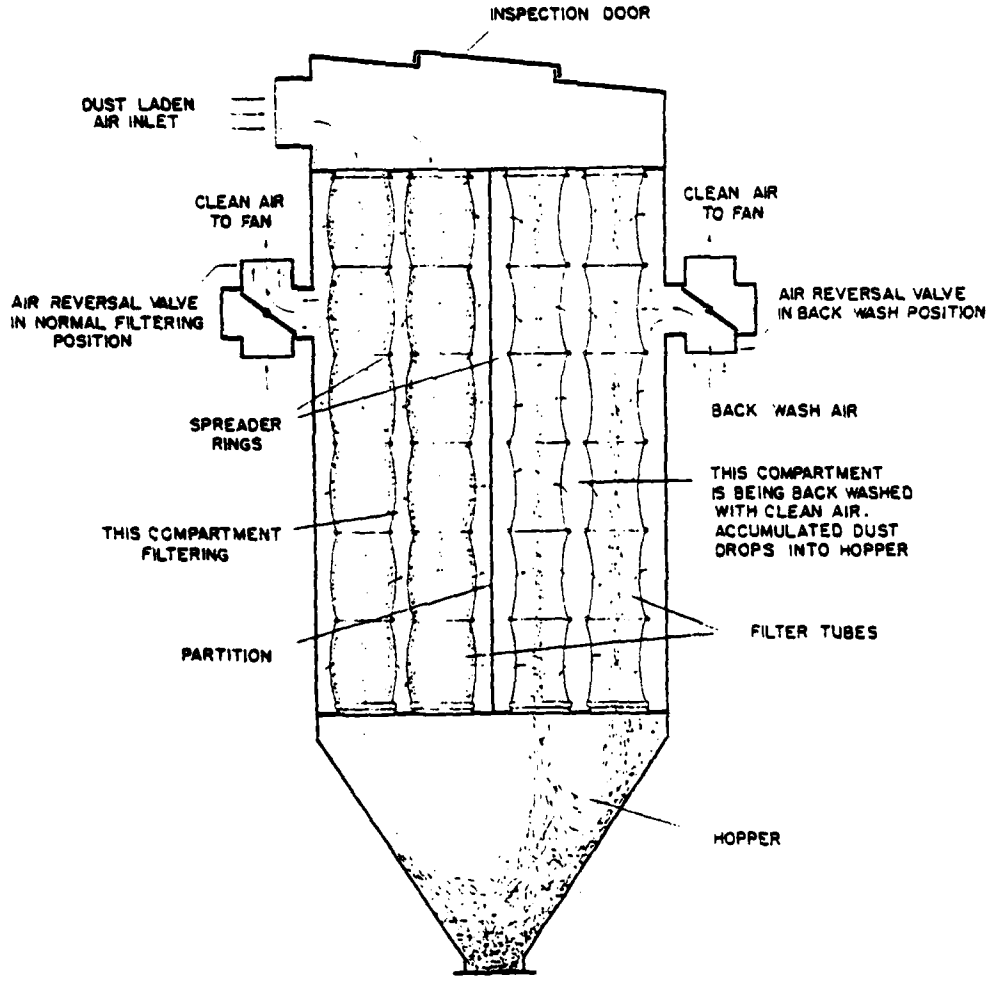

FIGURE 2-39

TYPICAL REVERSE-FLOW CLEANING BAGHOUSE[15]

Fabric flexing methods include mechanical shaking and rapping, sonic cleaning, collapse cleaning, and pressure-jet and pulse-jet cleaning. Mechanical shaking and rapping involve the use of a rocker arm-lever assembly to produce a motion at the top of the filter tubes as shown[15] in Figure 2-38. The motion may be generally horizontal (sometimes concave upwards, sometimes concave downwards), vertical, or cover a 90° arc from the bottom to the top of the swing. Vertical motion is sometimes accomplished by rapping. A variation that provides a more gentle shake than mechanical shaking is inducement of an oscillating flexing motion. The top of each bag is moved back and forth, in a relatively flat arc, causing ripples in the filter bags to dislodge the dust. This type of cleaning is best used with light, flocculent dusts. In cleaning by mechanical action, a baghouse compartment must usually be taken out of service and the filtering process interrupted. As a result, the capacity of the baghouse must be somewhat higher than would otherwise be required. A general problem in the use of mechanical shaking is that the fabric may be flexed excessively at the bottom, which is fastened, and fiber abrasion may result, thus shortening bag life.

Sonic cleaning employs sound generators that produce a low-frequency sound, causing the bags to vibrate gently. These vibrations, frequently combined with reverse air, gently loosen dust particles from the inner surface of the bags. Noise level is barely discernible outside the filter compartment. Sonic cleaning is often recommended for use with heavy, dense, and carbonaceous dusts also shown in Figure 2-38).

To clean filter bags by the collapsing technique, small reversals in pressure are created such that the pressure drop from the dirty air side to the clean air side is slightly negative. This causes the filter bag to deflate and discharge the dust cake. In some cases, the bag is slowly collapsed and "popped" open (see Figure 2-39).[15] Reverse-flow baghouses are equipped with an auxiliary fan that forces air through the bags in the direction of filtration. This backwash action collapses the bag and fractures the dust cake. When the bag is reinflated by being brought back on line, the fractured dust cake is dislodged into the hopper. If the unit operates under suction (the main fan located on the clean side of the baghouse), reducing the pressure in the baghouse may eliminate the need for an auxiliary fan.

In pulse-jet cleaning, a sharp pulse of compressed air is released, typically at the end of the filter bag, producing a combined effect of gas flow reversal in the

bag and fabric deformation. As shown[15] in Figure 2-40, the dust deposit can be dislodged with only a brief interruption of the filtering process in the bag. The time involved is only about 0.1 sec. The fabric receives a minimum of flexing.

An extension of the concept of pulse-jet cleaning is reverse jet cleaning, in which a ring of jets of moderately compressed air is continuously moved up and down the bag (see Figure 2-41),[15] usually at a rate of a few feet per minute. Abrasion and fabric wear, presented problems in the use of this method.

Pulse-jet cleaning is better applied to felted than to woven fabrics, which tend to be overly cleaned by the process and then to leak dust excessively in the next filtering cycle while the filter cake is being repaired.

For the same volume of gas cleaned, felted baghouse filters regenerated with pulsed jets of air can be operated at filtration velocities two to four times those used in baghouses equipped with woven filters cleaned by reverse air flow and mechanical shaking. On the other hand, a felted filter baghouse can have a pressure drop somewhat greater than that of a baghouse with woven bags. The power for compressing the pulsed air can also be significant, e.g., as much as that for the primary fan.

Pulse cleaning offers several advantages over either repressuring with shake assist. With off-line pulse cleaning, the gas-to-cloth ratio may be increased from 2.5:1 to 4.5:1 using fiberglass filters. This boost has a profound effect on the required plan area. Also, pulse cleaning implies that the filter bags will be suspended from a roof (tube sheet) rather than attaching to a floor (cell plate). Thus, by equipping the baghouse with an integral traveling crane, entire cartridges of bags may be extracted without physically entering the hostile environment of the baghouse. With tightening OSHA regulations, this advantage is becoming increasingly important.

An additional advantage offered by suspending the bags from a tube sheet is the ease with which a failed filter may be located and capped off. To find a failed bag, a special, inert, high-temperature fluorescent powder is injected into the breeching between the air preheater and the baghouse. An inspection of only the bag/tube sheet connection with an ultraviolet lamp identifies any failed bags. Inspection is accomplished without entering the baghouse.

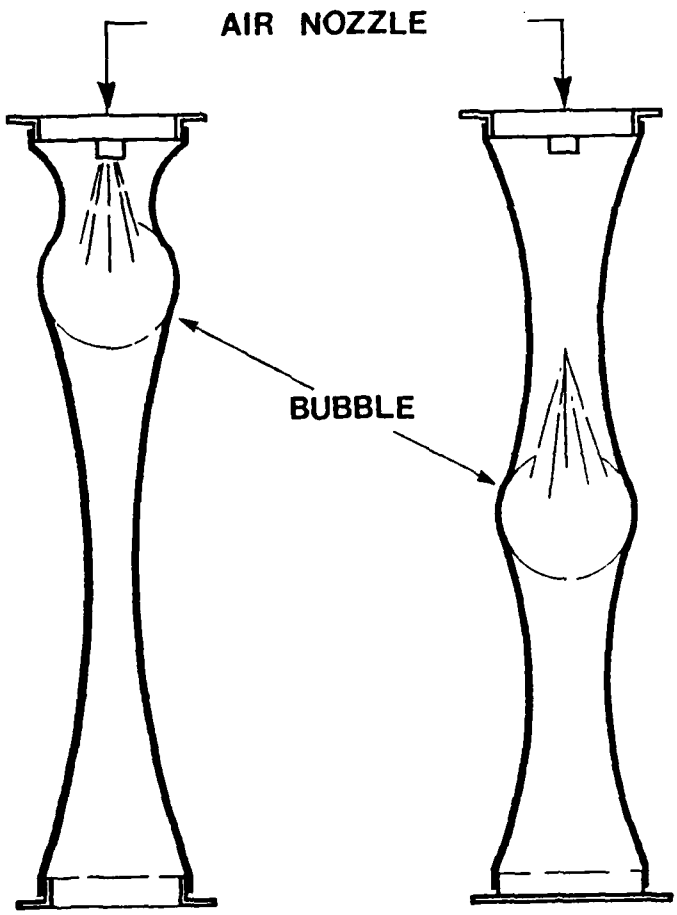

FIGURE 2-40

PULSE-JET CLEANING[15]

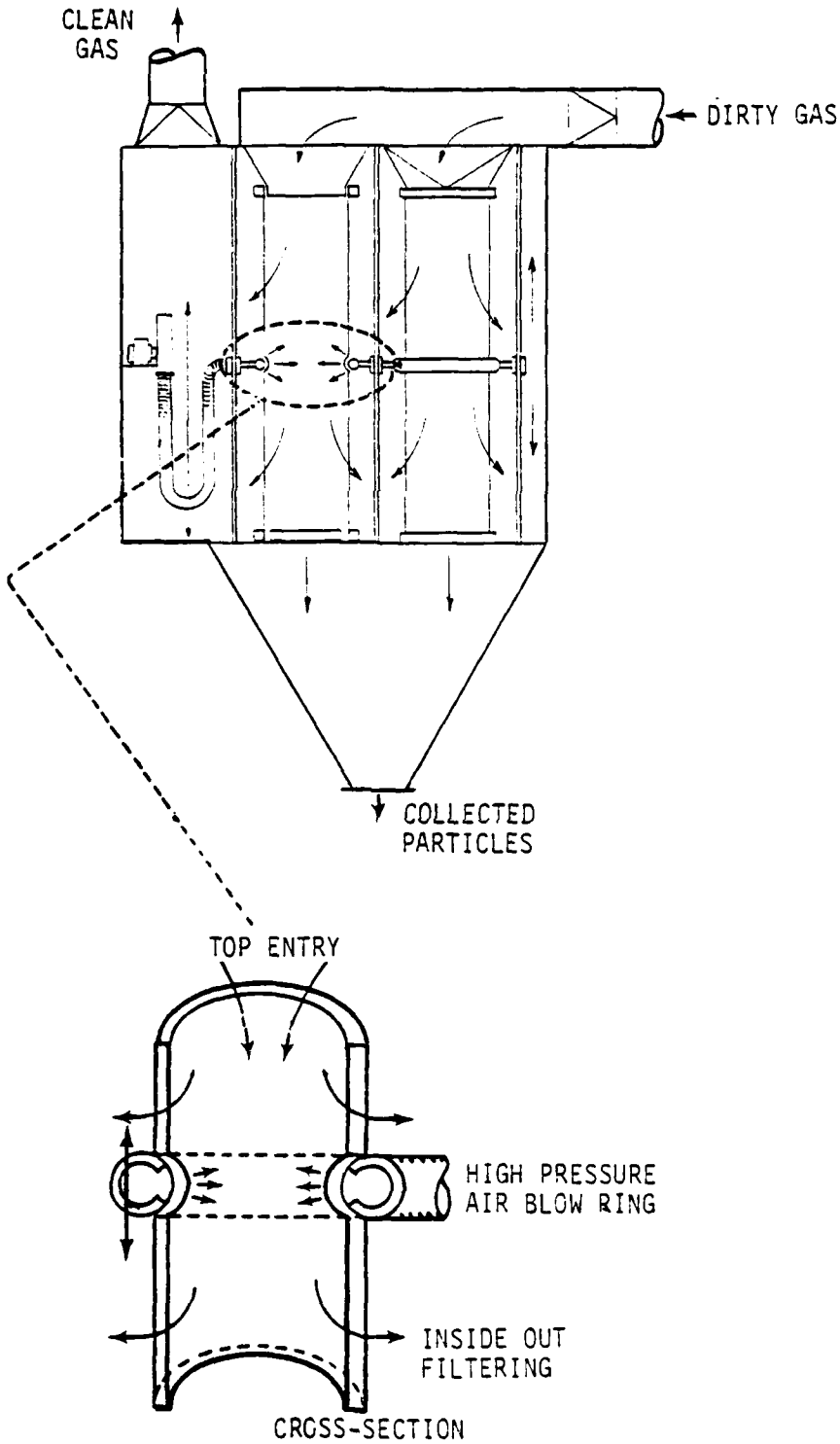

FIGURE 2-41

REVERSE-JET (TRAVELING RING TYPE) BAGHOUSE FILTER[15]

2.5.3 Design of Fabric Filters[28]

The design of industrial dust-collection equipment requires consideration of many factors. Figure 2-42 illustrates[15] the complex nature of the final selection of a fabric filter system. The most important design considerations include the operational pressure drop, cloth area, cleaning mechanism, fabric and fabric life, baghouse configuration, and costs. Exhaust volume through the usual single compartment fabric collector will not be constant because of the increasing resistance to air flow as the dust cake accumulates. The reduction in flow rate will be a function of the system pressure relationships, the exhaust fan characteristics, and the point of rating. Drop-off of exhaust volume is usually not severe in practice because pressure losses for the system of ducts and hoods usually equal or exceed those of the fabric collector; therefore, reduction in exhaust volume will cause a corresponding reduction in the pressure needs for that portion of the system. The design criteria applying to the collector itself will in turn affect those for other system components. However, during the early design phase it may not be possible to predict the interrelationships between design criteria and costs. Therefore, until realistic trade-offs can be established among collector size, fan requirements, hoods, and ducting, cost estimation within approximately $\pm$ 50 percent is acceptable.

Only those well experienced with fabric filter equipment should assume the responsibility for system design, especially of new equipment and new applications. The replacement of existing equipment is somewhat easier since one does have some practical guidelines. The following are the main design steps:

1. Define the effluent--mass flow rate, dust properties, gas properties, and process variations with time.

2. Approximate the collector design.

3. Approximate the fan and ducting needs.

4. Based on (2) and (3), select peripheral equipment, instruments, and controls.

5. Repeat (2) and (3) to minimize the estimated total cost.

6. Review for alternative effluent control methods.... budget; procure.

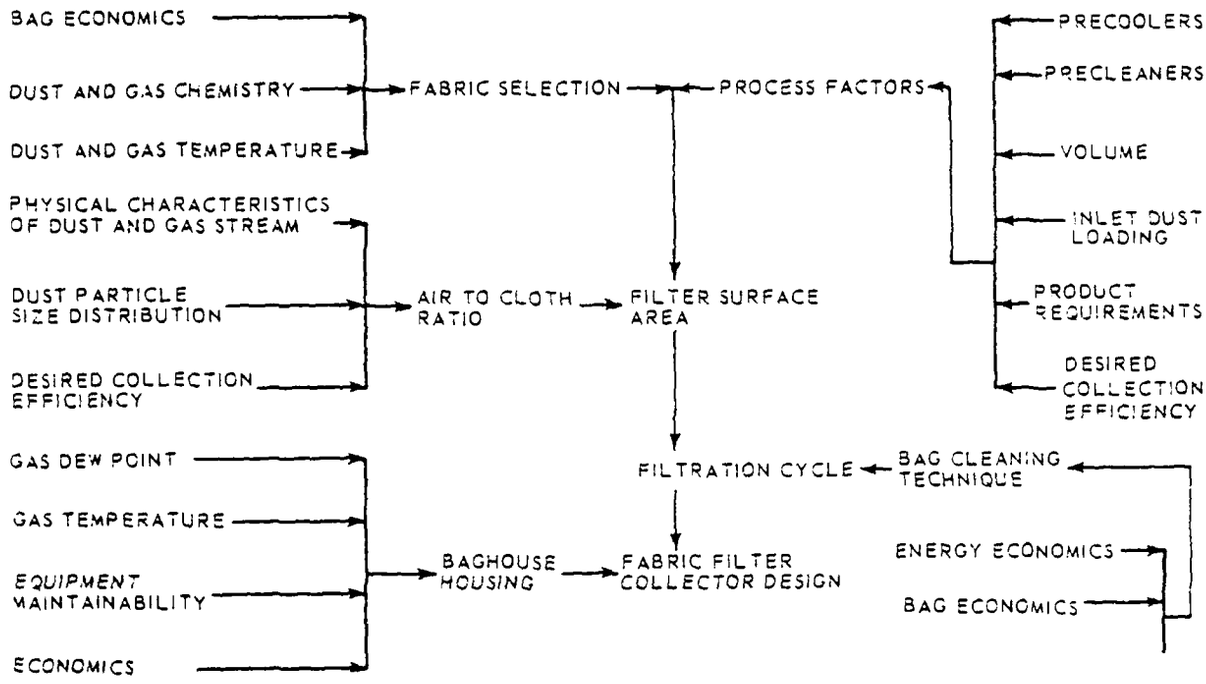

FIGURE 2-42

SYSTEM ANALYSIS FOR FABRIC FILTER COLLECTOR DESIGN [15]

The first three steps can be carried out approximately,without the cost data for individual system components, by using the guidelines set forth in the handbook.[28] In later stages, specific apparatus must be evaluated, in conjunction with equipment supplier support, to arrive at realistic estimates of initial costs.

Operating, maintenance, and overhead costs are also essential considerations in preparing design specifications. Total annual cost is probably the best criterion to use (see Chapter 4). Other criteria of performance that are difficult to assign a cost to, but which affect the overall quality of the system, include the emission level of the equipment, its reliability, simplicity of operation, etc. Although the latter factors are difficult to quantify, the design engineer must take them into account.

2.5.3.1 Description of Process Effluent to be Filtered

The definition of the problem constitutes the basis for its effective solution. In the case of fabric filtration, the crux of the problem is to define as completely as possible the process effluent properties. The desired information is listed[28] in Table 2-7.

If a process is characterized by variations in gas flow and/or particulate composition, the equipment must operate at peak loads without media plugging, as well as at reduced flows where condensation may occur. Any potential future increase in effluent loading should be considered in estimating design capacity, since an initial overdesign is often less costly than subsequent reconstruction.

TABLE 2-7

EFFLUENT AND FILTERING REQUIREMENTS[28]

1. Process Effluent

(a) Gas flow
- Average
- Maximum
- Temperature
- Water Content
- Other Constituents

(b) Dust flow
- Average
- Maximum
- Size distribution
- Size: < 1 < 5 <20 <50 μm
- % ___ ___ ___ ___
- Size: <80 <95 <99 μm
- % ___ ___ ___

2. System

(a) Preferred location (in) (out)side

(b) Space limitations, if any

(c) Ambient weather
- Range
- Temperature
- Snow, water, wind loads

(d) Weight requirements

(e) Cost considerations

TABLE 2-7 (Cont.)

Particle density
Bulk density
Est. range of K_2
Other properties _____

3. Exhaust

(a) Particulate level
(b) Gaseous requirements
(c) Visibility requirements
(d) Preferred exhaust location
(in) (out)side
(e) Distance from collector

2.5.3.2 Gas Flow

Determine the volume of gas emitted by the process prior to any corrective adjustments of temperature or dew point. Should there be temporal variations, the maximum, minimum, and normal operating flow should be estimated. The cost of the fabric filter system will be approximately proportional to the volume of gas emitted by the process. Existing leaks in the carrier gas system that will be eliminated by installation of the baghouse should be considered in the design.

2.5.3.3 Gas Properties

Determine the temperature and pressure of the carrier gas stream and estimate its approximate water content. Identify any abnormal gaseous constituents such as acid vapors, toxic and/or corrosive fumes, combustible or explosive materials, and condensibles. Determine whether composition and/or concentration vary significantly with time, particularly during process start-up or shutdown operations.

2.5.3.4 Dust Flow

Determine the weight (mass) rate of dust or fume generation by the process, again making certain that the quantity has been minimized as much as possible by process adjustment. Variable load conditions, particularly peak values, must be considered in determining filter capacity if overloading or plugging is to be avoided.

2.5.3.5 Dust Properties

A knowledge of effective densities for discrete and bulk particles as well as an estimated permeability (K_2) for the dust is also useful in establishing filter conditions.

The value of K_2, which is thought to reflect the integrated effects of particle dimensions (e.g., length to diameter ratio, cohesiveness between particles, dust cake rigidity, and spatial arrangement of particles in the dust cake), is a valuable design tool. On the other hand, individual measurement of the factors that presumably determine the K_2 are not, as yet, sufficiently understood to make their quantification possible. Therefore, the use of coefficients such as K_2 in calculations should be left to manufacturers and other experts in the design field. Some additional dust properties for which no strict quantitative definition is currently made, but which constitute important inputs for system design, are the abrasiveness, agglomerating characteristics, "seeping" tendency, adhesion of the dust cake to the fabric, and the softness or stickiness of the dust as a function of temperature or humidity.

2.5.3.6 Variability in Aerosol Composition

Allowance should be made for the fact that even without intentional modification of the gas temperature or the particle size properties, there may be radiation cooling, moisture leakage into the gas stream, agglomeration of the smaller particles and/or sedimentation of the larger particles, or other changes during transit through the system. Therefore, one must attempt to define the aerosol as it enters the filter unit.

2.5.3.7 Emission Requirements

The degree of particulate control that must be attained with the overall filter system should be determined early in the design process. This will usually be stated as a maximum tolerable weight emission rate rather than as a system efficiency. The requirements may also specify other factors that must be considered, e.g., toxic gases, odors, or visibility of particulate or steam plumes.

It must also be decided whether the filtered effluent can be discharged directly to the outside environment (with the attendant problems of heat loss, makeup ventilation,and visible exhaust) or released within the building. In the latter case, the problems of heat, materials toxicity, nuisance and/or hazard in the event of filter rupture, and noise take on added importance because of confinement.

2.5.3.8 Pressure Drop

Estimate the average pressure differential across the filter media and deposited dust layer during normal operating conditions. Although the value selected may be

somewhat arbitrary, several practical considerations, such as collector strength under pressure or vacuum, fan power requirements, and dust cake mechanics, point to a few inches of water as the optimum pressure drop. Lacking any better design guides, 3 inches of water is acceptable as a typical value. On the other hand, the use of high velocity filtration, felted fabrics, or the presence of a sticky or low porosity dust cake often require that optimum pressure drop be of the order of 8 to 10 inches of water. Pressure loss through the collector alone, exclusive of the media drop, is usually small compared to that of the loaded fabric. The pressure loss associated with the duct, hood and stack system will probably be in the same range as that for the collector with fabric.

The pressure drop through the combined fabric and dust layer can be treated as an independent variable in the sense that the design engineer can exercise considerable control over the cleaning mechanism. By increasing the intensity and/or frequency of cleaning, it is possible in some cases to reduce pressure drop to levels approaching those of the clean fabrics. If this concept is carried too far, however, the collection efficiency may be lowered, the fabric itself damaged, and the power costs for driving the cleaning mechanism increased to prohibitive levels. Thus, the selection of the optimum operating pressure loss becomes a matter of trade-offs based upon engineering judgment and field trials. Since the final operating pressure loss may not necessarily conform to the original design point, it is not practical to over-refine the preliminary estimates of average and peak pressure drops.

2.5.3.9 Air-to-cloth Ratio

This ratio (cubic feet per minute of air filtered per square feet of cloth filter area) is important in determining collector performance. the ratio (or its equivalent, filtering velocity) is closely related to dust deposit characteristics, collector configurations, collector efficiency, pressure drop, and maintenance requirements. Air-to-cloth ratios in current use range from less than 1:1 to more than 20:1. The choice depends on cleaning method and fabric and on characteristics of the particles.

There is no precisely determinable ratio for a given application as the choice also depends on estimates and trade-offs, such as between initial collector cost and recurring power costs. Consequently, there is no precise analytical method for determining the best air-to-cloth ratio. Instead, it is customary to select ratios based on similar previous experience, that is, ratios that have

been proven on similar combinations of cleaning method, fabric, and dust.

Tables 2-8 and 2-9 contain information on filtering velocities for various types of dusts in conventional shaker baghouses with woven fabrics and in reverse-jet baghouses, respectively.[29] Table 2-10 contains similar information for pulse-jet baghouses, and Table 2-11 shows recommended filtering velocities for various dusts in glass-cloth collectors.[29]

Each dust collector manufacturer has guidelines for the selection of air-to-cloth ratio, based on his experience with a variety of applications. These guidelines vary from manufacturer to manufacturer, largely as a result of differences in equipment. Four such guidelines are summarized in the handbook[28] for a shaking-bag collector, a glass cloth collector employing principally reverse flow plus flexural cleaning, a reverse-jet collector, and a reverse-pulse collector. These are typical of the guidelines that manufacturers have made publicly available. Normally, these guidelines should enable estimates to within at least 25 percent of the optimum design ratio. In unusual cases, and for a more exact estimate, consultation with an experienced manufacturer is advisable. Frequently, in new applications, a pilot study has been used to determine the best air-cloth ratio. Such studies can be misleading, however, unless they accurately model the proposed equipment and use a suitable aerosol.

There is an optimum air-to-cloth ratio for each set of filter system design parameters (i.e., dust to be filtered, configuration of filter system, cleaning mechanism to be employed, fiber material to be utilized, and configuration of the fiber media). However, for a given set of the above parameters, the total system cost versus air-to-cloth ratio relationship is rather flat near the optimum ratio. Thus, there is a tendency to minimize initial costs by selecting an air-to-cloth ratio toward the high end of the range. On the other hand, it is frequently reported that, with lengthy operation at two different filtration velocities, the lower of the two filtration velocities results in lower operational costs. Thus, one must carefully weigh the traditionally cited advantages of lowered air-to-cloth ratio (i.e., lower power costs, decreased maintenance and higher collection efficiency) against the larger initial capital costs associated with increased collector size and the penalty for its space occupancy. Furthermore, it is anticipated that with advancing fiber, fabric media, and cleaning technology, the optimum range of air-to-cloth ratios for any set of design and operating parameters will tend to increase.

TABLE 2-8

RECOMMENDED MAXIMUM FILTERING VELOCITIES FOR VARIOUS DUSTS AND FUMES IN CONVENTIONAL SHAKER BAGHOUSES WITH WOVEN FABRICS[29]

Dusts or Fumes	Maximum Filtering Ratios (cfm/ft^2 cloth area)	Dusts or Fumes	Maximum Filtering Ratios (cfm/ft^2 cloth area)
Abrasives	3.0	Coke	2.25
Alumina	2.25	Conveying	2.5
Aluminum oxide	2.0	Cork	3.0
Asbestos	2.75	Cosmetics	2.0
Baking powder	2.25-2.50	Cotton	3.5
Batch spouts for grains	3.0	Feeds and grain	3.25
Bauxite	2.5	Feldspar	2.5
Bronze powder	2.0	Fertilizer (bagging)	2.4
Brunswick clay	2.25	Fertilizer (cooler, dryer)	2.0
Buffing wheel operations	3.0-3.25	Flint	2.5
Carbon	2.0	Flour	2.5
Cement crushing and grinding	1.5	Glass	2.5
Cement kiln (wet process)	1.5	Granite	2.5
Ceramics	2.5	Graphite	2.0
Charcoal	2.25	Grinding and separating	2.25
Chocolate	2.25	Gypsum	2.5
Chrome ore	2.5	Iron Ore	2.0
Clay	2.25		
Cleanser	2.25		
Coca	2.25		

TABLE 2-8 (cont.)

Dusts or Fumes	Maximum Filtering Ratios (cfm/ft^2 cloth area)
Lampblack	2.0
Lead oxide	2.25
Leather	3.5
Lime	2.0
Limestone	2.75
Manganese	2.25
Marble	3.0
Mica	2.25
Oyster shell	3.0
Packing machines	2.75
Paint pigments	2.0
Paper	3.5
Plastics	2.5
Quartz	2.75
Rock	3.25
Sanding machines	3.25
Silica	2.75
Soap	2.25
Soapstone	2.25
Starch	2.25
Sugar	2.25
Talc	2.25
Tobacco	3.5
Wood	3.5

TABLE 2-9

RECOMMENDED MAXIMUM FILTERING VELOCITIES AND FABRIC FOR DUST AND FUME COLLECTION IN REVERSE-JET BAGHOUSES[29]

Material or Operation	Fabric	Filtering Ratios (cfm/ft^2)
Aluminum oxide	Napped cotton	11
Bauxite	Cotton sateen	10
Carbon, calcined	Napped cotton, wool felt	8[a]
Carbon, green	Orlon felt	7
Carbon, banbury mixer	Wool felt	8
Cement, raw	Cotton sateen	9
Cement, finished	Cotton sateen	10
Cement, milling	Cotton sateen	8
Chrome, (ferro)crushing	Cotton sateen	10
Clay, green	Cotton sateen	10
Clay, vitrified silicious	Cotton sateen	12
Enamel (porcelain)	Napped cotton	12
Flour	Cotton sateen	14[a]
Grain	Wood felt, cotton sateen	16
Graphite	Wool felt	7[a]
Gypsum	Cotton sateen, orlon felt	10
Lead oxide fume	Orlon felt, wool felt	8[a]
Lime	Napped cotton	10

TABLE 2-9 (Cont.)

Material or Operation	Fabric	Filtering Ratios (cfm/ft^2)
Limestone (crushing)	Cotton sateen	10
Metallurgical fumes	Orlon felt, wool felt	10[a]
Mica	Napped cotton	11
Paint pigments	Cotton sateen	10
Phenolic molding powers	Cotton sateen	10
Polyvinyl chloride (PVC)	Wood felt	10[a]
Refractory brick sizing (after firing)	Napped cotton	12
Sandblasting	Napped cotton, wood felt	6-8[a]
Silicon carbide	Cotton sateen	9-11
Soap and detergent powder	Dacron felt, orlon felt	12[a]
Soy bean	Cotton sateen	14
Starch	Cotton sateen	10
Sugar	Cotton sateen, wool felt	10[a]
Talc	Cotton sateen	11
Tantalum fluoride	Orlon felt	6[a]
Tobacco	Cotton sateen	12
Wood flour	Cotton sateen	10
Wood sawing operations	Cotton sateen	12
Zinc, metallic	Orlon felt, dacron felt	11
Zinc oxide	Orlon felt	8[a]
Zirconium oxide	Orlon felt	8

[a]Decrease 1 cfm/ft^2 if dust concentration is high or particle size is small.

TABLE 2-10

TYPICAL FILTERING VELOCITIES FOR VARIOUS DUSTS AND FUMES IN PULSE-JET BAGHOUSES[29]

Dusts or Fumes	Filtering Ratio ($Acfm/ft^2$)	Dusts or Fumes	Filtering Ratio ($Acfm/ft^2$)
Alumina	8-10	Leather dust	12-15
Asbestos	10-12	Lime	10-12
Bauxite	3-10	Limestone	8-10
Carbon black	5-6	Mica	9-11
Coal	5-10	Paint pigments	7-8
Cocoa, chocolate	12-15	Paper	10-12
Clay	9-10	Plastics	7-9
Cement	8-10	Quartz	9-11
Cosmetics	10-12	Rock dust	0-10
Enamel frit	9-10	Sand	10-12
Feeds, grain	14-15	Sawdust (wood)	12-15
Feldspar	9-10	Silica	7-9
Fertilizer	8-9	Slate	12-14
Flour	12-15	Soap, detergents	5-6
Graphite	5-6	Spices	10-12
Gypsum	10-12	Starch	8-9
Iron ore	11-12	Sugar	7-10
Iron oxide	7-8	Talc	10-12
Iron sulfate	6-8	Tobacco	13-15
Lead oxide	6-8	Zinc oxide	5-6

TABLE 2-11

RECOMMENDED FILTERING VELOCITIES FOR GLASS-CLOTH COLLECTORS[29]

Material	Filtering Ratio ($Acfm/ft^2$)
Carbon black generator furnace and channel black	1.1-1.3
Electric furnace and ferro alloy furnaces; most metallurgical fume	1.5-1.8
Cement and lime kilns, wet and dry process; open-hearth and oxygen-lanced open-hearth furnaces and smelters	1.3-2.0
Clinker coolers, refractory kilns and furnaces; coal-fired boilers (power plant)	2.0-2.3

2.5.3.10 Cleaning Mechanism and Fabric Selection

The selection of the cleaning mechanism and the filter fabric are best made together, since both items are closely related. For example, felted fabrics are almost exclusively cleaned by pulse- or reverse-jet air, whereas most woven fabrics are cleaned by other means. Of the relatively few choices of fiber media, most will be eliminated for reasons such as poor temperature and/or corrosion resistance or excessive cost. Of the several cleaning mechanisms used in filtration systems, only two or three will meet the specific requirements for a given installation, e.g., high, low or moderate dust loadings, continuous or intermittent operation, ease of removal of dust from the fabric; small floor area, minimal pressure drop, high efficiency.

By a process of elimination, therefore, a review of past successful filtering performance will usually show that only a few cleaning mechanism-fabric combinations are compatible and sufficiently attractive to warrant economic evaluation. The time required for cleaning also determines the choice of cleaning mechanism. This time should be a small fraction of the time required for dust deposition, since otherwise too large a fraction of the fabric will be out of service for cleaning at any given time. It is common with shake-cleaning equipment, for example, to have a cleaning-to-deposition time ratio of the order of 0.1 or less. Applying this criterion, having a ten compartment baghouse would mean that one compartment is out of service at all times. Therefore, the choice of cleaning mechanisms affects system size as well as fabric life, maintenance, etc.

Staple (Spun) vs. Filament Fibers

Natural fibers (cotton and wool) are available only in staple form. Individual fibers are limited to a few inches in length. Many man-made fibers are available in staple form, with a short given fiber length.

Staple fibers are used to produce fabric characteristics not always available with continuous (filament) fibers. These are larger yarns with greater bulk (and higher surface area), fabric of greater thickness, higher weight, higher permeability to air flow with tortuous, obscured air passages, etc. Such fabrics are useful for dusts that can be removed to the extent required, by the type of cloth cleaning method employed. Certain fumes or products undergoing a change of state while passing through the dust collector are known to condense on fiber ends and are generally more difficult to remove from staple fabrics.

Most man-made fibers are available in continuous filament and are generally used in multi-filament (bundle of fibers) form. Multi-filament fabrics show better release characteristics for certain dusts and fumes, especially with less vigorous methods of cloth cleaning. Filament fibers can produce fabric characteristics not always available with staple forms. These are smooth surface, high thread count, lightweight fibers operating with low residual dust burden (but not necessarily lowest pressure drop).

Where applicable, the staple fiber fabric usually operates at a lower pressure drop than the filament fiber fabric, all other conditions being equal.

The staple fiber fabric is available in all popular weaves. The filament fiber fabric is seldom used in the plain 1/1 (taffeta) weave.

Table 2-12 is a summary of the various types of filtration fabrics commercially available and their operating characteristics.[15]

2.5.3.11 Cloth Area

The amount of fabric in actual use at a given time is found by dividing the estimated total flow entering the collector by selected air-to-cloth ratio. The volume flow rate of the effluent entering the collector will not necessarily be the same as that discharging from the generating process, owing to temperature changes, the added volume of vaporized cooling water, and dilution air, which may be added deliberately for cooling purposes or accidentally by air leakage. The latter factor may contribute to a significant flow increase in systems operating under large negative pressures. In filter applications involving a varying flow, some judgment is required to decide whether to size the equipment for the peak flow, the average flow, or for some intermediate point. It is again necessary to seek a compromise between the increased cost of larger equipment and the increased cost, and the possible risk of fabric damage associated with short-term high pressure drops.

Except for certain systems that are operated intermittently, e.g., a few hours on-line followed by cleaning only during downtime, most filtration units will require reserve fabric capacity to allow for off-line cleaning, inspection, and maintenance. Since it is common practice to isolate temporarily defective filter units until it is convenient to replace them, additional reserve capacity may be required. The total or gross fabric area to be

TABLE 2-12

TYPICAL PROPERTIES OF COMMON FILTER MEDIA [15]

Fabric	Generic Name	Type Yarn	Maximum Temperature Range (°F) Long Periods of Time (months)	Short Periods of time (min)	Melting Temperature (°F)	Acid Resistance	Fluoride Resistance	Alkali Resistance	Flex and Abrasion Resistance	Relative Cost (approx.)	Supports Combustion
Cotton[a]	Natural fiber cellulose	Staple	180°	225°	302° decomposes	Poor	Poor	Fair-good	Fair-good	1.0	Yes
Wool[b]	Natural fiber protein	Staple	200°	250°	572° chars	Very good	Poor-fair	Poor-fair	Fair	2.75	No
Nylon[c]	Nylon polyamide	Filament spun	200°	250°	480°	Fair	Poor	Very good-excellent	Very good-excellent	2.5	Yes
Dynel®[d]	Modacrylic	Filament spun	180°	240°	325° softens	Good-very good	Poor	Good-very good	Fair-good	3.2	No
Polypropylene[e]	Polyolefin	Filament spun	200°	250°	333°	Excellent	Poor	Excellent	Very good-excellent	1.75	Yes
Orlon®[f]	Acrylic	Spun	240°	275°	482° softens	Good-excellent	Poor-fair	Fair	Fair	2.75	Yes
Dacron®[g]	Polyester	Filament spun	275°	325°	482°	Good	Poor-fair	Fair-good	Very good	2.8	Yes
Nomex®[h]	Nylon aromatic	Filament spun	425°	500°	700° decomposes	Fair	Good	Excellent	Very good-excellent	8.0	No
Teflon®[i]	Fluorocarbon	Filament spun	450°	500°	750° decomposes	Excellent	Poor-fair	Excellent	Fair	30.0	No
Fiberglass[j]	Glass	Filament spun bulked	500°	600°	1,470°	Fair-good	Poor	Fair	Poor	5.5	No
Polyethylene	Polyolefin	Filament spun	200°	-	-	Very good-excellent	Poor-fair	Very good-excellent	Good	2.0	Yes
Stainless steel[k] (type 304)	-	-	1,400°-1,500°	-	2,550°-2,650°	Excellent	-	Excellent	-	100.0	No

[a] Poor resistance to mildew and fungi; excellent selection in ventilation-type collector.
[b] Similar to those of cotton; good filterability.
[c] High tensile strength, good elasticity; unaffected by mildew and fungi; rugged fiber with excellent resistance to abrasion and alkalies; fair to poor resistance for most sodium salts.
[d] Good chemical and abrasion resistance and excellent dimensional stability; attacked by concentrated nitric acid, sodium hydroxide, and affected by most halogens; adversely affected by ketones, amines, cyclohexanone, and acetone.
[e] Chemically resistant to acids and alkalies; strong fiber, low moisture absorption; attacked at elevated temperatures by nitric acid and chlorosulfuric acid; has poor resistance to sodium and potassium hydroxide at high temperatures; not to be used with aromatics and chlorinated hydrocarbons.
[f] Resistance to moisture; not harmed by common solvents; not recommended for sulfuric acid (generally fair in environments above acid dewpoint); attacked by zinc chloride; good at elevated temperatures in acid conditions.
[g] High tensile strength, good dimensional stability, and excellent temperature resistance.
[h] Outstanding temperature resistance, high tensile strength, good resistance to abrasion.
[i] Can be used at elevated temperatures and possesses excellent chemical resistance.
[j] Low mechanical strength, hence vulnerable to abrasion; can be used at high temperatures and has high tensile strength.
[k] For extremely high temperatures.

installed can be estimated by multiplying the area to be in actual use at any time (net cloth area) by the term

$$(1 + T_R + F)$$

where T_R is the time ratio (cleaning to deposition) discussed in the preceding section and F is the fraction of the fabric area expected to be out of service at any time due to replacement, inspection, or maintenance operations. Judicious timing of the above procedures in relation to peak flow periods may, however, reduce the multiplying factor.

2.5.3.12 Cloth Life

Fabric deterioration often results from the combined assault of several factors, rather than from any single effect, such as thermal erosion, mechanical stress through repeated flexure, chemical attack, or abrasion. All possible modes of failure should be considered during the preliminary design phase.

Again, previous experience, especially that relating to similar fabric-cleaning-dust applications, may be the best and only guide. Extrapolating from experience, one might estimate that the reduction in fiber life through thermal erosion might double for a 20°F rise in temperature, or that the mechanical attrition rate might double when the frequency of cleaning is doubled.

Generally, it should be possible to estimate fabric life within a factor of two in situations where no direct experience can be cited. If performance data are available, estimated reliabilities may be upgraded to perhaps $\pm$ 20 percent, which is the order of dependability of the best pilot plant data. Having established a reasonable estimate of fabric life, one can then reach an annual cost figure for fabric media.

2.5.3.13 Number of Compartments

The basic information required in selecting the number of separate compartments is the allowable variation in gas flow with respect to process or plant ventilation, the availability of sizes of commercial units (compartments or filter house modules), and the expected frequency of maintenance. In small collectors, individual compartments may contain as little as 100 square feet of fabric surface, although collectors as large as approximately 50,000 cfm capacity may also have only one compartment. Multiple compartments of almost any size may be chosen, subject to availability. With the exception of reverse-jet and

pulse-jet units, at least one compartment will be out of service during the cleaning cycle. It may also be necessary to provide additional compartments for emergency, extended maintenance, or unexpected increases in process effluent.

2.5.3.14 Compartment Structure

Although many kinds of fabric collectors can, in principle, be operated either under negative or positive pressure, the larger units are often custom-designed for one condition or the other to minimize costs. Least expensive is the installation needing no housing at all, in which the particulate is collected on the interior surfaces of a positive pressure bag system. The danger of cooling below the dew point is an important consideration, particularly in a suction housing where infiltration leakage often occurs unless a more expensive gas-tight design is adopted. Thermal excursions of the collector during start-up and shutdown can affect the sealing characteristics of critical gasketed connections, depending on the structural materials. Interconnection of compartments via ducting or hoppers and their isolation, especially during the cleaning cycle, are important considerations. The generalized rules for designing the compartment structure are not well defined. Construction materials are likewise not generally prescribed. The most common housing material is steel. The choice of material depends on the nature of the dust and gas mixture and their flammability, corrosiveness, etc. Lacking a more specific guideline, one would do well to follow precedent in selecting materials of construction, metal gauges and dimensions.

2.5.3.15 System Pressure Drop

The total system pressure drop is that of the combined losses in the duct and the fabric filter unit. Ducting losses vary approximately as the square of the gas velocity and can be readily calculated by standard formulas or from tabular or graphical data. Pressure drops for non-linear shapes (elbows, Tee's, reducers, etc.) are usually expressed in equivalent length of straight duct of the same diameter. In conventional practice, one traces the largest branch (usually from the most remote source of dust generation), noting the temperature and gas flow through each succeeding section up to the collector inlet. A similar procedure is followed from the collector exit to the point where the filtered effluent is discharged to the inside or outside atmosphere. Ordinarily, the total pressure drop associated with the ducting alone will be in the range of 3 to 6 inches of water, although other values may apply in some circumstances.

One then adds to the estimated pressure drop through the ductwork, the collector pressure drop and that of any other component of the system, e.g., a centrifugal collector. The result will be the net pressure to be supplied by the fan when ambient pressures are the same at system inlet and outlet. Should the ambient pressure at the system inlet exceed that at the outlet, the net pressure requirement for the fan is decreased by this difference.* The opposite applies if the pressure gradient is reversed.

2.5.4 Advantages and Disadvantages of Fabric Filter System

The principal advantages of fabric filter systems include:

(1) Particle collection efficiency is very high and can be maintained at consistently high levels (usually greater than 99 percent).

(2) Efficiency and pressure drop are relatively unaffected by large changes in inlet dust loadings for continuously cleaned filters.

(3) Filter outlet air may be recirculated within the plant in many cases.

(4) The collected material is recovered dry for subsequent processing or disposal.

(5) There are no problems of liquid waste disposal, water pollution, or liquid freezing.

(6) There is no hazard of high voltage, simplifying maintenance and repair and permitting collection of flammable dusts.

(7) Use of selected fibrous or granular filter aids (precoating) permits the high-efficiency collection of submicron smokes and gaseous contaminants.

(8) Filter collectors are available in a large number of configurations, resulting in a range of dimensions and inlet and outlet flange locations to suit installation requirements.

*The kinetic energy or velocity pressure retained by the air leaving the fan may be deducted from the fan static pressure requirement, except when the air is charged to a large stagnant space, e.g., exhausting the fan outdoors. The kinetic deduction is frequently overlooked, however, on the justification that by omitting it, a safety factor is added. See the excellent discussion on fan selection in Fan Engineering by the Buffalo Forge Company.

Disadvantages in the use of fabric filters include:

(1) Fabric life may be shortened at elevated temperatures and in the presence of acid or alkaline particle or gas constituents.

(2) Temperatures much in excess of 500°F require special refractory mineral or metallic fabrics that are still in the developmental stage and can be very expensive.

(3) Hygroscopic materials, condensation of moisture, or tarry adhesive components may cause crusty caking or plugging of the fabric or require special additives.

(4) Certain dusts may require fabric treatments to reduce seeping of the dust or, in other cases, assist in the removal of the collected dust.

(5) Concentrations of some dusts in the collector (50 g/m^3) may represent a fire or explosion hazard if spark or flame is admitted by accident. Fabrics can burn if readily oxidizable dust is being collected.

(6) Replacement of fabric may require respiratory protection for maintenance personnel.

Fabric filters have been designed and installed to handle from 100 to 4.5 million acfm in a single installation. A well-designed (and properly maintained) baghouse is capable of handling the entire range of gas-borne stack emission particles, fine or coarse in size, light or heavy in grain loading. Its efficiency is approximately constant, regardless of varying moisture, temperature, particle size, or gas flow. On applications where the dust resistivity is high, the electrostatic precipitator can become large and expensive and the first cost of a fabric filter can become comparatively attractive.

2.5.5 Illustrative Examples

2.5.5.1 Example No. 1

Small-scale tests showed that filtration of an air stream containing one grain of particulates per cubic foot of air gave a maximum pressure drop of 5 inches of water at a flow rate of 3 ft^3/min per square foot of filtering surface.

a. Calculate the horsepower required for a fan for a flow rate of 6,000 ft^3/min.

b. Calculate the number of 0.5-ft diameter by 10 ft filtering bags required for the system. Assume an overall fan-motor efficiency of 63 percent.

Solution:

This example does not involve complete design of fabric filter. Only the following steps are necessary to solve it.

Step 3.

Air-to-cloth ratio = 3 cfm/ft^2 of filtering surface (given)

Step 4.

Filtering Area = 60,000 ft^3/min (given)

$$\frac{\text{Flow rate of air}}{\text{Air-to-cloth ratio}}$$

$$\frac{60{,}000\ \text{ft}^3/\text{min}}{3\ \text{cfm/ft}^2}$$

$$20{,}000\ \text{ft}^2$$

(a) For 63 percent fan-motor efficiency, the horsepower required for a fan is determined as follows:[30]

$$\text{hp} = \frac{\text{Air flow rate (cfm)} \times \Delta P\ \text{(inches of water)}}{4{,}000}$$

Where ΔP = pressure drop = 5 in. of water (given)

Then,

$$\text{hp} = \frac{(6{,}000\ \text{ft}^3/\text{min})\ (5\ \text{in. of water})}{4{,}000} = 7.5$$

(b) Filtering area per bag = $2\pi rH$

Where r = radius of bag = 0.25 ft (given)
H = height of bag = 10 ft (given)

Then, filtering area per bag

$$= 2 \times 3.14 \times 0.25\ \text{ft} \times 10\ \text{ft}$$
$$= 15.7\ \text{ft}^2$$

$$\text{Number of bags required} = \frac{\text{Total filtering area}}{\text{Filtering area per bag}}$$

$$\frac{20{,}000\ \text{ft}^2}{15.7\ \text{ft}^2}$$

$$= 127.3, \text{ say } 128$$

If shake or reverse air cleaning methods are used, an additional unit should be included in the design because one unit is always off-line during cleaning. If a pulse-jet cleaning method is used, no extra units are needed since the bags are cleaned while on-line.

2.5.5.2 Example No. 2

A plywood mill plans to install a fabric filter as an air cleaning device. How many bags, each 8 inches in diameter and 12 ft long, must be used to treat the exhaust gas, which has a particulate loading of 2 grains/ft^3 if the exhaust fan is rated at 7,000 ft^3/min.

Solution:

This example, too, does not involve the complete design and only those steps necessary to solve it will be used.

Step 4.

Consider the approximate superficial velocity equal to 2 ft/min.

$$\text{Total filtering area} = \frac{\text{Flow rate of air}}{\text{Filtering velocity}}$$

$$\frac{7{,}000\ ft^3/min}{2\ ft/min}$$

$$= 3{,}500\ ft^2$$

Area per bag = $2\pi rH$

Here, r = radius of bag = 4 in. = $\frac{4}{12}$ ft

H = height of bag = 12 ft

Then, area/bag = $2 \times 3.14\ (\frac{4}{12}\ ft)\ (12\ ft)$

$= 25.1\ ft^2$

$$\text{Number of bags required} = \frac{\text{Total filtering area}}{\text{Area per bag}}$$

$$\frac{3{,}500\ ft^2}{25.1\ ft^2}$$

$$= 139.4, \text{ say } 140$$

Assemble four compartments, each containing 6 x 6 = 36 bags. Therefore, the total number of bags is 144, the

total area is (144 x 25.1) = 3,618 ft^2, and the actual superficial velocity is less than 2 ft/min.

$$\text{Actual superficial velocity} = \frac{7{,}000\ ft^3/min}{3{,}618\ ft^2} = 1.93\ ft/min$$

If shake or reverse air cleaning methods are used, an additional unit should be included in the design because one unit is always off-line during cleaning. If a pulse-jet cleaning method is used, no extra units are needed since the bags are cleaned while on-line.

2.5.5.3 Example No. 3

A plant emits 50,000 acfm of gas at a dust loading of 5 grains/ft^3. The dust is collected by a fabric filter at 98 percent efficiency when the average filtration velocity is 10 ft/min. The pressure drop is given by

$$\Delta P = 0.2v + 5c_i v^2 t$$

where

ΔP = the pressure drop in inches of water
v = the filtration velocity in ft/min
c_i = the dust concentration in lb/ft^3 of gas
t = the time in minutes since bags were cleaned.

a. How many cylindrical bags, 1 ft in diameter and 15 ft high, will be needed?

b. If the fan can maintain the volume rate of flow up to a pressure drop of 8 inches of water, how frequently must the bags be cleaned?

Solution:

Step 4.

Average filtering velocity = 10 ft/min (given)
Gas flow rate = 50,000 acfm (given)

$$\text{Filtering area required} = \frac{\text{Gas flow rate}}{\text{Average filtering velocity}}$$

$$\frac{50{,}000\ acfm}{10\ ft/min}$$

$$= 5{,}000\ ft^2$$

(a) Area per bag = $2\pi rH + \pi r^2$

r = radius of bag = 0.5 ft
H = height of bag = 15 ft

$$\text{Area/bag} = 2 \times 3.14 \times (0.5\ \text{ft})\ (15\ \text{ft}) + 3.14\ (0.5\ \text{ft})^2$$
$$= 47.9\ \text{ft}^2$$

$$\text{No. of bags required} = \frac{\text{Total filtering area}}{\text{Area per bag}}$$

$$\frac{5{,}000\ \text{ft}^2}{47.9\ \text{ft}^2}$$

$$= 104.4, \text{ say } 104$$

Arrange into four units, each containing 26 bags.

(b) The pressure drop is given by

$$\Delta P = 0.2v + 5c_i v^2 t \qquad \text{(given)}$$

or

$$t = \frac{\Delta P - 0.2v}{5c_i v^2}$$

The following information is given:

P = pressure drop = 8 in. of water.
c_i = dust concentration = 5 grains/ft^3
$= \frac{5}{7{,}000}$ lb/ft^3
v = filtration velocity = 10 ft/min

Therefore,

$$t = \frac{(8\ \text{in. of water}) - 0.2\ (10\ \text{ft/min})}{5\left(\frac{5}{7{,}000}\ \text{lb/ft}^3\right)(10\ \text{ft/min})^2}$$

= 16.8 minutes between cleaning.

Note: For a cleaning period of this length, pulse-jet cleaning would not be used.

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CHAPTER 3

CONTROL EQUIPMENT FOR GASEOUS POLLUTANTS

National Ambient Air Quality Standards (NAAQS) have been established for sulfur dioxide, nitrogen oxides, hydrocarbons, and carbon monoxide. Four basic methods can be used to control the emissions of these gaseous pollutants into the atmosphere:

- absorption
- adsorption
- combustion
- condensation.

Within each method, various different devices are available. Selection of an optimum device for a specific job depends upon various factors including the contaminant properties, carrier gas characteristics, process factors, and operational factors.

3.1 STEP-BY-STEP DESIGN REVIEW PROCEDURE

After selecting a control device for a given job, its detailed design can be carried out by using the step-by-step procedure given in this section. A logical sequence of steps is presented for designing each of the following control equipment: (i) packed columns, (ii) plate columns, (iii) adsorption systems, (iv) flares, (v) thermal combustion reactors, and (vi) catalytic combustion reactors. Again, each step can be implemented by using the reference given for that step.

3.1.1 Step-by-Step Design Procedure for Packed Columns

		Reference
1.	Select a solvent	App. C.1.5.4
2.	Evaluate the equilibrium data	App. C.1.3
3.	Estimate the operating data	App. C.1.5.2 C.1.5.3
4.	Select the type of packing	App. C.1.6.1
5.	Calculate the column diameter	Sect. 3.2.4.1
6.	Calculate the optimum packing wetting rate	Sect. 3.2.4.1
7.	Calculate the height of a transfer unit	Sect. 3.2.4.1
8.	Calculate the number of transfer units required	Sect. 3.2.4.1
9.	Calculate the required column height	Sect. 3.2.4.1
10.	Calculate the pressure drop through the column	Sect. 3.2.4.1
11.	Estimate capital and operating costs of equipment	Sect. 4.5.5

3.1.2 Step-by-Step Design Procedure for Plate Columns

		Reference
1.	Select a solvent	App. C.1.5.4
2.	Evaluate the equilibrium data	App. C.1.3
3.	Estimate the operating data	App. C.1.5.2 C.1.5.3
4.	Select the type of plate	App. C.1.7
5.	Calculate the column diameter	Sect. 3.2.4.2
6.	Calculate the number of trays required	Sect. 3.2.4.2
7.	Calculate the required column height	Sect. 3.2.4.2
8.	Determine the pressure drop through the column	Sect. 3.2.4.2
9.	Estimate capital and operating costs of equipment	Sect. 4.5.5

3.1.3 Step-by-Step Design Procedure for Adsorption Systems

		Reference
1.	Select type of adsorbent	App. C.2.4
2.	Calculate the gas actual volumetric flow rate	Sect. 3.3.4
3.	Calculate the required quantity of adsorbent	Sect. 3.3.4
4.	Select required gas velocity through the bed	Sect. 3.3.4
5.	Calculate the required bed dimensions	Sect. 3.3.4
6.	Calculate the pressure drop through the bed	Sect. 3.3.4
7.	Estimate the capital and operating costs of equipment	Sect. 4.5.6, 4.6.2.1

3.1.4 Step-by-Step Design Review Procedure for Flares

		Reference
1.	Determine Btu content of waste gas	NA
2.	Determine the waste gas flow rate	NA
3.	Calculate the required stack diameter	Sect. 3.4.3.1
4.	Calculate the required stack height	Sect. 3.4.3.1

NA = not applicable

3.1.5 Step-by-Step Design Review Procedure for Thermal Combustion Reactors

		Reference
1.	Determine the Btu content of the waste gas	NA
2.	Determine the waste gas flow rate	NA
3.	Calculate the fuel requirement to sustain combustion in the incinerator	Sect. 3.4.3.2
4.	Select retention time and temperature required	NA
5.	Calculate the afterburner's required combustion chamber volume	Sect. 3.4.3.2
6.	Estimate the capital and operating cost of equipment	Sect. 4.5.7, 4.6.2.2

NA = not applicable

3.1.6 Step-by-Step Design Procedure for Catalytic Combustion Reactors

Variations and unknowns in many of the parameters in the describing equation(s) make it nearly impossible to provide a generalized procedure for the design of a catalytic reactor.

However, design considerations must definitely include:

1. Mass of catalyst for a given conversion at a particular flow rate.

2. Energy (fuel) requirements.

3. Estimation of capital and operating costs of equipment (Sect. 4.5.8 and 4.6.2.3).

3.2 CONTROL BY ABSORPTION

3.2.1 Introduction

The removal of one or more selected components from a gas mixture by absorption is probably the most important operation in the control of gaseous pollutant emissions. The process of absorption conventionally refers to the intimate contacting of a mixture of gases with a liquid so that part of one or more of the constituents of the gas will dissolve in the liquid. The contact conventionally takes place in some type of packed column or wet scrubber. The net effect is the transfer of a soluble gas from the gas stream into the liquid stream. The purpose of such gas scrubbing operations might include product recovery or process reaction, as well as pollutant removal, or even some combination thereof. For example, the pollutant sulfur dioxide removed from a contaminated gas stream may also be recovered to provide a valuable by-product in the form of sulfuric acid. The main emphasis in this part of the manual, however, will be solely on the elimination of pollutants or other objectionable components, such as odors, from gas streams.

Emphasis has been placed on presenting that information most likely to find practical use.

3.2.2 Description of Available Equipment

3.2.2.1 Packed Columns

Packed columns find normal application in the removal of pollutant gases and vapors. However, such systems have also been considered for the removal of particulate matter. They are usually vertical columns filled with packing or devices of large surface area. The liquid is distributed over and trickles down through the packed bed, thus exposing a large surface area to contact the gas. The countercurrent packed column (Figure 3-1) is the most common type of unit encountered in gaseous pollutant control for the removal of the undesirable gas, vapor, or odor. The gas stream (containing the pollutant) moves upwards through the packed bed against an absorbing or reacting liquor (solvent-scrubbing solution), which is injected at the top of the packing. This results in the highest possible efficiency. Since the solute concentration in the gas stream decreases as it rises through the column, there is constantly fresher solvent available for contact. This provides the maximum average driving force for the diffusion process throughout the packed bed.

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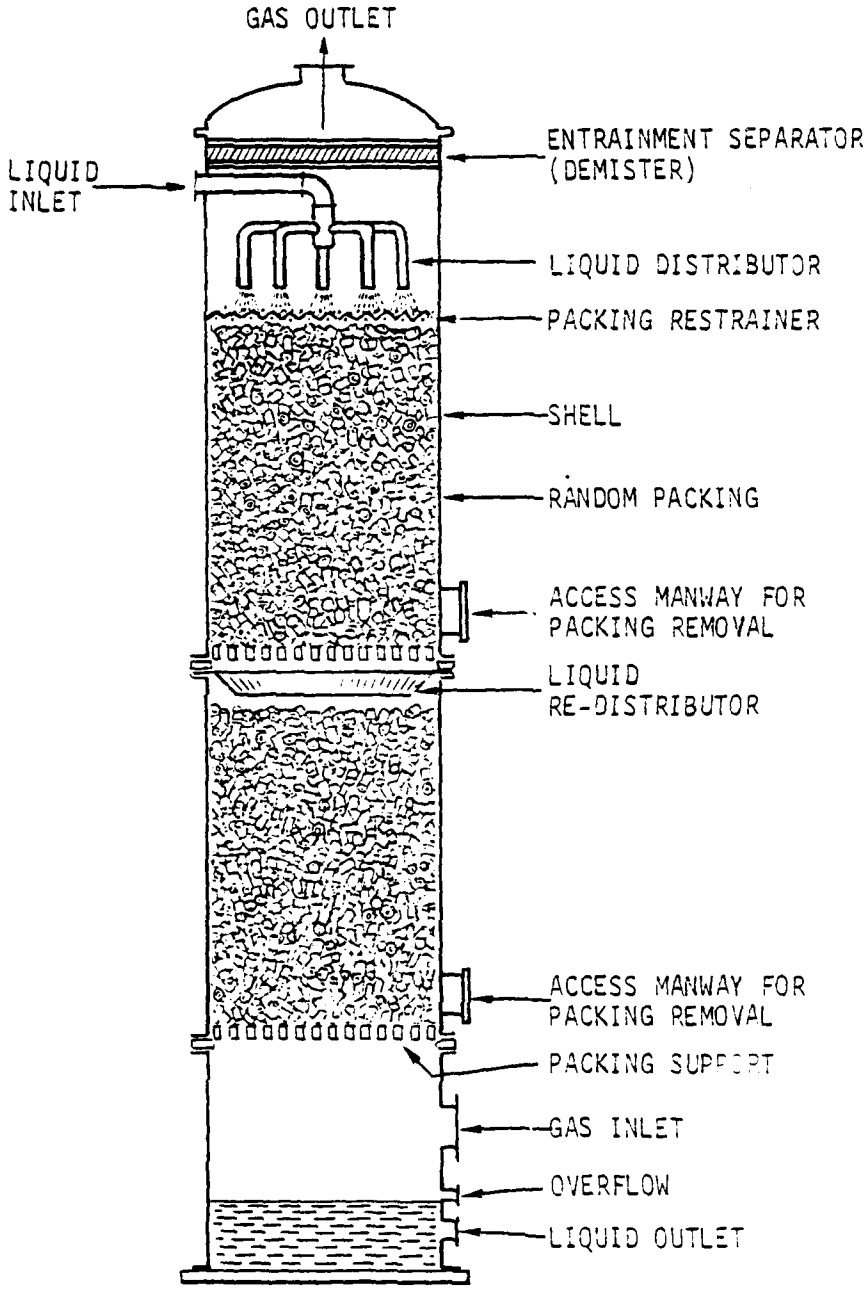


FIGURE 3-1

TYPICAL COUNTERCURRENT PACKED COLUMN [32]

There are four general ways in which the packed-bed scrubbers may be operated to remove particulates (Figure 3-2).[15] Cross-flow scrubbers will efficiently remove the relatively large-size particulates; they are also used to remove liquid particulates. The co-current flow scrubber is somewhat more efficient than the cross flow and is effective in removing both solid and liquid particulates. For the collection of particulates as small as 3 to 5 microns, packed countercurrent columns have been found to be most efficient.

3.2.2.2 Plate Columns

Plate columns are essentially vertical cylinders in which the liquid and gas are contacted in stepwise fashion (staged operation) on plates or trays, in a manner shown schematically for one type in Figure 3-3. The liquid enters at the top and flows downward via gravity. On the way, it flows across each plate and through a downspout to the plate below. The gas passes upward through openings of one sort or another in the plate, then bubbles through the liquid to form a froth, disengages from the froth, and passes on to the next plate above. The overall effect is a multiple countercurrent contact of gas and liquid. Each plate of the column is a stage, since on the plate, the fluids are brought into intimate contact, interphase diffusion occurs, and the fluids are separated.

3.2.3 Operating Characteristics

3.2.3.1 Packed Columns

1. The operating gas rates seldom exceed 60 to 75 percent of the rate that would cause flooding.

2. Column dimensions specified are usually in readily available off-the-shelf sizes (i.e., diameters to the nearest half foot and heights to the nearest foot).

3. Bed depths may vary from 6 inches to several feet depending upon the type of packing and the application.

4. Typical pressure drop through a packed column is 0.5 inch of water per foot of packing.

5. Liquid is generally redistributed every 6 to 10 feet of packed height to minimize channeling.

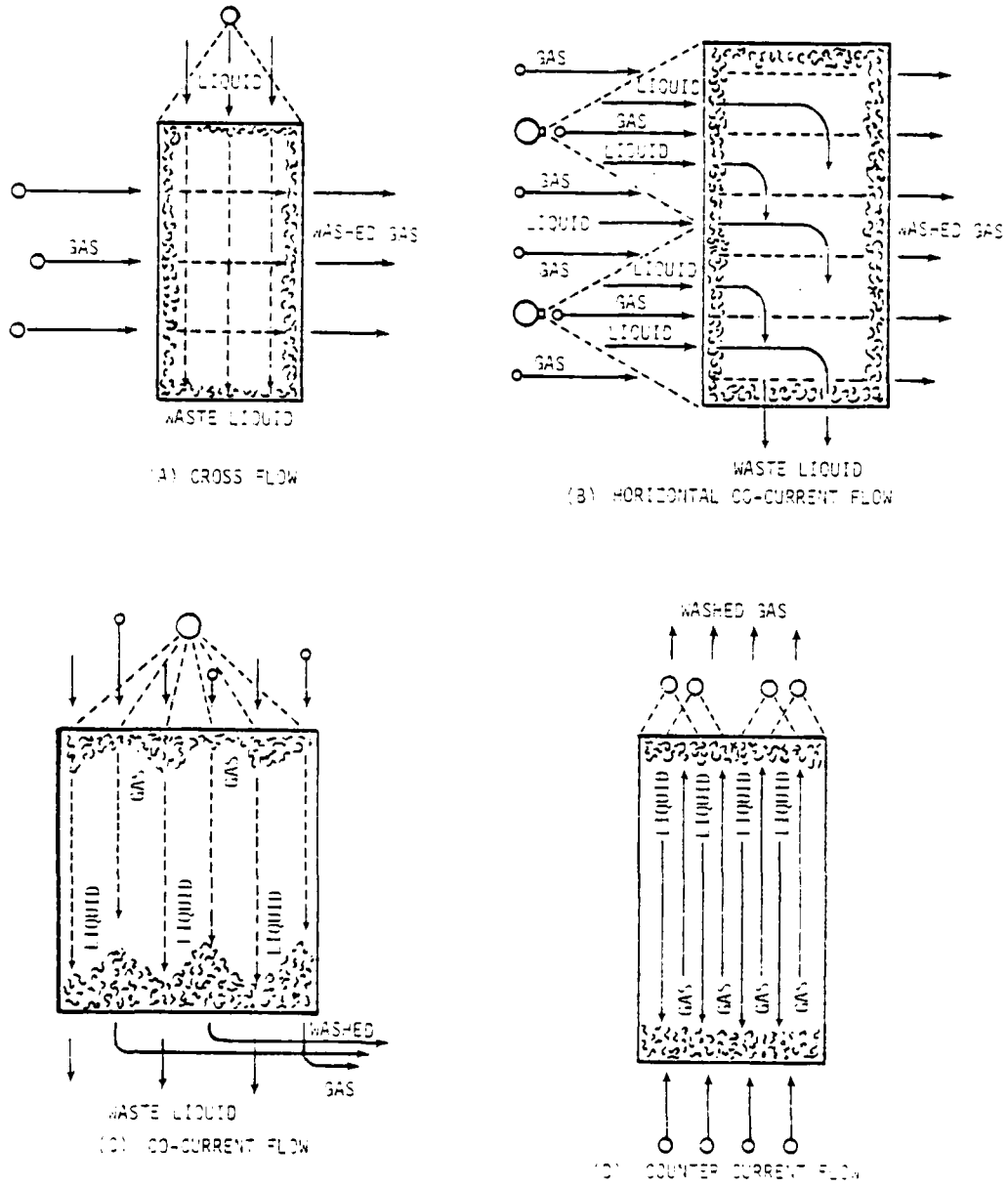


FIGURE 3-2

VARIOUS OPERATING METHODS FOR PACKED COLUMNS[15]

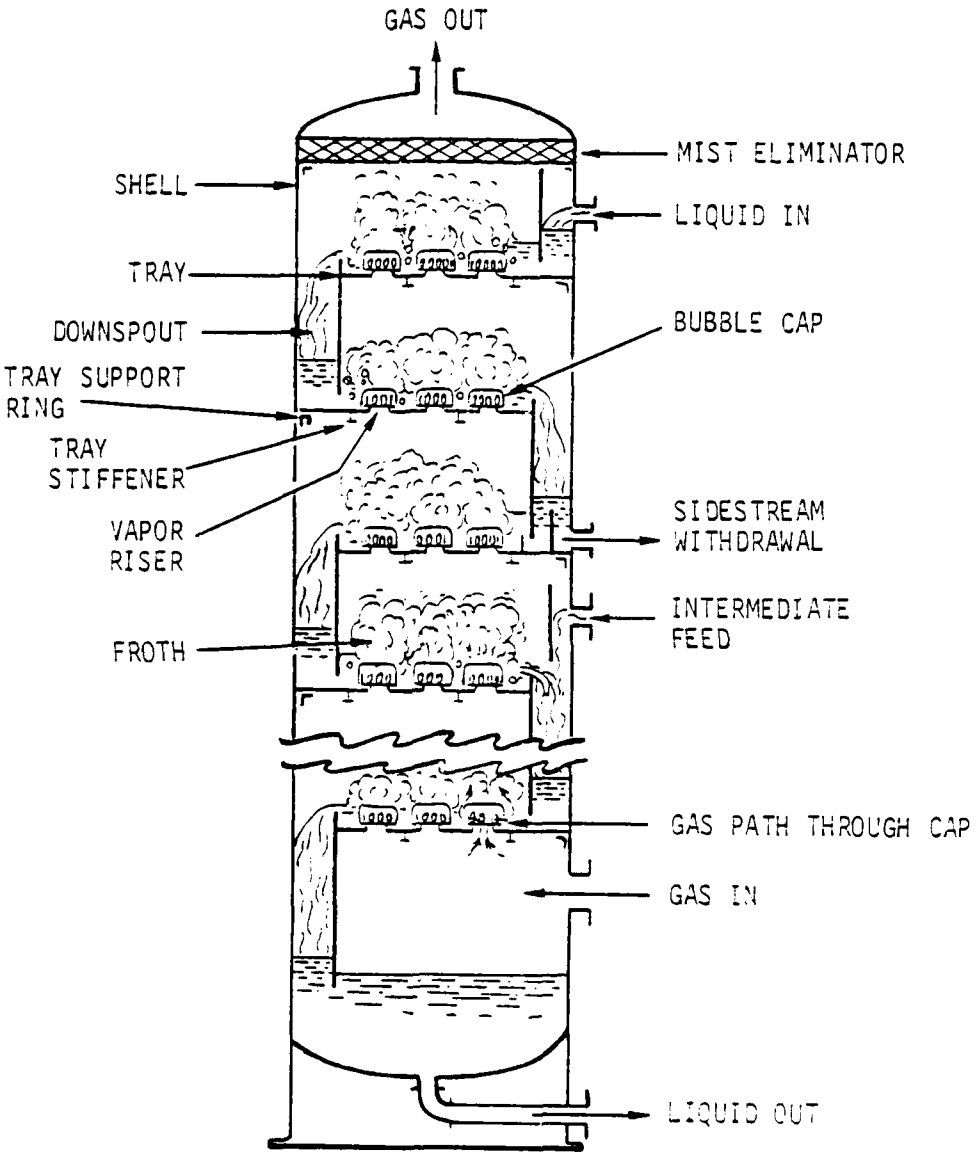


FIGURE 3-3

TYPICAL BUBBLE-CAP PLATE COLUMN [32]

3.2.3.2 Plate Columns

1. Liquid Flow

   a. Not over 0.165 ft^3 of liquid/(sec) (ft diameter) for single-pass cross flow trays.

   b. Not over 0.35 ft^3 of liquid/(sec) (ft weir length) for others.

2. Keep liquid gradient about 0.5 inch, with 1.25 inch maximum.

3.

Tower Diameter (ft)	Tray Spacing (inches)
4 or less	18-20
4-10	24
10-12	30
12-24	36

3.2.4 Design Review Procedure

3.2.4.1 Packed Columns

The streams entering and leaving the column and their constituents are first identified and their flow rates are determined by taking the component and overall mass balances.[31] The operating conditions are determined depending upon the gas streams to be treated. The physical dimensions of the column can then be calculated. The column must be of sufficient diameter to accommodate the gas and liquid and of sufficient height to insure that the required amount of mass is transferred with the existing driving force.

Procedure:

Step 1. Calculate the column diameter

a. Calculate the abscissa of Figure 3-4.

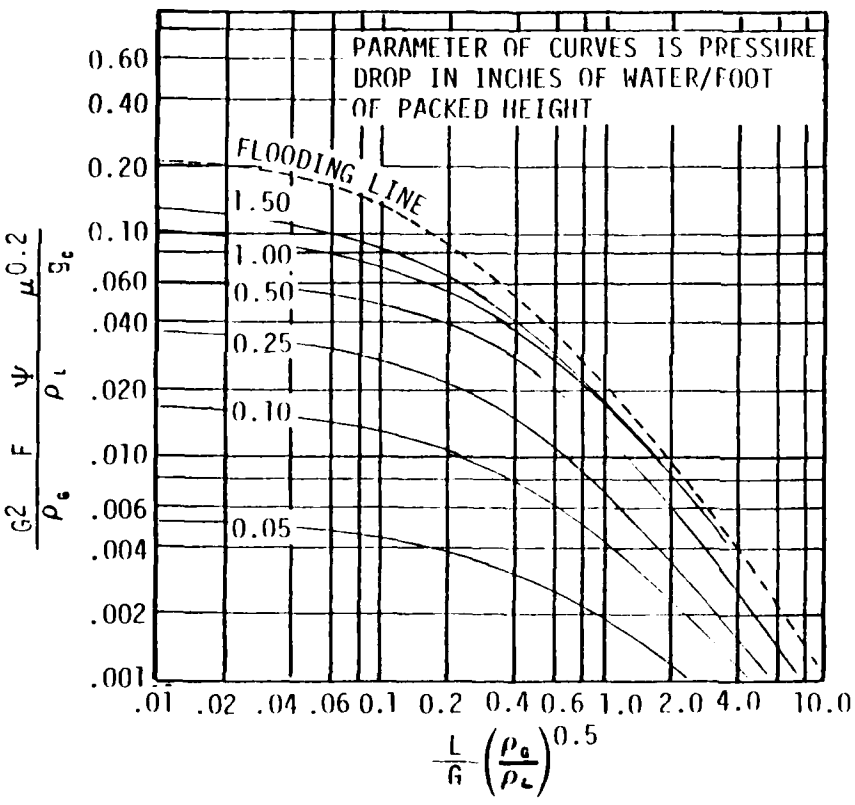
b. Proceed to flooding line and read the ordinate.

c. Solve the ordinate equation for G at flooding rate.

d. Calculate the column cross-sectional area, S, for the fraction of flooding velocity, F, chosen for operation:[32]

$$S = \frac{G'}{FG} \qquad (3.1)$$

where G' = mass flow rate of gas, lb/hr
G = superficial gas rate, lb/hr-ft^2



L = LIQUID RATE, LB/SEC SQ FT

G = GAS RATE, LB/SEC SQ FT

ρ_L = LIQUID DENSITY, LB/CU FT

ρ_G = GAS DENSITY, LB/CU FT

F = PACKING FACTOR

μ = VISCOSITY OF LIQUID, CENTIPOISE

Ψ = RATIO, $\frac{\text{DENSITY OF WATER}}{\text{DENSITY OF LIQUID}}$

g_c = GRAVITATIONAL CONSTANT = 32.2

FIGURE 3-4

GENERALIZED PRESSURE DROP CORRELATION TO ESTIMATE COLUMN DIAMETER[32]

e. Calculate column diameter, D_T, from following equation:

$$D_T = 1.13(S)^{0.5}$$

Step 2. Calculate the optimum (minimum) packing wetting rate.

In order to feed enough liquid into the column to effectively place a wet film of liquid over all packing, the packing must be evaluated for guidance in operation and design. Table 3-1 gives minimum wetting rate (MWR) for various packings.

TABLE 3-1

MINIMUM WETTING RATE

Packing	$MWR = \frac{\text{Liquid Rate, ft}^3\text{/hr-ft}^2\text{ Cross Section}}{\text{Packing Surface Area Per Column Volume, ft}^2\text{/ft}^3}$
Rings (Raschig, Lessing, etc.) through 3-in. diameter	0.85
Grid type (wooden, etc.) (pitch: 2 in.)	0.85
All packings larger than 3-in. diameter	1.30
Polished metal packings and poor wetting surfaces (some plastics, glazed porcelain, etc.)	1.30-2.50

The minimum mass flow rate of liquid per unit cross-sectional area of the column is:[32]

$$L_{min} = (MWR)\ (\rho_L)\ (a) \qquad (3.2)$$

where L = superficial liquid rate, lb/hr-ft^2
ρ_L = liquid density, lb/ft^3
a = surface area of packing per unit volume of bed, ft^2/ft^3.

The liquid rate at the loading point can be estimated from the graphic correlation presented as a plot of the volumetric gas-liquid ratio against the wetting rate.

This is accomplished as follows:

a. Calculate:

$$\frac{G''}{L''} = \frac{G'}{L'} \times \frac{\rho_L}{\rho_G}$$

where

G'' = volumetric gas flow rate, ft^3/hr
L'' = volumetric liquid flow rate, ft^3/hr
G' = mass flow rate of the gas, lb/hr
L' = mass flow rate of liquid, lb/hr
ρ_L = liquid density, lb/ft^3
ρ_G = gas density, lb/ft^3

b. Calculate the correction factor, F_a

where $$F_a = \left[\frac{\rho_G}{\rho_{air\ (at\ 20°C)}} \right]^{(0.5)} \qquad (3.3)$$

c. Using Figure 3-5, calculate the ordinate and proceed to the curve of the specific packing. Read off the abscissa the wetting rate, W_{max}, required for loading at this particular gas-liquid ratio.

This will correspond to a liquid load of

$$L_{max} = W_{max} \cdot \rho_L a \qquad (3.4)$$

The liquid load in the column should, therefore, lie between the two limits calculated.

$$L_{min} < L < L_{max}$$

This should insure that the packing is satisfactorily wetted with no danger of cascading, which could occur if the loading point were exceeded.

Step 3. Calculate the height of a transfer unit.

a. First, determine the height of a gas transfer unit by the following equation.[32]

$$H_G = \frac{\alpha G^{\beta}}{L^{\delta}} \left(\frac{\mu}{\rho_G \quad D_G} \right)^{0.5} \qquad (3.5)$$

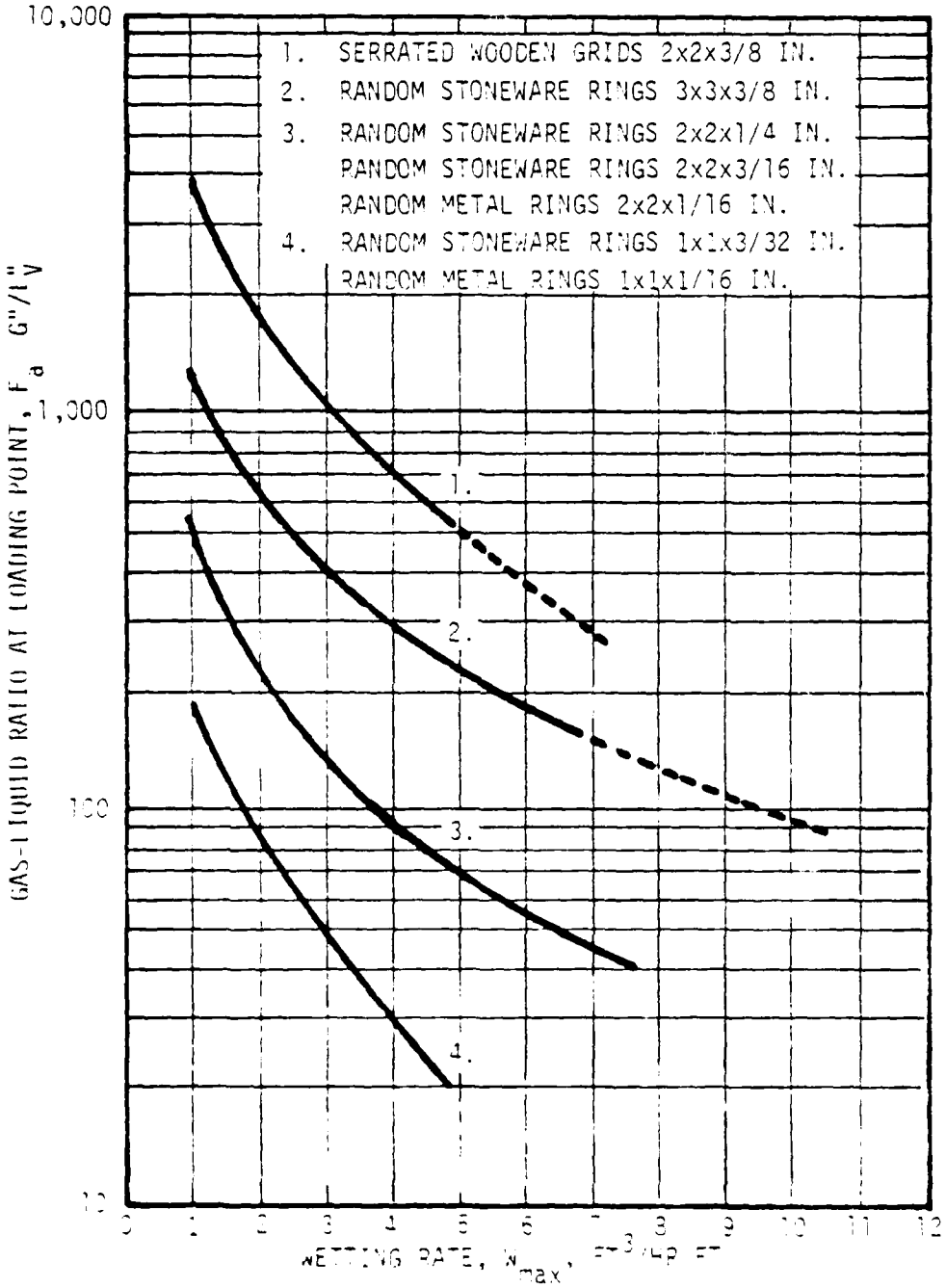


FIGURE 3-5

LOADING RATES FOR VARIOUS PACKINGS[32]

where

H_G = height of a gas transfer unit, ft
G = superficial gas rate, lb/hr-ft^2
L = superficial liquid rate, lb/hr-ft^2
α = a packing constant from Appendix Table C-4
β = a packing constant from Appendix Table C-4
δ = a packing constant from Appendix Table C-4
μ = gas viscosity, lb/hr-ft
ρ_G = gas density, lb/ft^3
D_G = gas diffusivity, ft^2/hr.

The group $(\mu/\rho\ D_G)$ is known as the Schmidt number as shown in Table C-6.

b. Determine the height of a liquid transfer unit by the following equation:[32]

$$H_L = \phi \left(\frac{L}{\mu_L}\right)^{\eta} \left(\frac{\mu_L}{\rho_L D_L}\right)^{0.5} \qquad (3.6)$$

where

H_L = height of a liquid transfer unit, ft
L = superficial liquid rate, lb/hr-ft^2
μ_L = liquid viscosity, lb/hr-ft
ϕ = a packing constant, Appendix Table C-5
η = a packing constant, Appendix Table C-5
ρ_L = liquid density, lb/ft^3
D_L = liquid diffusivity, ft^2/hr

c. Calculate the height of a transfer unit based on overall gas film coefficients:[32]

$$H_{OG} = H_G + (m)\ \frac{G'_m}{L'_m}\ (H_L) \qquad (3.7)$$

where

m = slope of the equilibrium curve
G'_m = gas rate, lb-moles/hr
L'_m = liquid rate, lb-moles/hr

Step 4. Calculate the number of transfer units required.

To calculate the number of transfer units required, based on overall gas film co-efficients[32] (for dilute solutions where operating and equilibrium lines are straight)

$$N_{og} = \frac{\ln\left[\frac{y_1 - mx_2}{y_2 - mx_1}\left(1 + \frac{mG_m}{L_m}\right) + \frac{mG_m}{L_m}\right]}{\left(1 - \frac{mG_m}{L_m}\right)} \qquad (3.8)$$

where

Y_1 = mole fraction of solute in gas stream entering column
Y_2 = mole fraction of solute in gas stream leaving column
m = slope of equilibrium line
X_2 = mole fraction of solute in liquid stream entering column
X_1 = mole fraction of solute in liquid stream leaving column
L_m = total liquid rate lb moles/hr-ft^2
G_m = total gas rate, lb moles/hr-ft^2.

Step 5. Calculate the required column height.

Column height (Z) = $H_{og} \times N_{og}$

Step 6. Calculate the column pressure drop.

a. Using Figure 3-4 with the actual mass flow percent of flooding and with the previously calculated ordinate and abcissa values, read the pressure drop, p, in inches water gauge/ft of packed height.

b. Total pressure drop = p x z.

3.2.4.2 Plate Columns

The most important design considerations for plate columns include calculation of the column diameter, type and number of trays to be used, actual tray layout and physical design, and tray spacing (which determines the

column height). To consider each of these to any great extent is beyond the scope of this manual. Details are available in any standard chemical engineering unit operations (i.e., distillation) or mass transfer text.[31] The discussion that follows, therefore, will be a relatively brief presentation of some of the general design techniques. Section 3.2.3.2 lists some recommended conditions and dimensions for tray columns that can be used as a general guide.

Step 1. Calculate column diameter.

a. Determine the superficial gas velocity for a given type of plate at flooding.[32]

$$V_F = C_F \left(\frac{\rho_L - \rho_G}{\rho_G}\right)^{0.5} \tag{3.9}$$

where

V_F = gas velocity through the net column cross-sectional area for gas flow, ft/sec

$\rho_L \rho_G$ = liquid and gas densities, respectively, lb/ft^3

C_F = empirical constant (see Appendix C-1.7.4 for individual plate types).

b. Determine the volumetric flow rate at the bottom of tower.

c. Determine tower cross-sectional area by dividing number from step (b) by velocity determined in step (a).

d. Tower diameter, D_T (ft), is computed by

$$D_T = 1.13\ (S)^{0.5}$$

where S = tower cross-sectional area, ft^2.

Step 2. Calculate column height.

This is determined from the product of the number of actual plates (theoretical plates dividing by overall plate efficiency) and the plate spacing chosen.

a. Determine the number of theoretical plates (where both operating line and equilibrium line are straight).[31]

$$N = \frac{\log\left[\frac{Y_{N_p+1} - mX_o}{Y_1 - mX_o}\left(1 - \frac{mG'_m}{L'_m}\right) + \frac{mG'_m}{L'_m}\right]}{\log\left(\frac{L'_m}{mG'_m}\right)} \qquad (3.10)$$

where

mX_o = gas composition in equilibrium with entering liquid
$Y_{Np}+1$ = solute concentration in gas stream entering column
Y_1 = solute concentration in gas stream leaving column
X_o = solute concentration in liquid stream entering column
m = slope of equilibrium curve
L'_m = total liquid rate, lb moles/hr
G'_m = total gas rate, lb moles/hr.

Step 3. Pressure Drop

As a rule of thumb, the following pressure drop values are used:

Pressure Column Operating at	Pressure Drop per Tray
35 mm Hg abs	0.011 inches water
1 atmosphere	1.94 inches water
300 psi	4.16 inches water

3.2.5 Advantages and Disadvantages of Absorption Systems

3.2.5.1 Advantages

1. Relatively low pressure drop.

2. Standardization in fiberglass reinforced plastic (FRP) construction permits operation in highly corrosive atmospheres.

3. Capable of achieving relatively high mass transfer efficiencies.

4. Increasing the height and/or type of packing or number of plates can improve mass transfer without purchasing a new piece of equipment.

5. Relatively low capital cost.

6. Relatively small space requirements.

7. Ability to collect particulates as well as gases.

3.2.5.2 Disadvantages

1. May create water disposal problem.

2. Product collected wet.

3. Particulates deposition may cause plugging of the bed or plates.

4. When FRP construction used, sensitive to temperature.

5. Relatively high maintenance costs.

3.2.5.3 Comparison of Packed and Plate Columns

Packed Column	Plate Column
• Lower pressure drop • Simpler and cheaper to construct • Preferable for liquids with high foaming tendencies	• Less susceptible to plugging • Less weight for (for structural support) • Less of a problem with channeling • Temperature surge will result in less damage

3.2.6 Illustrative Examples

3.2.6.1 Example 1 - Packed Columns

Design a packed tower to remove 95 percent of the ammonia from a gaseous mixture of 10 percent by volume of ammonia and 90 percent by volume of air. The gas mixture consists of 80 lb-moles/hr at 68°F and 1 atm. Water containing no ammonia is to be used as solvent and the packing will be 1-inch Raschig rings. The tower will be designed to operate at 60 percent of the flood point, and

isothermal conditions at 68°F will be assumed. The water will not be recirculated. Determine water flow rate, tower diameter, packed height, and tower pressure drop.

Solution:

Step 1.

Water will be used as a solvent as stated in the example.

Step 2.

a. Equilibrium data for the system ammonia-water are as follows:[30]

X	0.0206	0.0310	0.0407	0.0502	0.735	0.0962
Y	0.0158	0.0240	0.0329	0.0418	0.0660	0.0920

Plot the equilibrium curve as shown in Figure 3-6.

The curve is straight approximately to the point P, with a slope of about 0.75. Above point P, the slope is variable and higher than 0.75. Use 0.75 as the slope, m, of the equilibrium curve.

b. When the temperature rise of the solvent is negligible, apply the relation:

$$\frac{mG'_m}{L'_m} = 0.70 \qquad (3.11)$$

where

m = slope of equilibrium curve = 0.75
G'_m = gas rate = 80 lb-moles/hr
L'_m = liquid rate, lb-moles/hr

$$L'_m = \frac{(80 \text{ lb-moles/hr.})\ (0.75)}{0.70} = 85.8 \text{ lb-moles/hr.}$$

Step 3.

From the given gas flow rate, the calculated liquid rate, and the degree of absorption desired (95 percent of ammonia), tabulate gas and liquid flow rates at both ends of the tower:

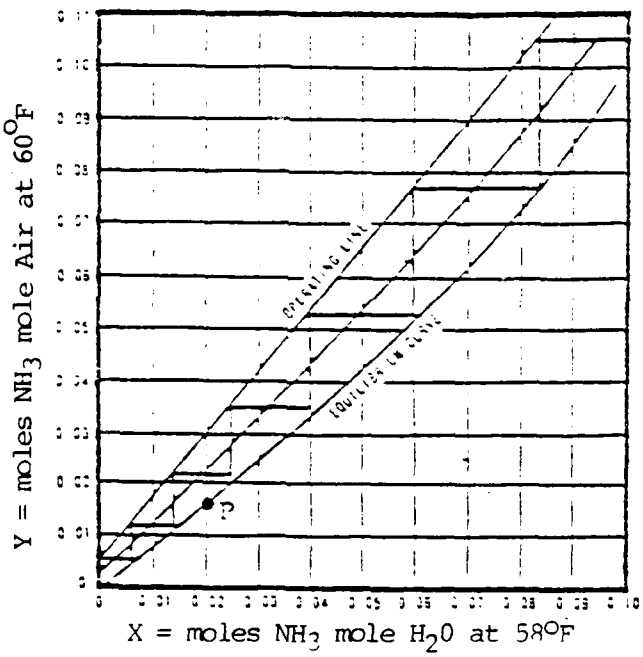


FIGURE 3-6

EQUILIBRIUM CURVE FOR AMMONIA-WATER SYSTEM.

	(lb-moles/hr)	(lb/hr)	Density (lb/ft^3)
Inlet gas (bottom)	80	2,221	0.0720
Outlet gas (top)	72.4	2,092	0.0750
Inlet water (top)	85.8	1,542	62.4
Outlet liquor (bottom)	93.4	1,671	62.4

Operating temperature = 68°F (given)
Operating pressure = 1 atmosphere (given)

Step 4.

One-inch Raschig rings will be used as the packing material as given in the problem.

Step 5.

The tower diameter is calculated as follows:

a. Use conditions at top of tower:

(i) $$\frac{L}{G}\left(\frac{\rho_G}{\rho_L}\right)^{0.5} = \frac{1{,}542 \text{ lb/hr}}{2{,}092 \text{ lb/hr}}\left(\frac{0.075 \text{ lb/ft}^3}{62.4 \text{ lb/ft}^3}\right)^{0.5}$$

$$= 0.0256$$

(ii) From Figure 3-4, the relationship

$$\frac{(G)^2(F)\ (\psi)\ (\mu_L)^{0.2}}{(\rho_G)\ (\rho_L)(g_c)} = 0.19 \text{ (for flooding line)}$$

(iii) Packing factor F = 160 (from Appendix Table C-3).
Since water is used as solvent, ψ = 1.

$$G^2 = \frac{0.19\ (\rho_G)(\rho_L)(g_c)}{(F)\ (\psi)(\mu_L)\ 0.2}$$

$$= \frac{0.19\ (0.075 \text{ lb/ft}^3)(62.4 \text{ lb/ft}^3)(32.2)}{(160)(1)(1)\ 0.2}$$

$$G = 0.423 \text{ lb/sec-ft}^2$$
$$= 1{,}522.8 \text{ lb/hr-ft}^2$$

(iv) Tower cross-sectional area,

$$S = \frac{G'}{fG}$$

G' = mass flow rate of gas = 2,092 lb/hr

At 60 percent of flooding, f = 0.6

$$\text{Then, } S = \frac{2{,}092 \text{ lb/hr}}{(0.6)(1522.8 \text{ lb/hr-ft}^2)}$$

$$= 2.29 \text{ ft}^2$$

b. Use conditions at bottom of tower:

(i) $$\frac{L}{G}\left(\frac{\rho_G}{\rho_L}\right)^{0.5} = \frac{1{,}671 \text{ lb/hr}}{2{,}221 \text{ lb/hr}}\left(\frac{0.072 \text{ lb/ft}^3}{62.4 \text{ lb/ft}^3}\right)^{0.5}$$

$$= 0.0256$$

(ii) From Figure 3-4, the relationship

$$\frac{(G)^2(F)(\psi)(\mu_L)^{0.2}}{(\rho_G)(\rho_L)(g_c)} = 0.19 \text{ (for flooding line)}$$

(iii) Packing factor F = 160 (from Appendix Table C-3)
Since water is used as solvent, $\psi = 1$.

$$G^2 = \frac{0.19\ (\rho_G)(\rho_L)(g_c)}{(F)(\psi)(\mu_L)^{0.2}}$$

$$= \frac{0.19\ (0.072 \text{ lb/ft}^3)(62.4 \text{ lb/ft}^3)(32.2)}{(160)(1)(1)^{0.2}}$$

$$G = 0.414 \text{ lb/sec-ft}^2$$

$$= 1{,}490 \text{ lb/hr-ft}^2$$

(iv) Tower cross-sectional area,

$$S = \frac{G'}{fG}$$

G' = mass flow rate of gas = 2,221 lb/hr

At 60% of flooding, f = 0.6

then, $$S = \frac{2{,}221 \text{ lb/hr}}{(0.6)(1490.4 \text{ lb/hr-ft}^2)}$$

$$= 2.48 \text{ ft}^2$$

c. Select the tower with uniform cross section of 2.48 ft^2

d. Tower diameter

$$D_T = 1.13(S)^{0.5}$$

$$= 1.13\ (2.48 \text{ ft}^2)^{0.5}$$

$$= 1.78 \text{ ft or } 21.4 \text{ inches}$$

Step 6.

Since liquid flow rate has already been determined, it is not necessary to find the optimum packing wetting rate.

Step 7.

a. Height of a gas transfer unit,

$$H_G = \frac{\alpha G^{\beta}}{L^{\delta}} \left(\frac{\mu}{\rho_G D_G}\right)^{0.5}$$

where $$G = \frac{2{,}221 \text{ lb/hr}}{2.48 \text{ ft}^2} = 896 \text{ lb/hr-ft}^2$$

$$L = \frac{1{,}542 \text{ lb/hr}}{2.48 \text{ ft}^2} = 622 \text{ lb/hr-ft}^2$$

α = 7.00 from Appendix Table C-4

β = 0.39 from Appendix Table C-4

δ = 0.58 from Appendix Table C-4

$\frac{\mu}{\rho_G D_G}$ = 0.66 from Appendix Table C-6

Then,

$$H_G = \frac{(7.00)(896 \text{ lb/hr-ft}^2)}{(622 \text{ lb/hr-ft}^2)^{0.58}}^{0.39} (0.66)^{0.5}$$

$$= 1.93 \text{ ft}$$

b. Height of a liquid transfer unit,

$$H_L = \phi \left(\frac{L}{\mu_L}\right)^{\eta} \left(\frac{\mu_L}{\rho_L D_L}\right)^{0.5}$$

where ϕ = 0.01 from Appendix Table C-5

η = 0.22 from Appendix Table C-5

L = 622 lb/hr-ft^2

μ_L = 2.42 lb/ft-hr

$\frac{\mu_L}{\rho_L D_L}$ = 570 from Appendix Table C-7

then,

$$H_L = 0.01 \left(\frac{622 \text{ lb/hr-ft}^2}{2.42 \text{ lb/ft-hr}}\right)^{0.22} (570)^{0.5}$$

$$= 0.81 \text{ ft}$$

c. Overall gas transfer unit,

$$H_{OG} = H_G + m \frac{G'_m}{L'_m} (H_L)$$

where m = slope of equilibrium curve = 0.75

G'_m = 80 lb-moles/hr

L'_m = 85.8 lb-moles/hr

then,

$$H_{OG} = 1.93 \text{ ft} + 0.75 \frac{80 \text{ lb moles/hr}}{85.8 \text{ lb moles/hr}} (0.81 \text{ ft})$$

$$= 2.50 \text{ ft}$$

Step 8.

The number of transfer units is graphically determined as follows:

a. For the gas stream,

mole fraction of NH_3 at bottom of tower, $Y_1 = 0.111$

mole fraction of NH_3 at top of tower, $Y_2 = 0.0056$

For the liquid stream,

mole fraction of NH_3 at bottom of tower, $X_1 = 0.088$

mole fraction of NH_3 at top of tower, $X_2 = 0.0$

b. Plot the operating line from data in (a) on the same graph used for the equilibrium curve.

c. The number of transfer units (NTU) is graphically determined as

NTU = 6

Step 9.

Required column height,

$$z = NTU \times H_{OG}$$

$$= 6 \times 2.50 \text{ ft} = 15 \text{ ft}$$

Step 10.

For actual mass flow rates at 60 percent of flooding:

Ordinate of Figure 3-4 = 0.069

Abscicca of Figure 3-4 = 0.0255

Intersection of these two correspond to

0.74 in. of water/ft of packed height

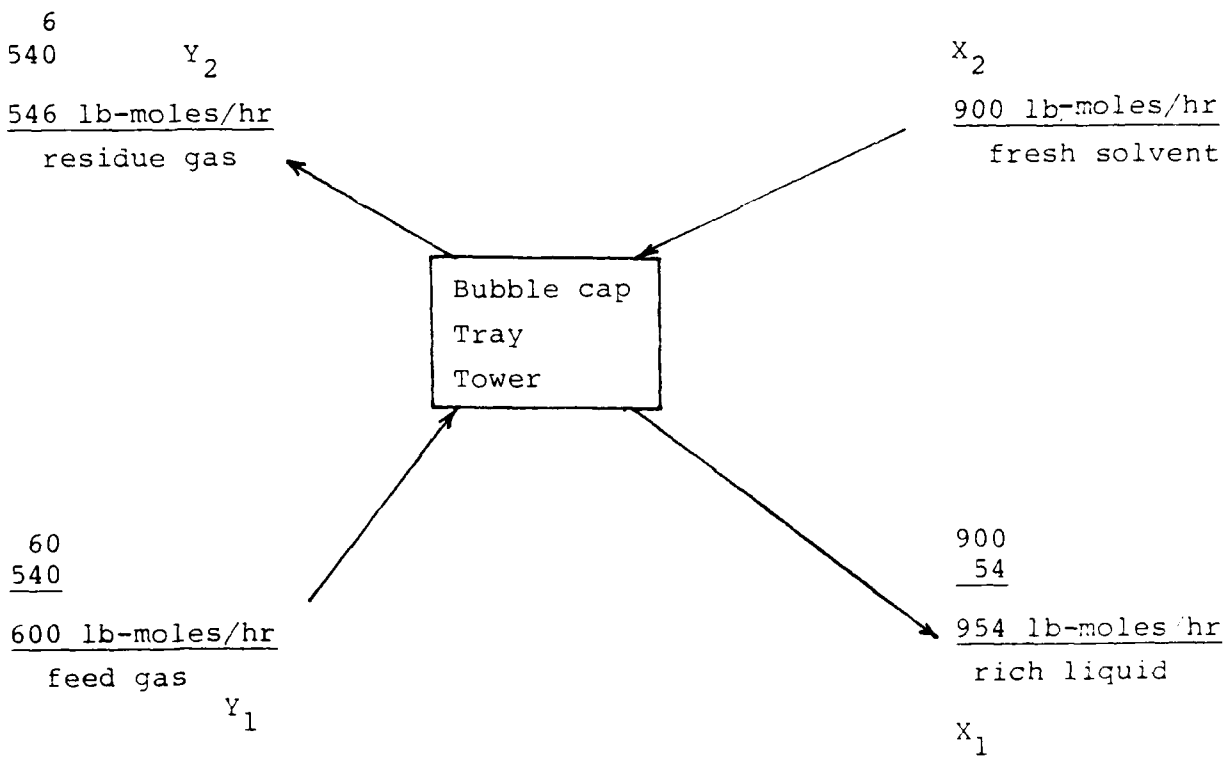
Pressure drop = 0.74 in. of water/ft of packed height x 15 ft

= 11.1 in. of water

3.2.6.2 Example 2-Plate Columns

Determine the number of actual plates and the diameter of a bubble cap plate tower for removing 90% of the ammonia from a gas stream containing 600 lb-moles/hr of gas at 68°F and 1 atm composed of 10% by volume of ammonia and 90% by volume of air.

Solvent rate expressed as moles solute/mole solvent is obtained from an operating line displaced substantially from the equilibrium curve as shown in the illustration that follows:



	Flow (lb moles/hr)	Flow (lb/hr)	Density (lb/ft^3)
Feed gas	600	16,680	0.0722
Residue gas	546	15,762	0.0750
Absorbent liquid	900	16,200	62.3
Rich liquid	954	17,118	62.3

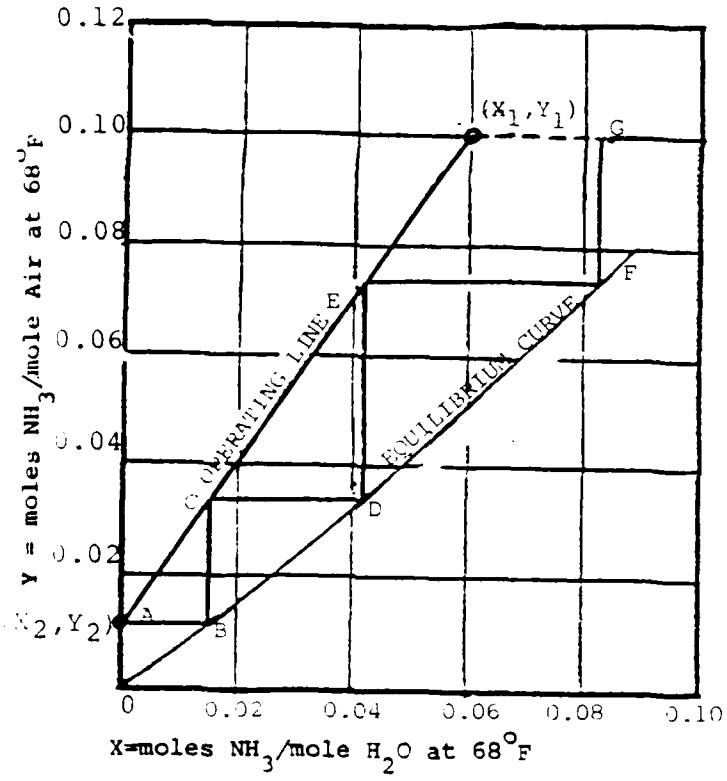


FIGURE 3-7

PLOT OF OPERATING LINE FROM THE CONDITIONS AT TOP AND BOTTOM OF BUBBLE CAP PLATE TOWER

Solution:

Step 1.

Water is selected as a solvent.

Step 2.

Equilibrium curve is shown in Figure 3-7.

Step 3.

(i) Calculate the mole ratios of solute in gas and liquid streams at both ends of the tower:

(a) Mole ratios at bottom of tower:

$$Y_1 = \frac{60}{540} = 0.111 \text{ mole } NH_3/\text{mole air}$$

$$X_1 = \frac{54}{900} = 0.06 \text{ mole } NH_3/\text{mole } H_2O$$

(b) Mole ratios at top of tower:

$$Y_2 = \frac{6}{540} = 0.011 \text{ mole } NH_3/\text{mole air}$$

$$X_2 = 0.0 \quad \text{mole } NH_3/\text{mole } H_2O$$

(ii) The operating line is plotted as shown in Figure 3-7 from the conditions at top and bottom of the column as determined in (i). A straight line is drawn between points (x_1, Y_1) and (X_2, Y_2).

Step 4.

Bubble cap plate is used as stated in the example.

Step 5.

(i) Superficial gas velocity at flooding,

$$V_F = C_F \left(\frac{\rho_L - \rho_G}{\rho_G}\right)^{0.5}$$

where $C_F = 0.17$ from Appendix Figure C-15

then,

$$V_F = 0.17 \left(\frac{62.3 \text{ lb/ft}^3 - 0.072 \text{ lb/ft}^3}{0.072 \text{ lb/ft}^3}\right)^{0.5}$$

$$= 5.00 \text{ ft/sec}$$

(ii) Volumetric flow rate at the bottom of tower, G'', is calculated as follows:

One lb-mole of an ideal gas occupies 359 ft^3 at 32^oF and 1 atm. Using the ideal-gas law equation, G" at 68^oF (528^oR) and 1 atm is given by,

$$G'' = (600 \text{ lb-moles/hr})(359 \text{ ft}^3/\text{lb-mole}) \times \frac{528^oR}{492^oR}$$

$$= 231{,}161 \text{ ft}^3/\text{hr}$$

$$= 64.21 \text{ ft}^3/\text{sec}$$

(iii) Cross-sectional area of tower,

$$S = \frac{G''}{V_F} = \frac{64.21 \text{ ft}^3/\text{sec}}{5.00 \text{ ft/sec}} = 12.84 \text{ ft}^2$$

(iv) Tower diameter,

$$D_T = 1.13 \ (S)^{0.5}$$

$$= 1.13 \ (12.84 \text{ ft}^2)^{0.5}$$

$$= 4.05 \text{ ft}$$

Step 6.

(i) Number of theoretical plates required:

A horizontal line AB is drawn from the operating line at the conditions at the top of the column to the equilibrium curve on Figure 3-7. Line BC is then drawn vertically from the equilibrium line back to the operating line. The step ABC is a theoretical plate. The stepwise procedure is repeated to the end of the operating line. The solution shows 2.45 theoretical plates.

(ii) Number of actual plates required:

With a viscosity, μ_L, of 1 centipoise for water and a slope of the equilibrium curve, m, of 0.83 (this assumes the equilibrium

curve to be straight over the area covered by the operating line), the value $m\mu_L$ is (1)(0.83) = 0.83 from Appendix Figure C-14; the overall plate efficiency is 72 percent. Actual plates required:

$$\frac{2.45}{0.72} = 3.4.$$ Use 4 bubble cap trays.

For towers operating at atmospheric pressure, maximum pressure drop = 3.33 in. of water/tray,

Hence, pressure drop = 3.33 in. of water/tray x 4 trays
= 13.32 in. of water

3.3 CONTROL BY ADSORPTION

3.3.1 Introduction

Gas adsorption is another useful method finding increasing importance in air pollution control. The advent of more stringent environmental control regulations has especially enhanced the attractiveness of adsorption as a method of control. More and more industries are recognizing the advantages of applying the principles of adsorption to remove toxic or obnoxious gases or vapors prior to discharging them to the atmosphere. When the removal of vapors, acid gases, or noxious odors is especially difficult using regular scrubbing techniques (absorption), adsorption may prove to be a more satisfactory solution.

It has been well established that the molecular forces at the surface of a liquid are in a state of unbalance or unsaturation. The same is true of the surface of a solid, where the molecules or ions in the surface may not have all their forces satisfied by union with other particles. As a result of this unsaturation, solid and liquid surfaces tend to satisfy their residual forces by attracting onto and retaining on their surfaces gases or dissolved substances with which they come in contact. This phenomenon of concentration of a substance on the surface of a solid (or liquid) is called adsorption. The substance thus attracted to a surface is said to be the adsorbed phase or adsorbate, while the substance to which it is attached is the adsorbent.

3.3.2 Description of Available Equipment

Two cylindrical vessels filled with an adsorbent in a vertical arrangement are customary for use in a continuous adsorption operation. While one bed is adsorbing vapors from the gas stream, the second bed is being stripped of adsorbate (desorbed) and regenerated. This process is shown in Figure 3-8.

3.3.3 Operating Characteristics

Activated Carbon

1. In general, removal of gaseous vapors by physical adsorption is practical for vapors with molecular weights greater than 45 and boiling points greater than 32°F (an exception to the molecular weight rule is methanol).

2. For average bed conditions, 100 pounds of carbon can efficiently treat 200 acfm of solvent laden air per hour.

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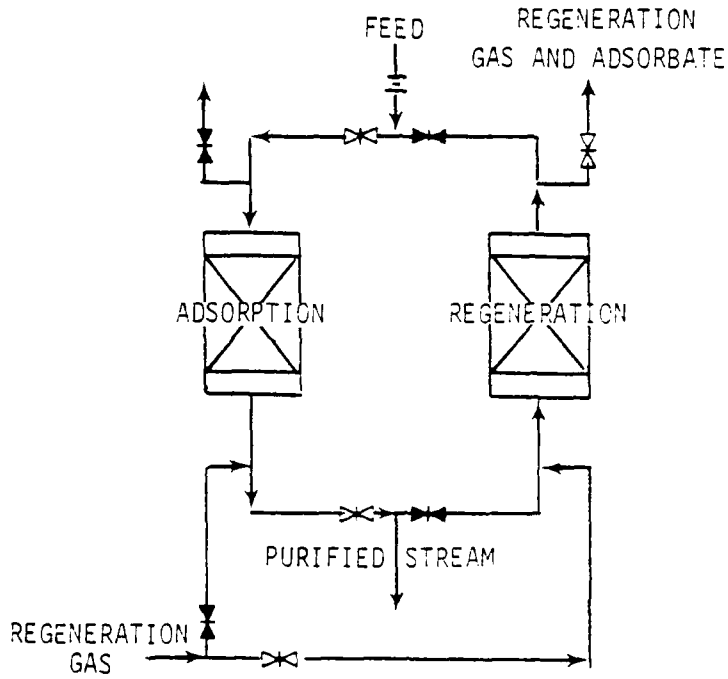

FIGURE 3-8

TWO-BED ADSORPTION SYSTEM [32]

3. Usually regenerate activated carbon with saturated steam at low pressure (5 psig).

4. Steam to solvent ratio usually varies between one and four pounds steam per pound of solvent desorbed from carbon (when volume percent solvent concentration is greater than 0.2). If solvent concentration is much less than 0.2 percent, the steam requirements can be as high as 30 pounds steam per pound of solvent.

5. Experience has shown that velocities of 60 to 70 fpm at the carbon bed are best.

TABLE 3-2

ADSORPTION ON ACTIVATED CARBON: TYPICAL PARAMETERS

Parameter	Odor Removal	Pollution Control	Solvent Recovery
Concentrations	1 PPMV*	1 PPMV-1% vol.	0.1-3% vol.
Bed depth	1/2 in.	9 in.	24 in.
Regeneration	Reactivate or discard	Regenerate in place	Steam
Adsorption cycle	1 year	1-8 hour	30-60 min
Pressure drop	0.25 in. H_2O	10-20 in. H_2O	20-30 in. H_2O
Air flow	2,000 acfm/ filter	1,000-100,000 acfm/bed	1,000-40,000 acfm/bed

3.3.4 Design Review Procedure

Step 1. Calculate the actual volumetric flow rate of gas stream.

$$acfm = scfm \times \frac{T_a}{T_r} \times \frac{P_r}{P_a} \qquad (3.12)$$

where
acfm = actual cubic foot per minute of gas stream
scfm = flow rate at standard conditions
T_r = reference temp. (60°F + 460) = 520°R
T_a = actual temperature of gas, °R
P_r = reference pressure, 14.7 psia
P_a = actual pressure, psia.

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* PPMV - parts per million (by volume)

Step 2. Calculate the required quantity of adsorbent.

(i) Volumetric flow rate of adsorbate, ft^3/hr,

V_a = acfm x (volume % of adsorbate) x 60 min/hr

(ii) Mass flow rate of adsorbate, lb/hr,

$$M = V_a \times \frac{T_r}{T_a} \times \frac{P_a}{P_r} \times \frac{1}{V_r} \times \text{(molecular wt. of adsorbate)}$$

where V_r = volume of one lb-mole of gas at standard conditions

(iii) Quantity of adsorbate to be adsorbed per cycle, lb,

$Q_a = (M)\ (E)\ (T)$

where E = % recovery required (fraction)
T = length of cycle time, hr.

(iv) Volume of adsorbent required = $\frac{Q_a(A)}{a}$ ft^3

where A = Adsorption capability, lb adsorbent/lb adsorbate
a = bulk density of adsorbent, lb/ft^3

Step 3. Calculate the required bed dimensions.

(i) Having chosen a superficial gas velocity, the cross-sectional area, S, of the bed is calculated as follows:

$$S\ (ft^2) = \frac{\text{acfm}}{\text{gas velocity (ft/min)}}$$

(ii) Diameter of the vessel, ft

$D_T = 1.13(S)^{0.5}$

(iii) Height of the bed (ft)

$$\frac{\text{Volume of adsorbent required } (ft^3)}{S}$$

Step 4. Calculate the pressure drop through the bed.[32]

$$\left(\frac{\Delta P}{z}\right) \times \frac{g_c E^3 d_p \rho_G}{(1-E)G^2} = \frac{150(1-E)}{N_{RE}} + 1.75 \qquad (3.13)$$

where ΔP = gas pressure drop, lb_f/ft^2 (lb_f = pound force)
Z = packing depth, ft
g_c = constant, 4.18×10^8 ($lb\text{-}ft/lb_f\text{-}hr^2$)
E = void volume (ft^3. voids/ft^3 packed volume)
d_p = effective particle diameter, ft
$= \frac{6\ (1-E)}{A_p}$
A_p = external surface of solid particles, ft^2/ft^3 packed volume
ρ_G = gas density, lb/ft^3
$N_{RE} = d_p\ G/\mu_G$
μ_G = gas viscosity, lb/ft-hr
G = gas mass velocity, $lb/hr\text{-}ft^2$.

3.3.5 Advantages and Disadvantages of Adsorption Systems

3.3.5.1 Advantages

1. Product recovery may be possible.

2. Excellent control and response to process changes.

3. No chemical disposal problem when pollutant (product) recovered and returned to process.

4. Capability of systems for fully automatic, unattended operation.

5. Capability to remove gaseous or vapor contaminants from process streams to extremely low levels.

3.3.5.2 Disadvantages

1. Product recovery may require an exotic, expensive distillation (or extraction) scheme.

2. Adsorbent progressively deteriorates in capacity as the number of cycles increases.

3. Adsorbent regeneration requires a steam or vacuum source.

4. Relatively high capital cost.

5. Prefiltering of gas stream may be required to remove any particulate capable of plugging the adsorbent bed.

6. Cooling of the gas stream may be required to get to the usual range of operation (less than 100°F).

7. Relatively high steam requirements to desorb high molecular weight hydrocarbons.

3.3.6 Illustrative Example

A 10,000 scfm (60°F, 1 atm) degreaser ventilation stream at 70°F and 20 psia contains 2,000 ppm by volume trichloroethylene (TCE). The contaminated air stream enters an adsorber, which is to recover 99.5 percent by weight TCE. The adsorbent is activated carbon with a bulk density of 36 lb/ft^3. The adsorption column cycle is set at 4 hr in the adsorption mode, 2 hr heating and desorbing, 1 hr cooling, and 1 hr standby. If the activated carbon is capable of adsorbing 28 lb TCE vapor/100 lb carbon before breakthrough, size the vessel.

Solution:

Step 1.

Activated carbon is used as an adsorbent.

Step 2.

$$\text{Gas flow rate} = \text{scfm} \times \frac{T_a}{T_r} \times \frac{P_r}{P_a}$$

$$= 10{,}000 \times \frac{530°R}{520°R} \times \frac{14.7\ \text{psi}}{20\ \text{psi}}$$

$$= 7{,}491.35\ \text{acfm}$$

Step 3.

(i) Volumetric flow rate of TCE,

$$V_a = 7491.35\ \text{acfm} \times 2{,}000 \times 10^{-6}$$
$$= 14.98\ ft^3/min$$
$$= 900\ ft^3/hr$$

(ii) Mass flow rate of TCE,

$$M = 900\ ft^3/hr \times \frac{520°R}{530°R} \times \frac{20\ \text{psi}}{14.7\ \text{psi}} \times \frac{1}{379\ ft^3}$$
$$\times (131.4\ \text{lb TCE}/lb_m)$$
$$= 416.5\ lb/hr$$

(iii) TCE to be adsorbed per cycle,

$$Q_a = (416.5\ lb/hr)\ (0.995)\ (4\ hr)$$
$$= 1{,}658\ lb$$

(iv) Volume of activated carbon required,

$$= \frac{(1{,}658 \text{ lb TCE})\ (100 \text{ lb carbon}/28 \text{ lb TCE})}{36 \text{ lb/ft}^3}$$

$$= 164.5 \text{ ft}^3$$

Step 4.

Superficial gas velocity of 100 ft/min is selected, for a vertical unit.

Step 5.

(i) Cross-sectional area of the bed,

$$S = \frac{7491.35 \text{ acfm}}{100 \text{ ft/min}}$$

$$= 74.91 \text{ ft}^2$$

(ii) Diameter of the vessel,

$$D_T = 1.13\ (S)^{0.5}$$
$$= 1.13\ (74.91 \text{ ft}^2)^{0.5}$$
$$= 9.78 \text{ ft}$$

(iii) Height of the bed $= \dfrac{164.5 \text{ ft}^3 \text{ of activated carbon}}{74.91 \text{ ft}^2}$

$$= 2.20 \text{ ft}$$

3.4 CONTROL BY COMBUSTION

3.4.1 Introduction

Gaseous pollutants discharged into the atmosphere can be controlled by chemical reaction. Three rapid oxidation methods are used to destroy combustible contaminants: (1) flares (direct flame combustion), (2) thermal combustion, and (3) catalytic combustion. The thermal and flare methods are characterized by the presence of a flame during combustion.

Catalytic combustion serves the same purpose, but uses a metallic catalyst to promote rapid oxidation at lower operating temperatures and with higher reaction rates.

To achieve complete combustion, i.e., the combination of the combustible elements and compounds of a fuel with all the oxygen they can utilize, sufficient space, time, and turbulence, and a temperature high enough to ignite the constituents must be provided.

The "three T's" of combustion--time, temperature, turbulence--govern the speed and completeness of the combustion reaction. For complete combustion, the oxygen must come into intimate contact with the combustible molecule at sufficient temperature, and for a sufficient length of time, in order that the reaction be completed. Incomplete reaction may result in the generation of aldehydes, organic acids, carbon, and carbon monoxide.

3.4.2 Available Equipment

3.4.2.1 Flares

3.4.2.1.1 Description

The flare system is used primarily as a safe method for disposing of excess waste gases. All process plants that handle hydrocarbons, hydrogen, ammonia, hydrogen cyanide, or other toxic or dangerous gases are subject to emergency conditions that require the immediate release of large volumes of such gases for protection of plant and personnel. Flares are used for the purpose. In operation, the gas containing the organics is continuously fed to and discharged from a stack, with the combustion occurring near the top of the stack and characterized by a flame at the end of the stack. Figure 3-9 shows a typical flare tip.

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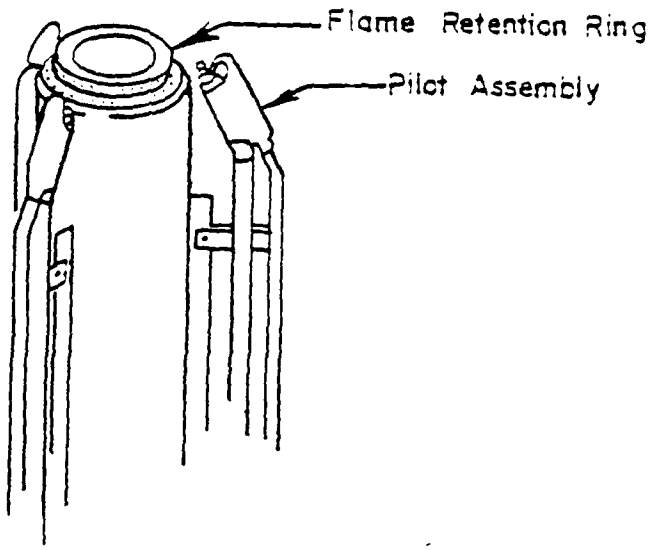

FIGURE 3-9

UTILITY FIELD FLARE TIP

3.4.2.1.2 Operating Characteristics

1. Flares generally operate in the 2,000° to 3,000°F range.
2. Used where concentration of pollutant is above the lower explosive limit (LEL) but below the upper explosive limit (UEL).*
3. Operating stack velocities are in excess of the flame propagation rate, and frequently exceed 200 ft/sec.
4. In general, a waste gas, with a heating value greater than 200 Btu/cu ft can be flared successfully. It is usually not feasible to flare a gas with a heating value below 100 Btu/cu ft.
5. It is a good practice to size flare stack on the basis of an exit velocity equal to 20 percent of the sonic velocity.

3.4.2.1.3 Advantages

1. Economical and safe disposal of large volumes of excess waste gases.
2. Low capital and operating costs.

3.4.2.1.4 Disadvantages

1. Inability to show compliance with air pollution codes by sampling the gas stream after combustion.
2. Diameter of flare stack depends upon the expected emergency gas flow rate.
3. Stack gas velocity usually limited to about 5,000 ft/sec to prevent flame extinction by blowout.
4. Forty to 50+ ft stacks require guy wires principally because a self-supporting stack at this height would require large and expensive foundations. Stacks of greater heights would require a steel structure for support and the costs would become very high.

3.4.2.2 Thermal Combustion Reactor

3.4.2.2.1 Description

Thermal incinerators or afterburners can be used over a fairly wide range of organic vapor concentrations. Reactions are conducted at elevated temperatures in order to

---

* LEL is defined as the concentration of pollutant in air, below which no explosion can take place because the heat of combustion is insufficient to propagate it. UEL is a maximum limiting concentration, above which an explosion cannot be produced because insufficient oxygen is available.

insure high chemical reaction rates for the organics. To achieve this temperature, it is necessary to preheat the feed stream with auxiliary energy. The burner may utilize air in the process waste stream as the combustion air for the auxiliary fuel, or may use a separate source of outside air for this purpose. Figure 3-10 shows a thermal combustion reactor with heat recovery.

3.4.2.2.2 Operating Characteristics

1. Unit requires operating temperatures in the 1,200° to 1,500°F range for combustion of most pollutants.

2. Residence times between 0.1 and 1 second are typical.

3. Equipment length to diameter ratio of 2.0 to 3.0 is usually specified.[32]

4. The average gas velocity at the reactor outlet can range from 10 to 50 ft/sec.

3.4.2.2.3 Advantages

1. Simplicity of construction.

2. Removal of submicron particles.

3. Virtuallly complete destruction of organic contaminants.

4. Small space requirements.

5. Low maintenance costs.

6. Capability of steam generation or heat recovery in other forms.

3.4.2.2.4 Disadvantages

1. High operating costs.

2. Potential for flashback and subsequent explosion hazard.

3.4.2.3 Catalytic Combustion Reactor

3.4.2.3.1 Description

Catalytic reactors are an alternative to thermal incinerators. For simple reactions, the effect of the presence of a catalyst is to:

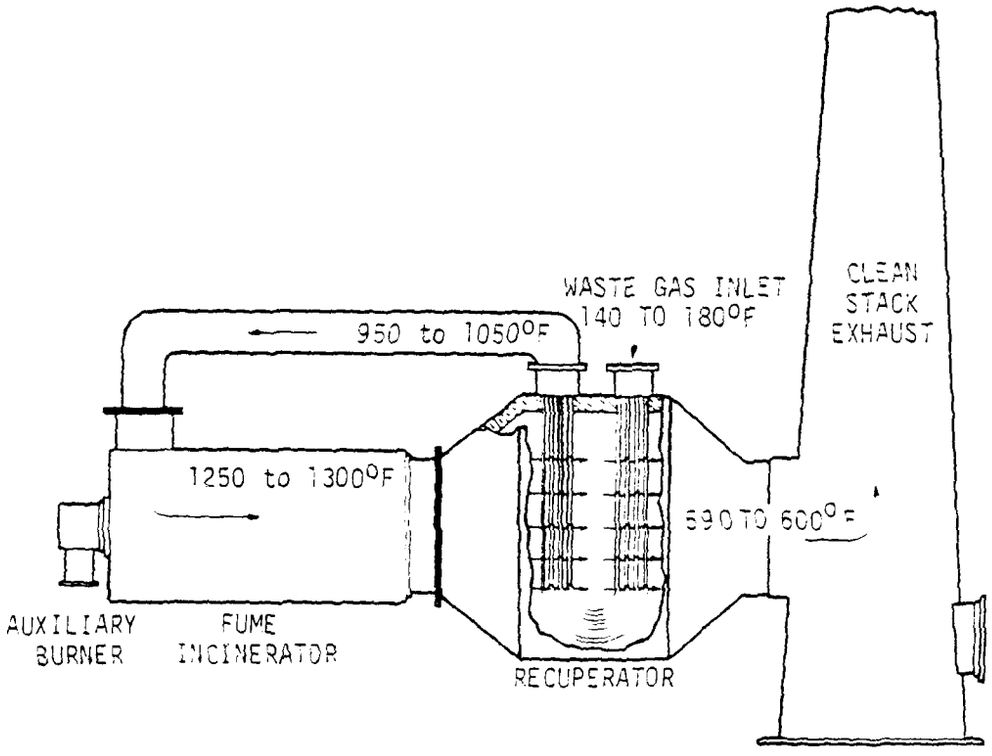

FIGURE 3-10

THERMAL COMBUSTION WITH ENERGY (HEAT) RECOVERY [32]

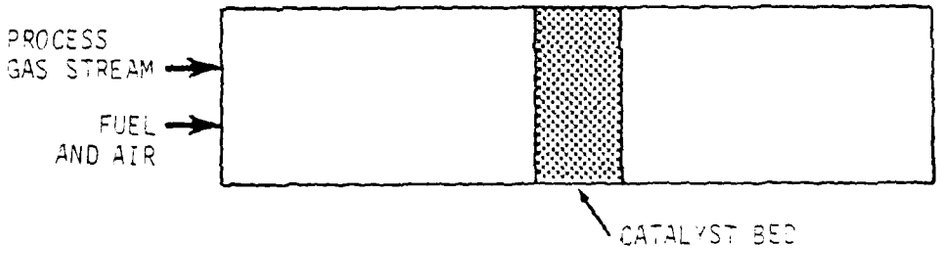

FIGURE 3-11

CATALYTIC COMBUSTION REACTOR [32]

1. Increase the rate of reaction.

2. Permit the reaction to occur at a lower temperature.

3. Permit the reaction to occur at a more favorable pressure.

4. Reduce the reactor volume.

5. Increase the yield of a reactant(s) relative to the other(s).

Metals in the platinum family are recognized for their ability to promote combustion at relatively low temperatures. Other catalysts include various oxides of copper, chromium, vanadium, nickel, and cobalt. Figure 3-11 shows a catalytic reactor.

3.4.2.3.2 Operating Characteristics

1. The gas stream is usually delivered to the reactor by a fan at a velocity in the 10 to 30 ft/sec range.

2. Reactor temperature is usually in the 650° to 800°F range.

3. A reactor length to diameter ratio less than 0.5 is usually specified.

4. The smaller the size of the catalyst used, the more efficiently it operates; however, the pressure drop through the reactor will increase.

5. Typically 0.5 to 2 cu ft catalyst/1,000 scf waste stream for 85 to 95 percent conversion of hydrocarbons in the waste stream.

3.4.2.3.3 Advantages

1. Low fuel requirements.

2. Little or no insulation requirements.

3. Reduced flashback problems and explosion hazards.

3.4.2.3.4 Disadvantages

1. High initial cost.

2. Catalyst poisoning.

3. Particulate must first be removed.

4. Catalyst regeneration problems.

3.4.3 Design Review Procedure

3.4.3.1 Flares

Step 1. Calculate the diameter of flare tip, inches,

$$d = \left(\frac{W}{1,370}\sqrt{\frac{T}{M}}\right)^{0.5} \qquad (3.14)$$

where
W = mass flow rate of waste gases lb/hr
T = temperature of vapor, °R
M = molecular weight of vapor stream.

Step 2. Calculate the required stack height.

The height of a stack is a function of its relative distance from personnel or equipment.

Recommended heat intensity values are listed below:

Operating personnel - 1,500 Btu/hr-ft^2
Equipment - 3,000 Btu/hr-ft^2

$$q = \frac{960\ W\sqrt{M}}{4\ [X^2 + H(H + 120D)]} \qquad (3.15)$$

where
q = heat intensity, Btu/hr-ft^2
W = vapor flow rate, lb/hr
M = molecular weight of vapor to be burned
X = distance from base of stack to nearest operating personnel or equipment, ft
H = stack height, ft
D = diameter of stack, ft.

Step 3. Calculate the steam required for smokeless flaring.

$$W_{steam} = W_{HC}\ (0.68 - 10.8/M) \qquad (3.16)$$

where
W_{steam} = lb/hr of steam
W_{HC} = lb/hr of hydrocarbon
M = molecular weight of hydrocarbon.

3.4.3.2 Thermal Combustion Reactors

Step 1. Calculate combustion chamber volume.

$$V = \frac{(acfm)}{60}\ t$$

where V = combustion chamber volume (following burner and prior to heat exchangers, dilution air, etc.)

acfm = exhaust gas flow rate to afterburner, actual ft^3/min

t = total average residence time, seconds

Step 2. Calculate the fuel required to sustain combustion.

Most industrial combustion calculations for fuel requirements involve a two-step calculation.

(1) The heat load (Btu/min) required to raise the temperature of the process gas stream to the desired combustion temperature.

(2) The energy released (Btu/scf of fuel) and available for heating the gas stream to combustion temperature. This term is often referred to as the available heat for the fuel (HA).

Dividing (1) by (2) yields the fuel rate in scfm. This rough calculation provides a result that is often satisfactory for engineering (design) purposes.

3.4.3.3 Catalytic Combustion Reactors

Variations and unknowns in many of the parameters in the describing equation(s) make it nearly impossible to provide a generalized procedure for the design of a catalytic reactor.

3.4.4 Advantages and Disadvantages of Combustion Systems

3.4.4.1 Advantages

1. Simplicity of operation.

2. Capability of steam generation or heat recovery in other forms.

3. Virtually complete destruction of organic contaminants.

3.4.4.2 Disadvantages

1. Relatively high operating costs (particularly associated with fuel requirements).

2. Potential for flashback and subsequent explosion hazard.

3. Catalyst poisoning (in the case of catalytic incineration).

3.4.5 Illustrative Examples

3.4.5.1 Example 1--Thermal Combustion

A total of 35,900 scfm (32°F, 1 atm) of process gas at 400°F is to be combusted at 1,400°F to remove the pollutants. Assume that the fuel is natural gas (heat content - 690 Btu/scf fuel), stoichiometric external air is supplied for combustion, and that the gas stream is essentially air. Design a thermal reactor for the purpose. Also assume that this application requires a residence period of 0.5 sec. The bulk gas velocity is set at 20 ft/sec.

Solution:

Step 1.

Btu content of the waste gas is accounted for in Step 3.

Step 2.

Molar flow rate of gas stream

$$= \frac{35{,}900 \text{ scfm}}{359 \text{ ft}^3/\text{lb-mole}} = 100 \text{ lb-moles/min.}$$

Step 3.

The average heat capacity of air over the temperature range in question is approximately 7.5 Btu/lb-mole-°F. The enthalpy change of the gas stream is then:

$$H = 7.5 \text{ Btu/lb-mole-°F } (1{,}400°F - 400°F)$$
$$= 7{,}500 \text{ Btu/lb-mole}$$

The energy requirement rate, Q, is given by:

$$Q = (7{,}500 \text{ Btu/lb-mole}) \times (100 \text{ lb-moles/min})$$
$$= 750{,}000 \text{ Btu/min}$$

The fuel rate, FR, is simply obtained by dividing Q by heat content of fuel. Thus,

$$FR = \frac{750{,}000 \text{ Btu/min}}{690 \text{ Btu/scf}}$$

$$= 1{,}088 \text{ scfm}$$

Step 4.

Retention time = 0.5 sec (given)
Combustion temperature = 1,400°F (given)

Step 5.

One can show that there are approximately 11.5 scf of combustion products for every 1.0 scf of natural gas burned in stoichiometric air. Thus, after fuel combustion, the combustion products flow rate is:

11.5 (1,088 scfm) = 12,500 scfm

Since the process gas stream flow rate is 35,900 scfm, the <u>total</u> flue gas flow rate in the reactor section is:

35,900 + 12,500 = 48,400 scfm

The <u>actual</u> flow rate, acfm, is:

acfm = 48,400 scfm x (1,860°R/492°R) = 183,000

The cross-sectional area, A, required for flow is then:

$A = (183{,}000 \text{ acfm})/(20 \text{ ft/sec})\ (60 \text{ sec/min}) = 152.5 \text{ ft}^2$

If one reactor is employed, the diameter is:

$D = 1.13\ (A)^{0.5} = 1.13\ (152.5 \text{ ft}^2)^{0.5} = 13.95 \text{ ft}$

The length of the reactor, L, is obtained from residence time information.

L = (20 ft/sec) (0.5 sec) = 10 ft

The L/D ratio is:

L/D = 10 ft/13.95 ft = 0.717

This value is rather low. Several smaller diameter units, with the same total volume, are suggested to bring the L/D ratio above 2.0.

3.4.5.2 <u>Example 2--Catalytic Combustion</u>

A feed rate of 25 lb/min of a certain pollutant is introduced into a catalytic reactor. A conversion of 0.74 is required in order to meet new local pollution standards. Determine the mass of catalyst and the number of 20-ft long, 1.5-in. 10 Gauge (GA) tubes necessary to achieve this conversion. The rate of reaction may be represented as a function of the conversion by:

$$R_A = -0.15\ (1 - x_A)\ \ :lb_m/(lb_{cat})\ (min) \qquad (3.17)$$

where, R_A = rate of reaction of pollutant

x_A = conversion (fraction)

The bulk density of the catalyst is 36 lb/ft^3.

<u>Solution</u>:

(i) $$W = -F_A \int_0^{x_A} \frac{dx_A}{R_A} \qquad (3.18)^{32}$$

where W = mass of catalyst required, lb.

F_A = feed rate of pollutant = 25 lb/min

x_A = required conversion = 0.74

substituting for R_A, F_A and x_A,

$$W = 25 \int_0^{0.74} \frac{dx_A}{0.15(1-x_A)}$$

$$= \frac{25}{0.15} \left[-\ln\ (1-x_A) \right]_0^{0.74}$$

$$= 224.5 \text{ lb catalyst}$$

(ii) Volume of the reactor, V, is given by

$$V = \frac{224.5 \text{ lb}}{36 \text{ lb/ft}^3} = 6.24 \text{ ft}^3$$

(iii) Cross-sectional area of 1.5 in. 10 BWG tube is 0.00828 ft^2.

Then, volume/tube = (0.00828 ft^2) (20 ft) = 0.1656 ft^3.

No. of tubes = 6.24 ft^3/0.1656 ft^3 = 37.7

The design calls for 38 tubes.

3.5 CONTROL BY CONDENSATION

3.5.1 Introduction

Condensers can be used to collect condensable emissions discharged to the atmosphere, particularly when the vapor concentration is high. Condensation is usually accomplished by lowering the temperature of the gaseous stream, although an increase in pressure will produce the same result. The former approach is usually employed by industry because pressure changes (even small ones) on large volumetric gas flow rates are often economically prohibitive.

There are two basic types of condensers used for control --contact and surface. In contact condensers, the gaseous stream is brought into direct contact with a cooling medium so that the vapors condense and mix with the coolant (Figure 3-12). The more widely used system, however, is the surface condenser (or heat exchanger) in which the vapor and cooling medium are separated by a wall (Figure 3-13). Since high removal efficiencies cannot be obtained with low condensable vapor concentrations, condensers are typically used for pretreatment prior to some other more efficient control device such as a reactor (incineration), absorber, or adsorber.

3.5.2 Description of Available Equipment

3.5.2.1 Direct Contact Condensers

In a direct-contact condenser, a stream of water or other cooling liquid is brought into direct contact with the vapor to be condensed. The liquid stream leaving the chamber contains the original cooling liquid plus the condensed substances. The gaseous stream leaving the chamber contains the noncondensable gases and such condensable vapors as did not condense; it is reasonable to assume that the vapors in the leaving gas stream are saturated. It is then the temperature of the leaving gas stream that determines the collection efficiency of the condenser.

3.5.2.2 Surface Condensers

Surface condensers are usually referred to as heat exchangers in the chemical process industry. Surface condensers are employed in the air pollution equipment industry for recovery, control, and/or removal of trace impurities or contaminants. In the surface type, the coolant does not contact the gas stream. There are various types of surface condensers including the shell and tube, fin-fan, finned hairpin, finned tube section, and tubular.

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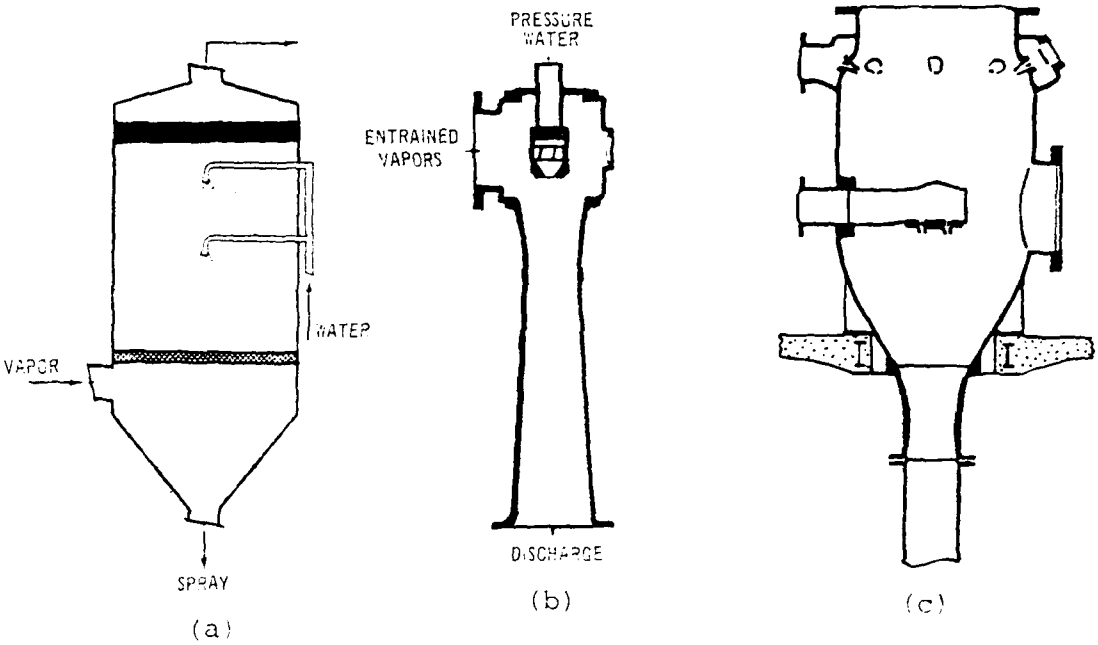


FIGURE 3-12

DIRECT CONTACT CONDENSERS

(a) spray chamber (b) Jet (c) Barometric

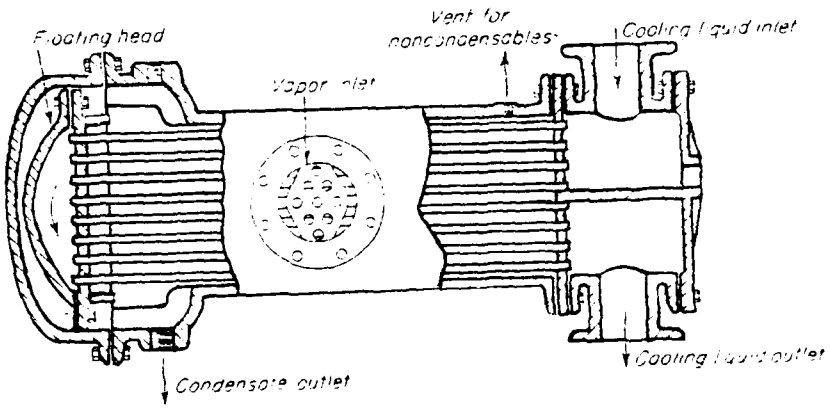


FIGURE 3-13

SURFACE CONDENSER

3.5.3 Operating Characteristics

1. In a surface condenser, the overall heat transfer coefficient is usually low, on the order of 10 to 25 Btu/hr-ft^2-°F for organic vapor/noncondensable gas systems.

2. In a contact type condenser, approximately 16 pounds of water at 70°F are required to condense 1 pound of steam at 250°F and cool the condensate at 140°F.

3.5.4 Design Review Procedure

3.5.4.1 Direct Contact Condensers

Direct contact condensers involve the simultaneous transfer of heat and mass. Design procedures available for absorption, humidification, cooling towers, etc. may be applied with some modifications. However, detailed analysis of this material is beyond the scope of this manual and, therefore, is not included. The reader is referred to any one of a number of excellent heat and mass transfer texts, especially that by Treybal.[31]

In practice, very crude empirical correlations, not available in the literature, are usually employed in designing and predicting performance of this type of unit.

3.5.4.2 Surface Condensers

The usual energy (enthalpy) balance calculations can be applied to determine the rate of heat transfer in a condenser.

$$Q = H_F - H_V - H_L \qquad (3.19)$$

where

Q = rate of heat transfer
H = enthalpy with the subscripts F, V, and L referring to the feed, vapor, and liquid, respectively.

Standard equations are employed to design the heat exchangers:

$$Q = UA\,\Delta T \qquad (3.20)$$

where

A = surface area required
Q = rate of heat transfer
U = overall heat transfer coefficient
ΔT = temperature difference driving force.

The cooling medium is typically on the tube side of the condenser with the vapor on the shell side. Water or some type of refrigerant is often used as the cooling medium, depending on the desired condensing temperature.

3.5.5 Advantages and Disadvantages of Condenser Systems

3.5.5.1 Advantages

1. Pure product recovery (in the case of surface condensers).

2. Water used as the coolant in a surface condenser does not contact the contaminated gas stream and can be reused after cooling.

3.5.5.2 Disadvantages

1. Relatively low removal efficiency for gaseous contaminants (at concentrations typical of pollution control applications).

2. Coolant requirements may be extremely expensive.

3.5.6 Illustrative Example

A 23,150 ft^3/min stream at 530°F and 14.9 psia containing 60 percent air, 4 percent n-octane, 24 percent n-nonane, and 12 percent n-decane is to be cooled to the temperature of 150°F. Calculate the mole-fraction of each hydrocarbon in the vapor phase of the exit stream. Size the unit required to accomplish this separation.

Solution

(i) Mole-fractions of all the components in the condensate and the exit gas stream are first calculated.

Basis: 1.0 mole of feed

At equilibrium,

$$y_i = K_i X_i \qquad \text{(a)}$$

where y_i = mole-fraction of any component i in vapor phase
X_i = mole-fraction of any component i in liquid phase
K_i = equilibrium constant of component i

Also, $$X_i = \frac{Z_i}{L + K_i\,(1-L)} \qquad \text{(b)}$$

where Z_i = mole-fraction of component i in feed stream
L = fraction of feed condensed.

For different values of L, X_i is calculated using (b) for all the components in feed, and ΣX_i is determined. Value of L corresponding to $\Sigma X_i = 1$ gives the fraction of feed condensed. For these values of X_i values of y_i are determined using (a). Trial and error is required for the purpose.

L = 0.37, when used in (b), satisfies $X_i = 1$. Corresponding values of y_i are calculated and given below.

Component	K at 150°F and 14.9 psia		yi
Air			0.9530
Octane	0.15		0.0129
Nonane	0.05		0.0295
Decane	0.015		0.0050
		Σ =	1.0004

The fraction of the inlet hydrocarbons remaining in the vapor phase is therefore

$$\frac{(0.0474)\ (0.63)}{(0.4)} = 0.075 \text{ or } 7.5\%$$

The vapor removal efficiency of the condenser operating at 150°F and 14.9 psia is then

$$1 - 0.075 = 0.925 \text{ or } 92.5\%$$

The design calculations for the condenser (heat exchanger) are presented below.

(ii) Molar flow rate of the feed is determined.

Volumetric flow rate = 23,150 ft^3/min
= 1,389,000 ft^3/hr

Average molecular weight of feed, M, is given by

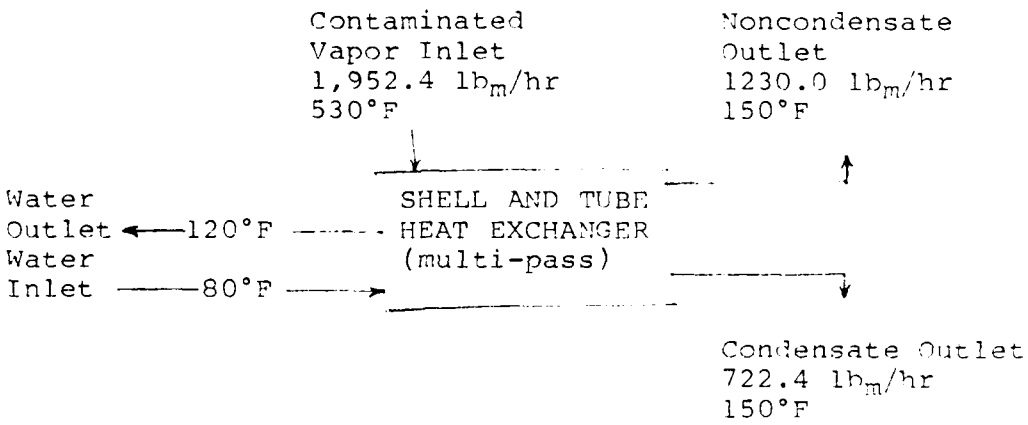
$$M = 0.60\ (29) + 0.04\ (114) + 0.24\ (128) + 0.12\ (142)$$
$$= 69.72\ lb/lb_m\ (lb_m = \text{pound-mole})$$

The molar flow rate is then,

$$(1{,}389{,}000\ ft^3/hr) \times \frac{520°R}{990°R} \times \frac{14.9\ psi}{14.7\ psi}\left(\frac{1}{379\ ft^3/lb_m}\right)$$

$$= 1{,}952.4\ lb_m/hr$$

A line diagram of the system is shown below.



To determine the heat transferred and water requirements, it is necessary to perform a heat balance.

$$H = m\langle c_p \rangle \Delta T$$

where $\langle c_p \rangle$ = average heat capacity, Btu/lb_m-°F

Basis: Reference temperature of 0°F, 1 hr

<u>Heat Balance</u>

In the following calculations to determine the amount of heat that must be removed in the condenser, heat capacity values are averaged over the temperature range indicated (530° to 150°F).

1. 1,171.4 lb_m/hr air is cooled from 530°F to 150°F:
   1,171.4 (7.325 Btu/lb_m°F) (530-150)°F=3.26 MM Btu/hr

2. 15.87 lb_m/hr octane (g) is cooled from 530°F to 150°F:
   15.87 (61.6 Btu/lb_m°F) (530-150)°F = 0.37 MM Btu/hr

3. 62.23 lb_m/hr octane (g) is cooled from 530°F to 150°F:

   62.23 (61.6 Btu/lb_m°F) (530-150)°F = 146 MM Btu/hr

4. 62.23 lb_m/hr octane is condensed at 150°F:
   62.23 (15918.50 Btu/lb_m) = 0.99 MM Btu/hr

5. 36.29 lb_m/hr nonane (g) is cooled from 530°F to 150°F:

   36.29 (69.05 Btu/lb_m°F) (530-150)°F = 0.95 MM Btu/hr

6. 432.3 lb_m/hr nonane (g) is cooled from 530°F to 150°F:
432.3 (69.05 Btu/lb_m°F) (530-150)°F = 11.34 MM Btu/hr

7. 432.3 lb_m/hr nonane is condensed at 150°F:
432.3 (17603.56 Btu/lb_m) = 7.61 MM Btu/hr

8. 6.15 lb_m/hr decane (g) is cooled from 530°F to 150°F:
6.15 (76.55 Btu/lb_m°F) (530-150)°F = 0.18 MM Btu/hr

9. 228.2 lb_m/hr decane (g) is cooled from 530°F to 150°F:
228.2 (76.55 Btu/lb_m°F) (530-150)°F = 6.64 MM Btu/hr

10. 228.2 lb_m/hr. decane is condensed at 150°F:
228.2 (18355.32 Btu/lb_m) = 4.19 MM Btu/hr

Total Amount of Heat to be Removed = 37.00 MM Btu/hr

The quantity of water required to remove this amount of heat is:

$$(m_{H_2O})\ (C_{p'H_2O})\ (\Delta T) = 37 \times 10^6 \text{ Btu/hr}$$

$$m_{H_2O} = \frac{37 \times 10^6 \text{ Btu/hr}}{(1 \text{ Btu/lb-°F})\ (120-80)\text{°F}}$$

$$= 924{,}750 \text{ lb/hr}$$

or approximately 1,850 gpm.

A value of 20 Btu/hr-ft^2-°F will be employed for U, the overall heat transfer coefficient, and it will be assumed constant. Note: A more rigorous approach to calculating U may be found in a heat transfer book.[33]

$$Q = (f)\ (U)\ (A)\ (\Delta T)$$

T = logarithmic mean temperature difference = 155.6°F
f = 0.9 (one shell/four tube passes)[33]

The area required (under ideal conditions) is then:

$$A = (37 \times 10^6 \text{ Btu/hr})/(0.9)(20 \text{ Btu/hr-ft}^2\text{-°F})(155.6\text{°F})$$
$$= 13{,}210 \text{ ft}^2$$

In summary, the condenser design specifications are:

Q = 37 MM Btu/hr
A = 13,210 ft^2
m_{H_2O} = 1,850 gpm.

4. COSTING METHODS FOR AIR POLLUTION CONTROL EQUIPMENT

4.1 Introduction

4.2 Estimating Equipment Cost

4.3 Estimating Operating and Maintenance Costs

4.4 Updating Costs to Future Periods

4.5 Data for Estimating Capital Equipment
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- 4.5.2 Fabric Filters
  - 4.5.2.1 Intermittent Pressure, Mechanical Shaker Baghouse
  - 4.5.2.2 Continuous Suction or Pressure, Pulse-Jet Baghouse
  - 4.5.2.3 Continuous Pressure, Mechanical Shaker Baghouse
  - 4.5.2.4 Continuous Pressure, Reverse Air Baghouse
  - 4.5.2.5 Custom Pressure on Suction Baghouse
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- 4.5.4 Mechanical Collectors
- 4.5.5 Absorption Units--Packed and Plate Towers
- 4.5.6 Adsorbers
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  - 4.5.7.1 Thermal Incinerators Without Heat Exchangers
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- 4.5.10 Quenchers
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- 4.5.12 Fans and Motors
  - 4.5.12.1 Backward-Curved Blade Fans
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  - 4.5.13.1 Straight Duct
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  - 4.5.13.4 Stacks

4.6 Estimating Operating and Maintenance Costs
- 4.6.1 Particulate Pollutants Control Devices
  - 4.6.1.1 Cyclone Collectors
  - 4.6.1.2 Electrostatic Precipitators
  - 4.6.1.3 Fabric Filter Systems
  - 4.6.1.4 Wet Scrubbers
- 4.6.2 Gaseous Pollutants Control Devices
  - 4.6.2.1 Carbon Adsorbers
  - 4.6.2.2 Thermal Incinerators
  - 4.6.2.3 Catalytic Incinerators
- 4.6.3 Maintenance and Installation Cost Factors and Equipment Life Guidelines

4.7 Life Cycle Costing

CHAPTER 4

COSTING METHODS FOR AIR POLLUTION CONTROL EQUIPMENT

4.1 INTRODUCTION

There are usually many alternatives available to resolve an air pollution emission control problem. To select the optimum method, each solution should be thoroughly evaluated before implementation. Steps such as substitution of fuels and raw materials or modification of processes should not be overlooked as possible solutions.

The first consideration is the degree of reduction of emissions necessary to meet emission standards. The degree of emission reduction or the collection efficiency required is dependent upon the relationship between emissions and emission standards. The important factors to be considered next are the gas stream and gaseous contaminant characteristics. High gas temperatures, for example, usually preclude the use of adsorbers. Each alternative will have specific capital and operating costs associated with it, and the components of this cost should be carefully examined. Those alternatives meeting the requirements of both the process and the plant facility can then be evaluated in terms of uniform annual cost and, on this basis, the gas cleaning system can be selected.

The actual cost of installing and operating air pollution control equipment is a function of many direct and indirect cost factors. An analysis of the control costs for a specific source should include an evaluation of all relevant factors. The control system must be designed and operated as an integral part of the process; this will minimize the cost of control for a given emission level. The definable control costs are those that are directly associated with the installation and operation of control systems. These expenditure items, from the control equipment user's point of view, have a breakdown for accounting purposes as follows:

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DEVELOPMENT OF DESIGN REVIEW PROCEDURES FOR ARMY AIR POLLUTION --ETC(U)
JUL 80 A J BUONICORE, J P BILOTTI DAMD17-79-C-9051

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1. Capital Investment

   Engineering studies
   Land
   Control hardware
   Auxiliary equipment

2. Operating and Maintenance Costs

   Pressure drop across the system
   Electricity, fuel, and scrubbing liquor
   Maintenance costs

3. Capital Charges

   Interest
   Taxes
   Insurance

4.2 ESTIMATING EQUIPMENT COSTS

The general cost curves and equations presented in this chapter represent "ballpark" equipment costs for the various types of emission control devices. Installation costs for the equipment will depend on such factors as physical location of the equipment within the plant, degree of assembly, availability of utilities, overtime requirements, equipment transportation and difficulty of loading/unloading, and complexity of instrumentation and controls.

Turn-key cost estimates by most suppliers also include engineering and contingency costs. Engineering is generally estimated at 10 percent of the total equipment and installation cost. This includes start-up and performance testing besides the normal system design engineering. Contingencies are also included in the cost estimates. These contingencies cover unexpected costs due to inflation, union slowdowns and strikes, delays in receipt of materials, start-up and guarantee testing problems, subcontractor price adjustments, and other unforeseen problems. Contingency costs are generally estimated at 10 percent of the total costs.

The capital costs for a control system are itemized as follows:

(1)	Estimated equipment costs (control device + auxiliaries)	= $_____
(2)	Tax and freight at 7% of (1)	= $_____
(3)	Installation costs (see Table 4-15)	= $_____
(4)	Subtotal (1 + 2 + 3)	= $_____
(5)	Engineering at 10% of (4)	= $_____
(6)	Subtotal (4 + 5)	= $_____
(7)	Contingencies at 10% of (6)	= $_____
(8)	Total estimated capital costs (6 + 7)	= $_____

4.3 ESTIMATING OPERATING AND MAINTENANCE COSTS

The cost of operating and maintaining control equipment depends on such factors as the quality and suitability of the control equipment and the user's understanding and vigilance in its operation. Management policies and attitude also play a role in this area of cost.

Maintenance and operating costs are difficult to assess, but are frequently a significant portion of the overall annual cost of controlling air pollutant emissions. Although combined operating and maintenance costs may be as low as 15 percent of the annualized total cost for dry centrifugal collectors and electrostatic precipitators, they may be as high as 90 percent for a high-efficiency wet collector.

The annual operating cost is the expenditure incurred in operating a control device at its designed collection efficiency for a period of one year. This cost depends on and accounts for the following factors: (1) gas volume cleaned; (2) pressure drop across the system; (3) total time the device is operated; (4) consumption and costs for electricity, fuel, and scrubbing liquor; and (5) mechanical efficiencies of the fan and the pump. Section 4.6 gives the theoretical cost equations developed to reflect these factors.

Maintenance cost is defined as the expenditure required to sustain the operation of a control device at its designed efficiency. This objective is best accomplished with a scheduled maintenance program and proper replacement of defective and worn parts. The maintenance cost is assumed to be proportional to the gas flow rate of the device. Maintenance cost is an estimate based on present knowledge and understanding of control system applications. These costs are expressed as average annual values computed over the useful life of the devices. Actually, maintenance costs, as in any mechanical device, will increase with the age and wear of the equipment. Factors for calculating maintenance costs are presented in Table 4-15.

4.4 UPDATING COSTS TO FUTURE PERIODS

The equations given in the following sections of this chapter contain correction factors to update the costs through September 1979 levels.

No attempt is made to predict future costs, since this is beyond the scope of this manual. In general, it is recommended that equipment costs be updated using the Chemical Engineering Fabricated Equipment Index. Table 4-1 summarizes the annual index through September 1979.

TABLE 4-1

CHEMICAL ENGINEERING FABRICATED EQUIPMENT COST INDEX[34]

Year	Index
October 1, 1979	266.5
1978	238.6
1977	216.6
1976	200.8
1975	192.2
1974	170.1
1973	142.5
1972	136.3
1971	130.3
1970	122.7
1969	115.1
1968	109.9
1967	106.2
1966	104.8
1965	103.4
1964	102.7
1963	101.7
1962	101.0
1961	100.1
1960	101.2

4.5 DATA FOR ESTIMATING CAPITAL EQUIPMENT COSTS[35]

4.5.1 Electrostatic Precipitators

The Deutsch equation for electrostatic precipitators is given below in Equation 4.1, and the area solved for in Equation 4.2. Prices are generally a function of net plate area (or specific collecting area, SCA).

$$\eta = 1.0 - e^{(wA/Q)} \qquad (4.1)$$

or

$$A = -Q \ln (1-\eta)/w \qquad (4.2)$$

where

η = efficiency
w = drift velocity, ft/sec
A = specific collecting area, ft^2
Q = flow rate, actual ft^3/sec
e = Naperian log base.

For example, for grey iron foundries, the precipitation rate parameter, w, is typically 0.12 ft/sec. If 99 percent cleaning efficiency is required on a flow rate of 10,000 acfm into the precipitator, the specific collecting area is calculated as follows:

$$A = (-10{,}000 \text{ cfm}) (\ln(1-.99))/(0.12 \text{ ft/sec} \times 60 \text{ sec/min}) = 6{,}396 \text{ ft}^2.$$

Use 6396 ft^2 for the area in Equation 4.3 to find the price for the insulated precipitator or Equation 4.4 for the uninsulated precipitator, depending on design requirements.

Dry-type Electrostatic Precipitator Purchase Price ($) vs. Plate Area ($ft^3$)	
• Insulated Electrostatic Precipitator $ = 172,898 + 5.28 x Area	(4.3)
• Uninsulated Electrostatic Precipitator $ = 103,739 + 3.43 x Area	(4.4)
(Equations reflect October 1979 costs.)	

4.5.2 Fabric Filters

Prices for mechanical shaker, pulse-jet, reverse-air, and custom fabric filters (baghouses) are listed in Equations 4.5 through 4.22. Prices are based on net cloth

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TABLE 4-2

BAG PRICES ($/ft^2 Cloth Area)

Class	Type	Dacron	Orlon	Nylon	Nomex	Glass	Polypropylene	Cotton
Standard	Mechanical Shakers							
	<20,000 ft^2 (cloth area)	0.46	0.79	0.92	1.44	0.59	0.79	0.52
	>20,00 ft^2	0.39	0.66	0.81	1.31	0.52	0.66	0.46
	Pulse Jet*	0.72	1.18	-	1.64	-	0.90	-
	Reverse Air	0.39	0.72	0.85	1.31	0.52	0.66	0.46
Custom	Mechanical Shaker	0.26	0.39	0.52	0.79	0.33	0.39	0.46
	Reverse Air	0.26	0.39	0.52	0.79	0.33	0.39	0.46

*For heavy felt, multiply prices by 1.5.
(Numbers reflect October 1979 costs.)

TABLE 4-3

APPROXIMATE GUIDE TO ESTIMATE GROSS CLOTH AREA

Net Cloth Area (sq ft)	Gross Cloth Area (sq ft)
1- 4,000	Multiply by 2
4,001- 12,000	Multiply by 1.5
12,001- 24,000	Multiply by 1.25
24,001- 36,000	Multiply by 1.17
36,001- 48,000	Multiply by 1.125
48,001- 60,000	Multiply by 1.11
60,001- 72,000	Multiply by 1.10
72,001- 84,000	Multiply by 1.09
84,001- 96,000	Multiply by 1.08
96,001-108,000	Multiply by 1.07
108,001-132,000	Multiply by 1.06
132,001-180,000	Multiply by 1.05
180,001 and greater	Multiply by 1.04

area, which is calculated by dividing the gas volume entering the baghouse by the required air-to-cloth (A/C) ratio. For example, to handle 100,000 acfm, an A/C = 2.0 requires 50,000 ft^2 net cloth area. The prices for bags may be determined from Table 4-2. From Table 4-3, obtain factors to calculate gross cloth area (for example, at 50,000 ft^2 the factor is 1.11). The price of glass bags for the example is thus:

$$50{,}000\ ft^2 \times 1.11 \times \$0.52/ft^2 = \$28{,}860$$

Extras such as stainless steel construction, insulation, and suction are add-ons to the basic baghouse cost. Basic baghouse prices are flange-to-flange, not including bags. They do, however, include a 10-foot support clearance and inlet and exhaust manifolds. Pressure baghouses are designed for 12 inches of water and suction baghouses are designed for 20 inches of water. Custom baghouses are more a function of specific requirements than of operating pressure construction. Hence, prices for custom units do not differentiate between pressure or suction. All baghouses are priced as factory assembled.

4.5.2.1 Intermittent Pressure, Mechanical Shaker Baghouse Price ($) vs. Net Cloth Areas ($ft^2$)

Basic Price (Bags Not Included)	$ = 4,025 + 2.22 (Net Area)	(4.5)
Stainless Steel Add-On	$ = 2,956 + 1.21 (Net Area)	(4.6)
Insulation Add-On	$ = 2,376 + 0.98 (Net Area)	(4.7)
Suction Add-On	$ = 1,658 + 0.14 (Net Area)	(4.8)

(Equations reflect October 1979 costs).

4.5.2.2 Continuous Suction or Pressure, Pulse-Jet Baghouse Price ($) vs. Net Cloth Area ($ft^2$)

Basic Price (Bags Not Included)	$ = 6,411 + 9.13 (Net Area)	(4.9)
Stainless Steel Add-On	$ = 1,784 + 5.38 (Net Area)	(4.10)
Insulation Add-On	$ = 5,740 + 2.77 (Net Area)	(4.11)

(Equations reflect October 1979 costs).

4.5.2.3 Continuous Pressure, Mechanical Shaker Baghouse Price ($) vs. Net Cloth Area ($ft^2$)

Basic Price (Bags Not Included)	$ = 7,993 + 4.15 (Net Area)	(4.12)
Stainless Steel Add-On	$ = 7,939 + 2.11 (Net Area)	(4.13)
Insulation Add-On	$ = 2,671 + 2.11 (Net Area)	(4.14)
Suction Add-On	$ = 2,714 + 0.31 (Net Area)	(4.15)

(Equations reflect October 1979 costs.)

4.5.2.4 Continuous Pressure, Reverse-Air Baghouse Price ($) vs. Net Cloth Area ($ft^2$)

Basic Price (Bags Not Included)	$ = 30,342 + 3.59 (Net Area)	(4.16)
Stainless Steel Add-On	$ = 12,565 + 1.93 (Net Area)	(4.17)
Insulation Add-On	$ = 12,680 + 1.93 (Net Area)	(4.18)
Suction Add-On	$ = 2,033 + 0.39 (Net Area)	(4.19)

(Equations reflect October 1979 costs).

4.5.2.5 Custom Pressure or Suction Baghouse Price ($) vs. Net Cloth Area ($ft^2$)

Basic Price (Bags Not Included)	$ = 140,535 + 3.74 (Net Area)	(4.20)
Stainless Steel Add-On	$ = 69,159 + 1.93 (Net Area)	(4.21)
Insulation Add-On	$ = 52,561 + 1.93 (Net Area)	(4.22)

(Equations reflect October 1979 costs.)

4.5.3 Venturi Scrubbers

The equipment cost for a venturi scrubber can be determined using Equation 4.23. To price a scrubber using these curves and equations, use the following steps:

A. Determine gas volume entering the venturi section and use this value in thousands acfm for a 1/8-in. thick carbon steel (C.S.) in Equation 4.23. For example, at 100,000 acfm, the price is approximately $47,200.

B. Determine the pressure drop across the scrubber required to obtain the desired efficiency (see Section 2.3.3) and find the required metal thickness for the design inlet volume[35] from Figure 4-1. For 100,000 acfm and 30-in. ΔP, the required metal thickness is 1/4-in. plate (always come up to the next standard plate).

C. From Figure 4-2, find the price adjustment factor for the design inlet volume and the material thickness[35] found in Step B. For 100,000 acfm and 1/4-in. plate, the factor is approximately 1.6. Thus, the carbon steel scrubber price is $47,200 x 1.6 = $75,520.

D. If stainless steel (S.S.) construction, rubber or fiberglass lining, or variable venturi section is to be included, refer to Table 4-4 and adjust the price accordingly. For stainless steel construction, the adjusted price would be $75,520 x 1.8 = $135,940.

Note: Radial tip fans are commonly used in scrubbers.

Fabricated 1/8-in. Thick Carbon Steel Venturi Scrubber Price ($) vs. Gas Volume (1,000 acfm):

$$\$ = 8{,}547 + 489\ (Q) - 1.02\ (Q^2) \qquad (4.23)$$

(Equation reflects 1979 cost.)

Note: This equation is not applicable to installations greater than 200,000 acfm.

4.5.4 Mechanical Collectors

Equations 4.24 through 4.38 provide a means of estimating capital equipment costs for mechanical collectors (Equations 4.24 through 4.27) and associated components (Equations 4.28 through 4.38) as a function of the inlet area (IA) in ft^2. The mechanical collection system consists of the mechanical collectors or cyclones, supports, dust hopper, and scroll outlet. All costs are updated to October 1979 levels. Use Equations 4.24 through 4.27 for pricing 10 Guage (GA) and 3/16-in. carbon steel and stainless steel mechanical collectors.

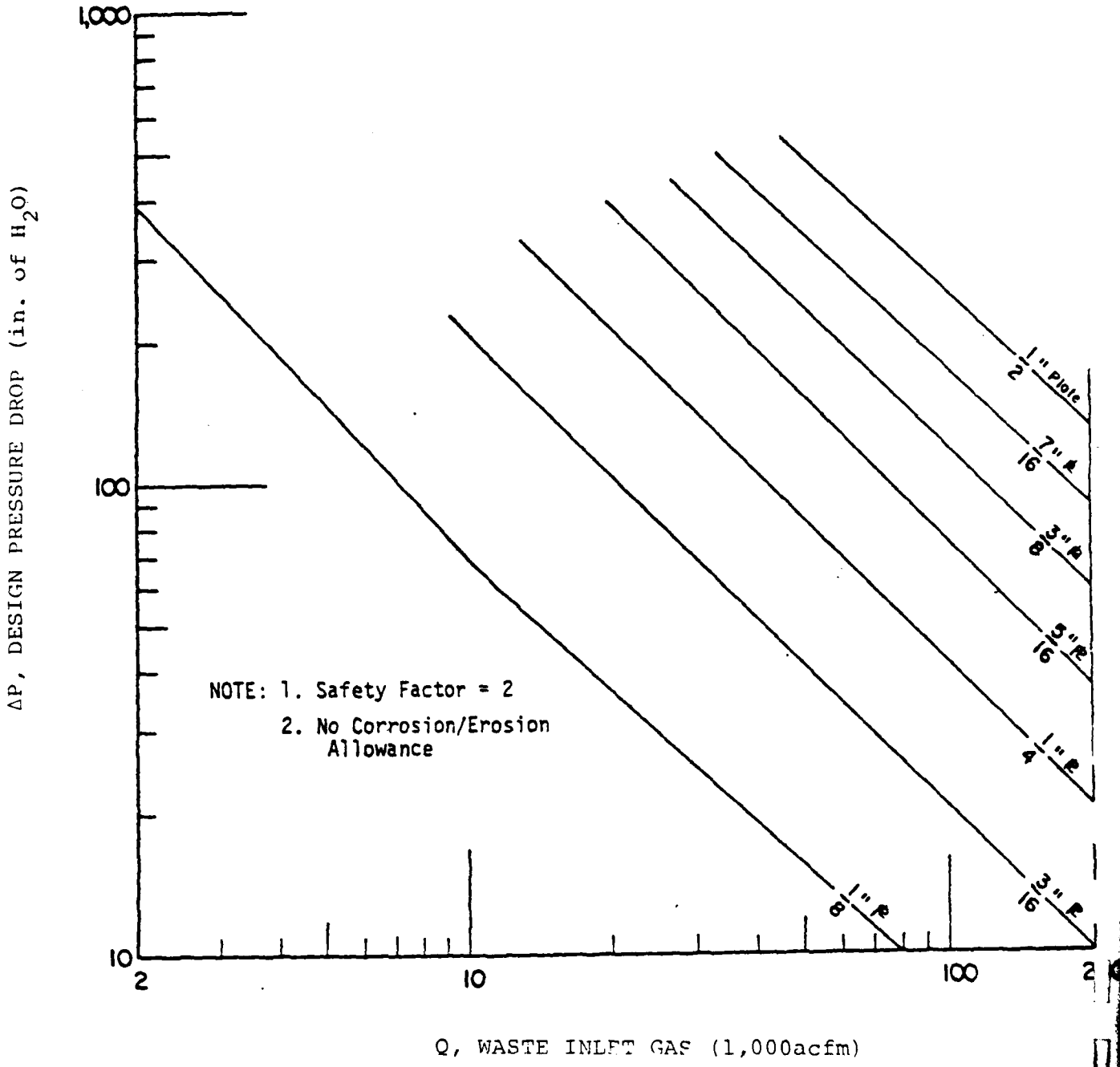


FIGURE 4-1

PRESSURE DROP VS. PLATE THICKNESS AND VOLUME[35]

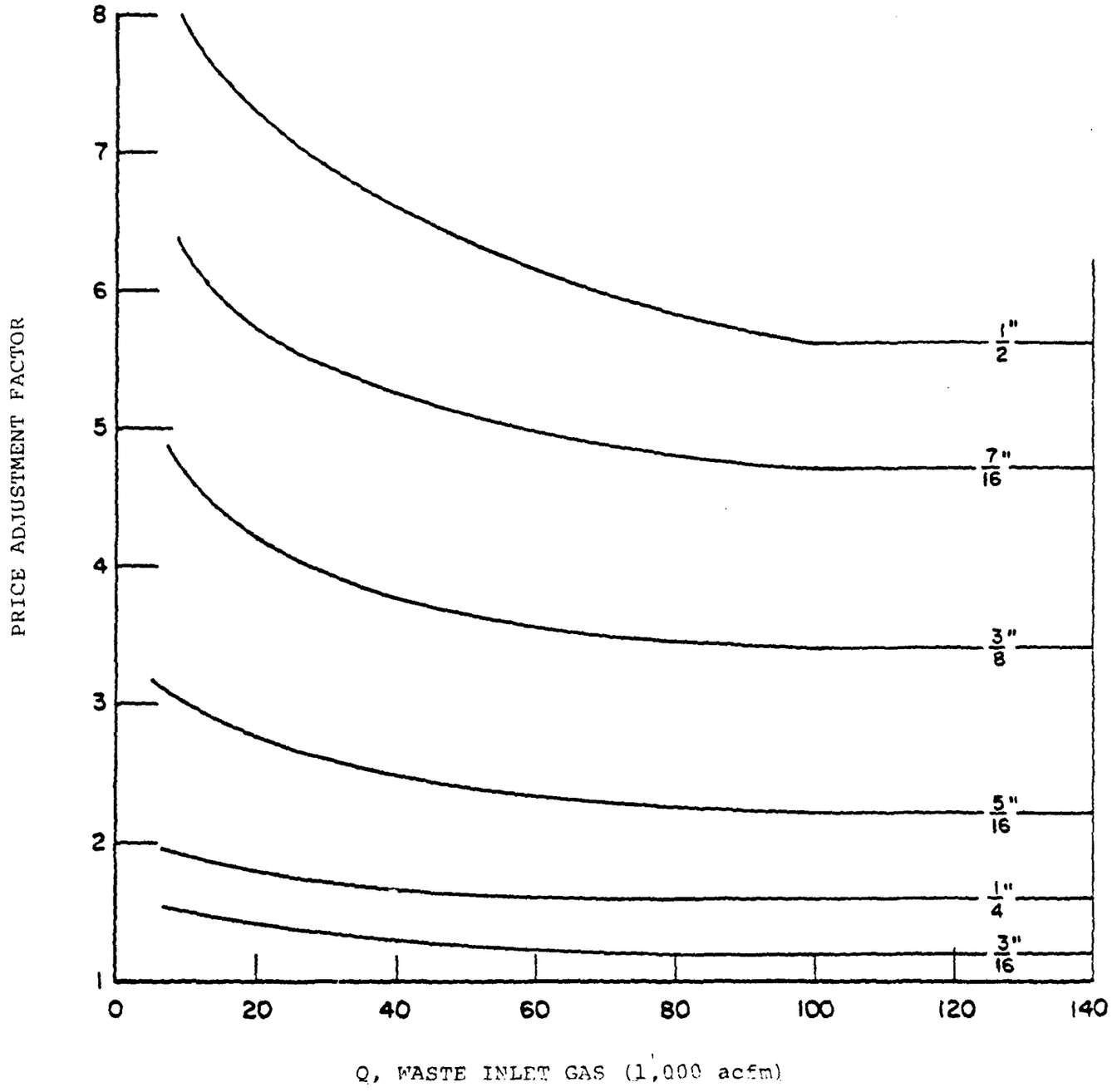


FIGURE 4-2

PRICE ADJUSTMENT FACTORS VS. PLATE THICKNESS AND VOLUME [35]

TABLE 4-4

PRICE ADJUSTMENTS FOR VENTURI SCRUBBERS

	Variation	Price Adjustment
A.	Metal Thickness (other than 1/8 in.)	See Figure 4-2
B.	316 Stainless Steel	3.5 x C.S. cost
C.	304 Stainless Steel	2.5 x C.S. cost
D.	3/16 in. Rubber-lined	Add $5.20/ft^2
E.	Fiberglass-lined	Add 15% of price for 1/8 in. carbon steel scrubber to total price
F.	Manual Variable Venturi	Add $4,138
G.	Automatic Variable Venturi	Add $7,640

Note: Costs shown reflect October 1979 prices.

3/16-in.C.S.	$ = 1,383 + 692 (IA)	(4.25)
10 GA S.S.	$ = 1,795 + 1320 (IA)	(4.26)
3/16-in. S.S.	$ = 2,080 + 2143 (IA)	(4.27)

Use Equations 4.28 through 4.30 for pricing collector supports vs. varying inlet area.

IA = 1-2 ft^2	$ = 690 + 180 (IA)	(4.28)
IA = 2-6 ft^2	$ = 1,240 + 170 (IA)	(4.29)
IA = >6 ft^2	$ = 2,345 + 145 (IA)	(4.30)

Dust hoppers may be constructed either from carbon or stainless steel and in various thicknesses. Equations 4.31 through 4.34 yield pricing information for 10 GA and 3/16-in. carbon and stainless steel dust hoppers.

10 GA C.S.	$ = 180 + 94 (IA)	(4.31)
3/16-in. C.S.	$ = 193 + 112 (IA)	(4.32)
10 GA S.S.	$ = 276 + 236 (IA)	(4.33)
3/16-in. S.S.	$ = 276 + 358 (IA)	(4.34)

Mechanical collector scroll outlets are also available in either carbon or stainless steel in various thicknesses. Equations 4.35 through 4.38 yield pricing information for 10 GA and 3/16-in. carbon and stainless steel scroll outlets.

10 GA C.S.	$ = 310 + 200 (IA) - 3.7 $(IA)^2$	(4.35)
3/16-in. C.S.	$ = 375 + 260 (IA) - 5.0 $(IA)^2$	(4.36)
10 GA S.S.	$ = 560 + 523 (IA) - 14.6 $(IA)^2$	(4.37)
3/16-in. S.S.	$ = 621 + 748 (IA) - 15.7 $(IA)^2$	(4.38)

4.5.5 Absorption Unit Costs--Packed and Plate Towers

From the view of mass transfer, the most efficient operation of an absorption column occurs at the load point, and columns are normally designed to operate at the velocity at which loading starts. However, occasionally, it may be more economic to operate at velocities well below loading. For example, if the packing and column constructions are inexpensive, but pressure drop is an important design consideration, it may be better to use larger cross sections and much lower gas velocities, even at the expense of extra packing height. On the other hand, if column construction and packing costs are high (e.g., for highly corrosive materials) and if pressure drop is relatively unimportant, it may be more economical to use a smaller cross-sectional area, coupled with some increase in height, with velocities up to 80 percent of flooding.

Unless the design is restricted by external factors, such as fan or compressor size available for handling a particular material, the column should be designed for an economic balance over the entire plant. Ideally, any absorption column should be designed so that the total annual costs, including depreciation, power costs, operating costs, maintenance, etc., are at a minimum.

Figure 4-3 illustrates the usual variation of annual cost with the gas rate. The value of the gas rate corresponding to the minimum value of the total annual cost is the economic rate. However, there is a range of values of gas velocity over which the total annual cost does not vary to a large extent, and in practice deviations of +25 percent from the optimum are generally of little importance.

The purchased cost for plate and packed columns may be divided into three groups:

(1) shell cost, including heads, skirts, manholes, and nozzles. Use Figure 4-4 and Table 4-5 to estimate these costs.[35]

(2) cost of internals, including trays, packing and accessories, packing supports, and distributor plates. Use Figures 4-5 through 4-8 and Tables 4-6 and 4-7 to estimate these costs.[35]

(3) auxiliary equipment cost, such as platforms, ladders, handrails, and insulation. These have not been included in this section. For exact costs, it is best to consult the particular manufacturer.

The cost figures and graphs presented here may be satisfactorily used for estimation purposes.

4.5.6 Adsorbers

Prices for carbon adsorbers are presented in Equations 4.39 to 4.41 as a function of the total number of pounds of carbon in the unit. The total (gross) number of pounds is determined by the adsorption and regeneration rates of carbon for the emission being controlled, as well as the flow rate of the adsorbate. To calculate the net pounds of carbon required for adsorption, first refer to Table 4-8 for a listing of carbon adsorption efficiencies for various solvents.[35] These efficiencies represent the ratio of pounds of solvent collected per 100 pounds of carbon, per hour, under conditions of 100°F and 200 cfm.

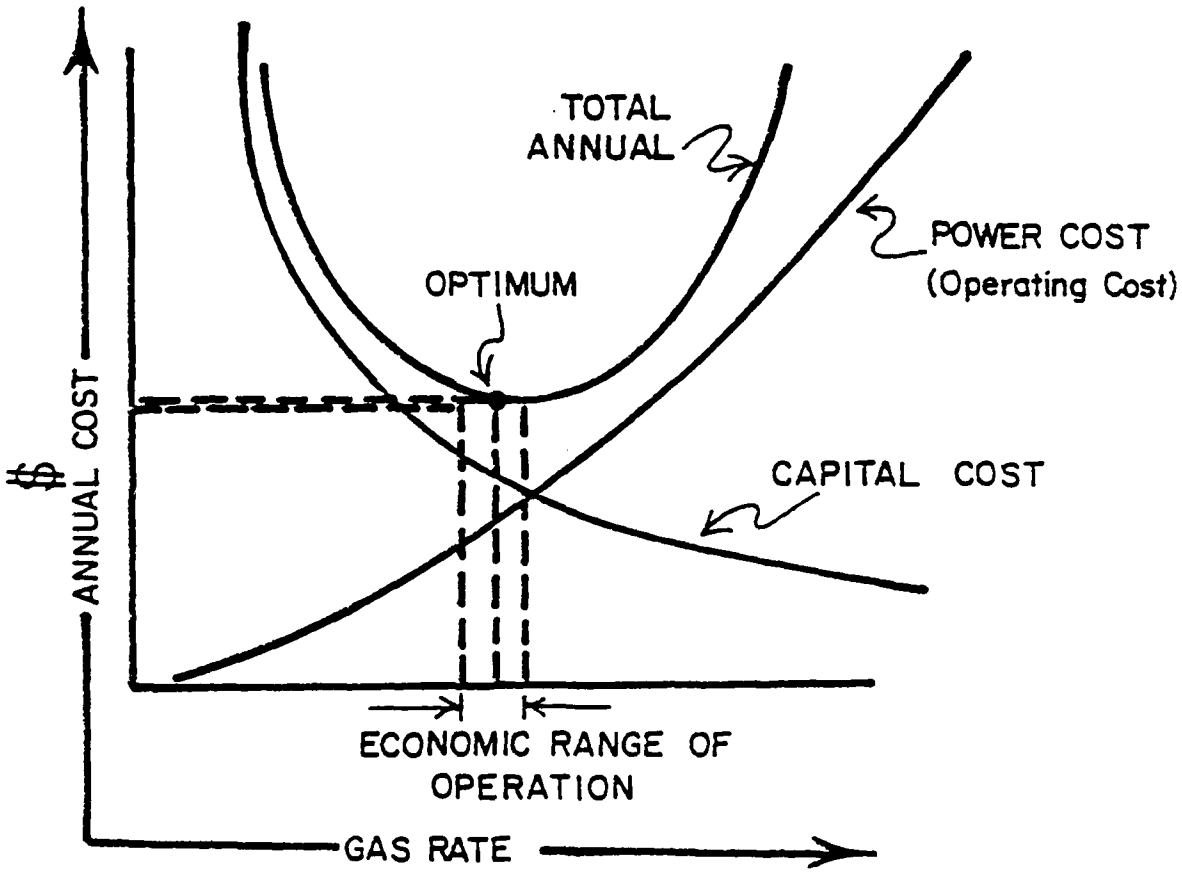
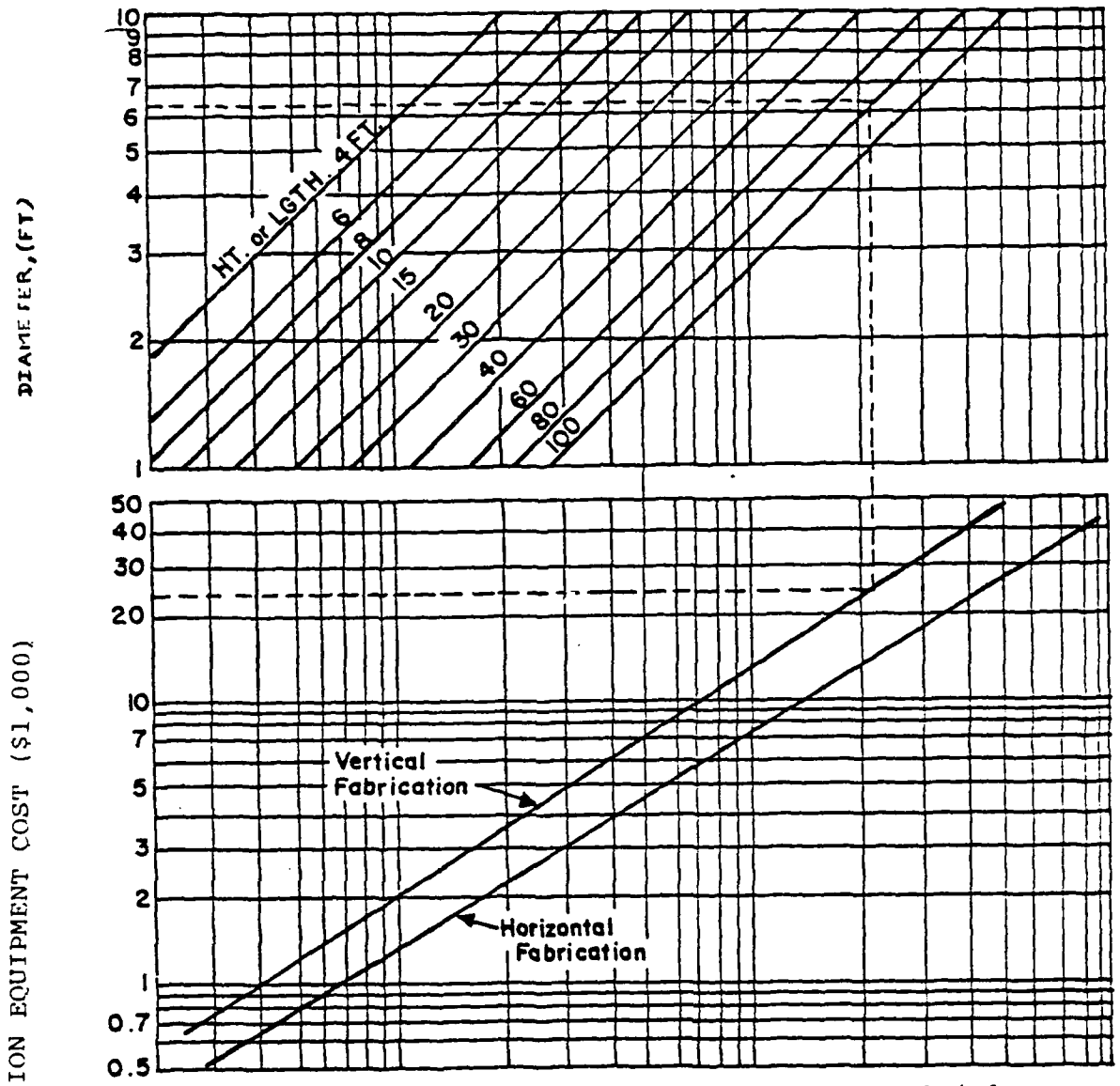


FIGURE 4-3

ANNUAL COST AS A FUNCTION OF GAS RATE



Note: Multiply costs by 2.42 to update to Oct. 1979

REQUIRED

Diameter, ft
Length, ft
Desigh pressure, psig
Shell material
Fabrication (horiz. or vert.)

Basis of chart
Carbon steel material
50-psi design pressure
Average nozzles and manways
ASME code construction
Shop fabrication

INCLUDED

- Vertical:
  - Shell and 2 heads
  - Nozzles and manways
  - Skirt, base ring and lugs
  - Tray supports
- Horizontal:
  - Shell and 2 heads
  - Nozzles and manways
- Saddles, 2

FIGURE 4-4

ABSORPTION COLUMN BASE-COST ESTIMATING CHART [35]

TABLE 4-5[35]

COST ESTIMATING ABSORPTION COLUMNS

COLUMN COST, S = (BASE COST) (F_m) (F_p) (INDEX)

Adjustment Factors:

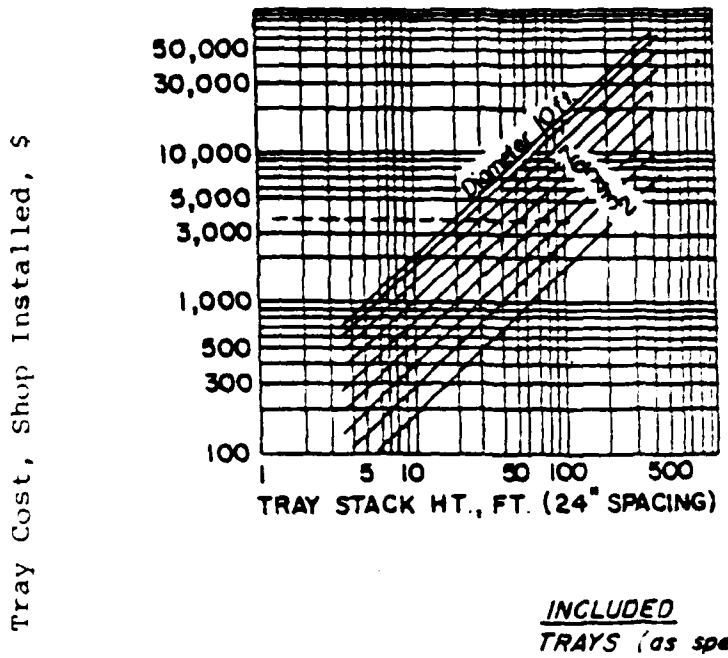
	F_m	F_m	Pressure Factor	
Material	Clad	Solid	psi	F_p
carbon steel	1.00	1.00	up to 50	1.00
stainless steel (316)	2.25	3.67	100	1.05
monel	3.89	6.34	200	1.15

Field Installation Module*	Vertical Fabrication	Horizontal Fabrication
Direct Cost Factor (Materials & Labor)**	3.04	2.29
Bare Module Factor***	4.34	3.29

*Module (process) represents direct cost of a process circuit. The cost includes equipment together with piping and instrumentation, minor steelwork (such as ladder platforms, supports, concrete foundations and sub-structures), insulation and paint.

**Direct Cost Factor relates f.o.b. equipment dollars to the cost of the equipment together with cost of fielr materials and field labor necessary to install the equipment on a prepared jobsite.

***Bare Module Factor includes all the direct and indirect cost elements in the process module.



Note: Multiply the costs by 2.42 to update to Oct. 1979

INCLUDED
TRAYS (as specified)
SUPPORTS, ALL FITTINGS
SHOP FABRICATION & INSTALLATION

FIGURE 4-5
TRAY COSTS [35]

TABLE 4-6

COST ESTIMATING--TRAYS FOR PLATE COLUMNS[35]

Tray Cost, $ = (Base Cost $(F_s + F_t + F_m)$ Index)

Adjustment Factors:

Tray Spacing (in.)	F_s	Tray Type	F_t	Tray Material	F_m
24	1.0	Valve	0.4	Carbon steel	0.0
18	1.4	Sieve	0.0	Stainless	1.7
12	2.2	Bubble Cap	1.8	Monel	8.9
		Kock Cascade	3.9		

TABLE 4-7

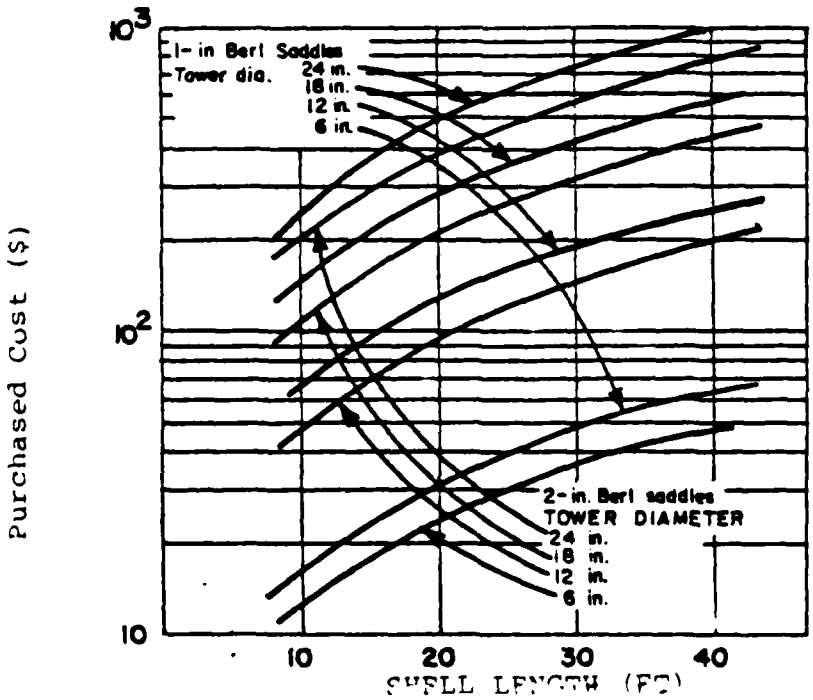
COST OF TYPICAL PACKING MATERIALS[35]

(Based on Oct. 1979 Costs)

Packing Cost, $ = (Materials & Labor, $/ft^3) (Packing Volume, ft^3) Index

Raschig Rings:	Size, (inches)			
	1	1½	2	3
	Materials and Labor ($/ft^3)			
Stoneware	12.58	10.4	8.5	7.0
Porcelain	16.94	14.0	11.4	9.4
Stainless	169.8	110.8	76.6	55

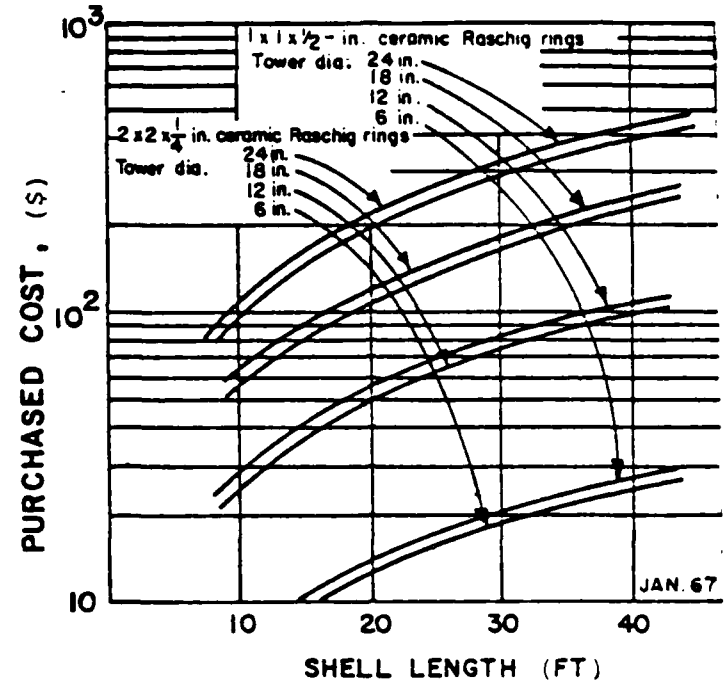
Berl Saddles:	Size, (inches)		
	3/4	1	1½
	Materials and Labor ($/ft^3)		
Stoneware	45.5	35.0	18.9
Porcelain	50.0	38.5	21.0



Note: Multiply the costs by 2.51 to update to Oct. 1979

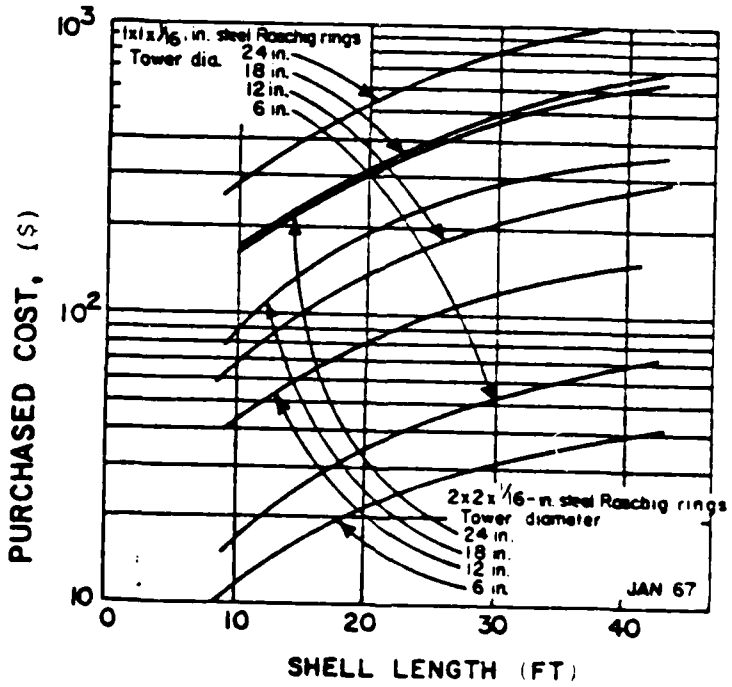
FIGURE 4-6

COST OF BERL-SADLLE COLUMN PACKINGS[35]



Note: Multiply th costs read off the graph by 2.51 to obtain prices for October 1979 values

FIGURE 4-7
COST OF CERAMIC RASCHIG-RING COLUMN PACKINGS[35]



Note: Multiply the costs read off the graph by 2.51 to obtain prices for Oct. 1979 values.

FIGURE 4-8
COST OF STEEL RASCHIG-RING COLUMN PACKING[35]

TABLE 4-8
LOWER EXPLOSIVE LIMITS (LEL) AND
CARBON ADSORPTION EFFICIENCIES FOR
COMMON POLLUTANTS[35]

Pollutant	Lower Explosive Limit (%) by volume in air)	Carbon Adsorption Efficiency (%)*
Acetone	2.15	8
Benzene	1.4	6
n-Butyl acetata	1.7	8
n-Butyl alcohol	1.7	8
Carbon tetrachloride	n	10
Chloroform	n	10
Cyclohexane	1.31	6
Ethyl acetate	2.2	8
Ethyl alcohol	3.3	8
Heptane	1	6
Hexane	1.3	6
Isobutyl alcohol	1.68	8
Isopropyl acetate	2.18	8
Isopropyl alcohol	2.5	8
Methyl acetate	4.1	7
Methyl alcohol	6.0	7
Methylene chloride	n	10
Methyl ethyl ketone	1.81	8
Methyl isobutyl ketone	1.4	7
Perchloroethylene	n	20
Toluene	1.27	7
Trichloroethylene	n	15
Trichloro trifluoroethane	n	8
VM & P Naptha	0.81	7
Xylene	1.0	10

*Efficiencies are based on 200 cfm of 100°F solvent-laden air, with no other impurities, * per hundred pounds of carbon per hour. Solvent recovery is 90-95%. Concentrations of solvent will alter efficiencies somewhat, but for estimation purposes those figures are satisfactory for 25 ppm and greater.

Note: n indicates no LEL.

Select the efficiency for the solvent to be controlled (for mixtures of solvents, see the reference noted below). Next, determine the rate of solvent emission in pounds per hour. For example, suppose a source produces 35 lb/hr of toluene; since the efficiency for toluene is 7 percent, 100 lb of carbon can adsorb 7 lb of toluene per hour. Therefore a total of

$$35 \text{ lb/hr} \times \frac{100 \text{ lb}}{7 \text{ lb}} = 500 \text{ lb of carbon per hour}$$

are required per hour of operation. To obtain the actual amount of carbon required, this must be multiplied by the number of hours of operation before regeneration.

Next, determine the steam regeneration rate for the solvent being collected, and calculate the number of beds and gross pounds of carbon required. If the regeneration rate (including coding) equals the collection rate, two beds will be required, and the gross weight of carbon must be twice the net weight. If the regeneration rate is one-half the collection rate, three beds will be needed, and the gross weight of carbon must be three times the net weight.

Equations 4.39 and 4.40 can be used for typical commercial applications such as dry cleaning plants and metal cleaning operations, as well as for industrial applications such as lithography and petrochemicals. Industrial requirements include heavier materials for high steam or vacuum pressure designs, and more elaborate safety controls against explosions and hydrocarbon breakthrough.

Prices for stationary bed carbon adsorber with steam regeneration vs. pounds of carbon (C) are listed in Equations 4.39 through 4.41.

A. Manual Operation

$$\$ = 138 + 7.37\ (C) \qquad (4.39)$$

B. Automatic Operation

$$\$ = 1{,}250 + 13.00\ (C) - 0.0005\ (C^2) \qquad (4.40)$$

Note: 1. Includes adsorber, carbon, blower or fan controls, steam regeneration equipment.
2. Equations reflect October 1979 costs.

Prices for custom adsorption unit vs. pounds of carbon are given by Equation 4.41.

C. Custom Adsorption Units

$$\$ = 79{,}123 + 6.84\ (C) \qquad (4.41)$$

Note: 1. Includes fans and pumps.
2. Equation reflects October 1979 costs.

4.5.7 Thermal Incinerators

4.5.7.1 Prices for Thermal Incinerators Without Heat Exchanger (1000 $) vs. Capacity, Q(1,000 acfm), and Residence Time, RT

A. RT = 0.2 sec.

$$\$ = 15.84 + 0.67\ (Q) - 0.0023\ (Q^2) \qquad (4.42)$$

B. RT = 0.5 sec.

$$\$ = 18.48 + 1.09\ (Q) - 0.01\ (Q^2) \qquad (4.43)$$

C. RT = 1.0 sec.

$$\$ = 21.88 + 1.52\ (Q) - 0.0083\ (Q^2) \qquad (4.44)$$

D. RT = 2.0 sec.

$$\$ = 28.87 + 2.39\ (Q) = 0.0017\ (Q^2) \qquad (4.45)$$

Note: 1. RT = Residence time for incineration.

2. Accuracy: ± 50%.

3. Price includes: incinerator, fan or blower, controls and instruments.

4. Price varies with retention time, materials of construction, and heat value of pollutant.

5. Equations reflect October 1979 costs.

4.5.7.2 Thermal Incinerators With Heat Exchangers Price (1,000 $) vs. Capacity, Q(1,000 acfm)

A. $$\$ = 31.46 + 2.70\ (Q) - 0.0035\ (Q^2) \qquad (4.46)$$

Note: 1. Based on 40 to 60 percent heat recovery.

2. Accuracy: ± 50%.

3. Price includes: incinerator, heat exchanger, fan or blower, damper controls and instruments.

4. Prices vary with:
   a. retention times
   b. materials of construction
   c. special controls
   d. heat content of pollutant.

5. Equation reflects October 1979 costs.

4.5.8 Catalytic Incinerators

Catalytic Incinerator Price (1,000 $) vs. Capacity, Q (1,000 acfm)

A. Package Units

$$\$ = 16.60 + 1.25\ (Q) \qquad (4.47)$$

B. Custom Units Without Heat Exchangers

$$\$ = 22.54 + 2.77\ (Q) - 0.014\ (Q^2) \qquad (4.48)$$

C. Custom Units With Heat Exchangers

$$\$ = 38.75 + 2.77\ (Q) \qquad (4.49)$$

4.5.9 Spray Chambers

Spray Chamber Devices (1000 $) vs. Inlet Gas Volume, Q (1,000 acfm)

A. $$\$ = 52.15 + 0.290\ (Q) \qquad (4.50)$$

Note:
1. Based on chamber velocity = 600 fpm.
2. Length/diameter = 3.0.
3. Carbon steel construction.
4. Does not include refractory.
5. Spray chamber costs include vessel and support rings, platform, spray system, and controls.
6. Equation reflects October 1979 costs.

4.5.10 Quenchers

Quencher costs (1,000 $) vs. Inlet Volume, Q (1,000 acfm)

A. $$\$ = 8.65 + 0.27\ (Q) \qquad (4.51)$$

Note:
1. Carbon steel construction.
2. Does not include refractory.
3. Does not include pumps and piping.
4. Quencher cost includes: vessel, spray nozzles and supports.
5. Equation reflects October 1979 costs.

4.5.11 Radiant Coolers

Prices for Fabricated 40-ft "U" Tube Heat Exchangers with Hoppers and Manifolds (1,000 $) vs. Number of Branches (X)

A. 12-in. Diameter Tube

$$\$ = -1.71 + 2.77\ (X) \qquad (4.52)$$

B. 18-in. Diameter Tube

$$\$ = -1.60 + 4.39\ (X) \qquad (4.53)$$

C. 36-in. Diameter Tube

$$\$ = -1.62 + 8.96\ (X) \qquad (4.54)$$

Note: 1. Prices include manifolds, hoppers, and tube.

2. Material sized for 10 GA tubes and 7 GA hoppers and manifolds.

3. For heights other than 40 ft see accompanying Table 4-9.

4. Prices reflect October 1979 costs.

4.5.12 Fans and Motors

There are three types of fans available for use:

1. backward-curved blade
2. radial tip
3. forward-curved blade.

Because dust buildup on blades can throw them off balance and cause excessive vibrations, fans with forward-curved blades should never be considered for applications where there is dust or particulate matter in the gas stream. The cost of this design has, therefore, not been included in this manual.

4.5.12.1 Backward-Curved Blade Fans

Fans with backward-curved blades are priced as a function of the actual air flow rate, pressure drop at standard conditions, and class, as given[35] in Figure 4-9. For example, if a Class III fan is to operate at sea level with gas temperature of 70°F and is to handle a gas volume of 20,000 cfm at 10 inches of water, the price would be $4,500.

However, in many cases a fan would not be operated at standard conditions, and cost adjustments must be made (use Table 4-10). For example, if actual conditions are:

a. gas temperature = 300°F
b. altitude = 1,000 ft
c. actual cfm = 5,000
d. actual P = 10-in. static pressure

then the fan is priced as follows:

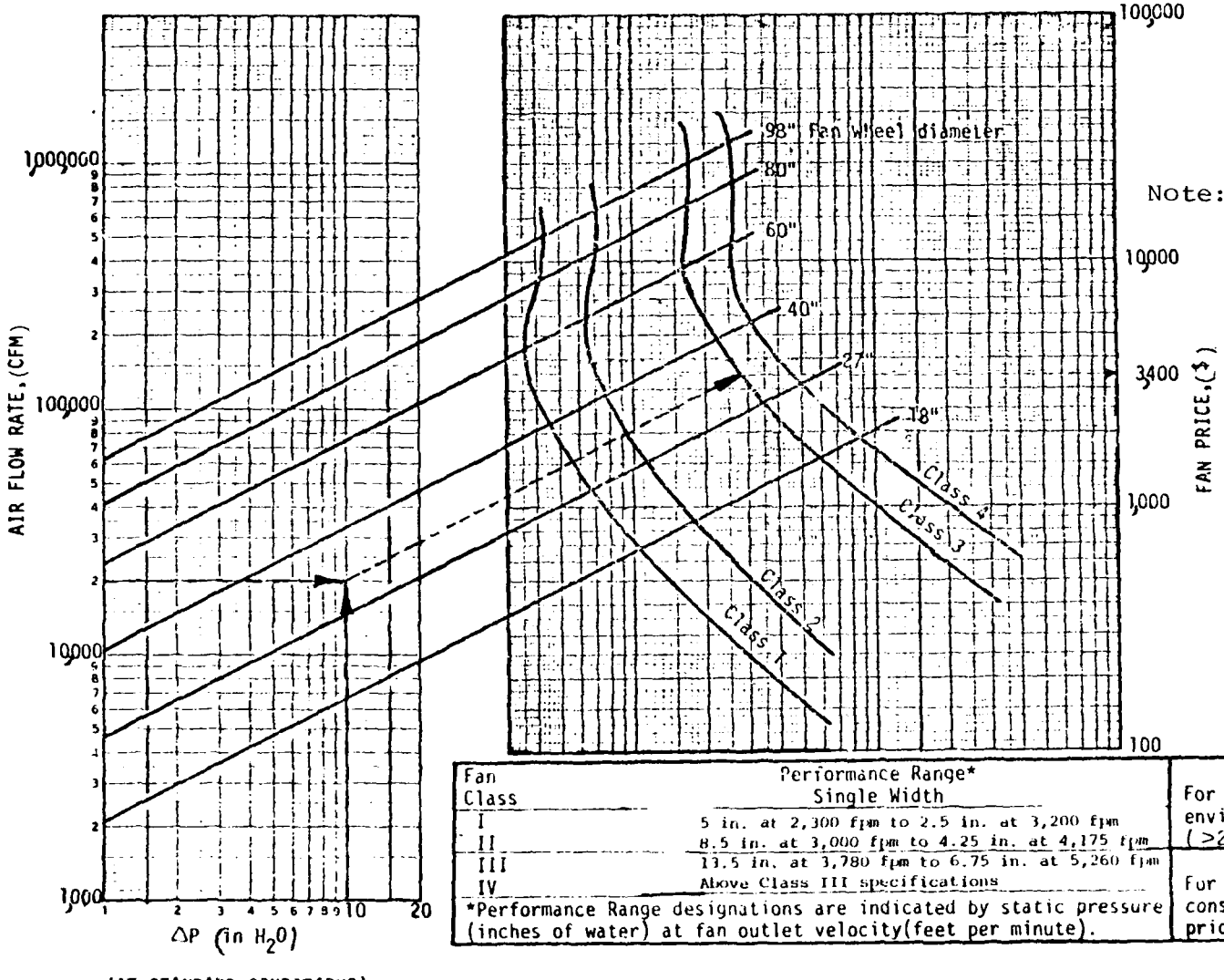
TABLE 4-9

PRICE ADJUSTMENTS FOR VARYING HEIGHTS FROM 40-ft UNIT

DEDUCT/ADD Per Branch ($).

Height (ft)	12 in.	18 in.	36 in.
10	-824	-1,278	-2,666
20	-549	- 855	-1,784
30	-275	- 422	- 877
40	0	0	0
50	+275	+ 422	+ 877
60	+549	+ 855	+1,784

DATA VALID FOR DECEMBER 1975



(AT STANDARD CONDITIONS)

Note: Multiply fan prices by 1.9 to obtain 1977 costs

Fan Class	Performance Range* Single Width	
I	5 in. at 2,300 fpm to 2.5 in. at 3,200 fpm	For high temperature environment add 3% (>250°F, < 600°F)
II	8.5 in. at 3,000 fpm to 4.25 in. at 4,175 fpm	
III	13.5 in. at 3,780 fpm to 6.75 in. at 5,260 fpm	
IV	Above Class III specifications	For stainless steel construction-multiply price by 2.5 .

*Performance Range designations are indicated by static pressure (inches of water) at fan outlet velocity(feet per minute).

FIGURE 4-9

PRICES OF FANS WITH BACKWARD-CURVED BLADES VS. CLASS, CFM, AND ΔP FOR ARRANGEMENT NO. 1[35]

1. Obtain fan sizing factor from Table 4-10 (300°F at 1,000 ft = 0.672).

2. Actual 10-in. static pressure/0.672 = 15 inches at standard conditions.

3. For a Class IV fan, at 50,000 cfm and 15 inches (see Figure 4-9), the price is $6,800. An additional 3 percent for high temperature applications yields an estimated price of $6,800 x 1.03 = $7,000.

4.5.12.2 Radial Tip Fans

The method of estimating prices for radial tip fans is the same as for backward-curved blade fans. Prices for radial tip fans operating under 20 inches static pressure are given[35] in Figure 4-10. Figure 4-11 provides the data for determining the fan RPM and motor brake horsepower for radial tip fans.[35] Refer to Figure 4-14 and Table 4-11 to obtain the motor and starter prices once the bhp has been determined.[35]

For radial tip fan applications involving greater than 20 inches static pressure, Figures 4-12 and 4-13 should be used to estimate the fan and motor prices, respectively. The static pressure must be converted to standard conditions as before, using Table 4-10.

4.5.12.3 Motors and Starters

The prices for the motor and the starter are obtained from Figure 4-14. Start with a gas flow rate and a static pressure at standard conditions of 20,000 cfm and 10 inches, respectively; find the point with those coordinates, and draw lines parallel to the fan rpm guidelines and the bhp guidelines. Read the fan rpm on the scale to the right, read the bhp on the scale to the left, then read the price for the type of starter needed and for the basic drip-proof motor at the selected rpm (a guide to determining motor rpm is given[35] in Table 4-12). Fan rpm is about 1,600 and the motor bhp is 44. According to Table 4-12, the motor rpm should be 1,800; hence, the corresponding price is about $600.

If a magnetic starter is selected, the price is about $350. Prices for motor types other than drip-proof may be estimated using Table 4-13. A totally enclosed motor for this example would cost $600 x 1.5 = $900. Use Table 4-14 for the selection of a motor type.

For conditions other than standard, the following steps must be taken to establish the motor and starter price. Again consider the 300°F application as before.

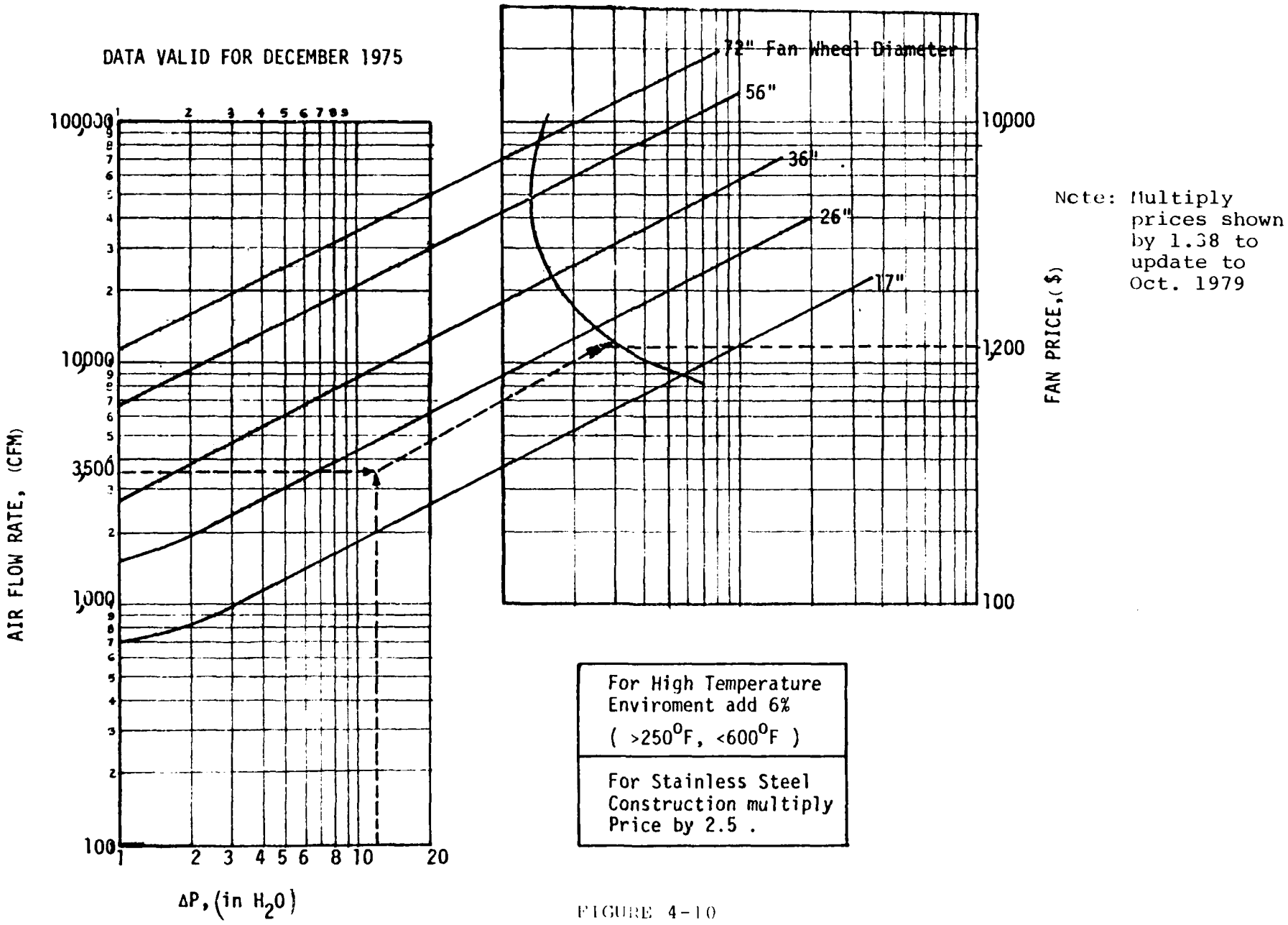


FIGURE 4-10

RADIAL TIP FAN PRICES VS. SCFM AND ΔP

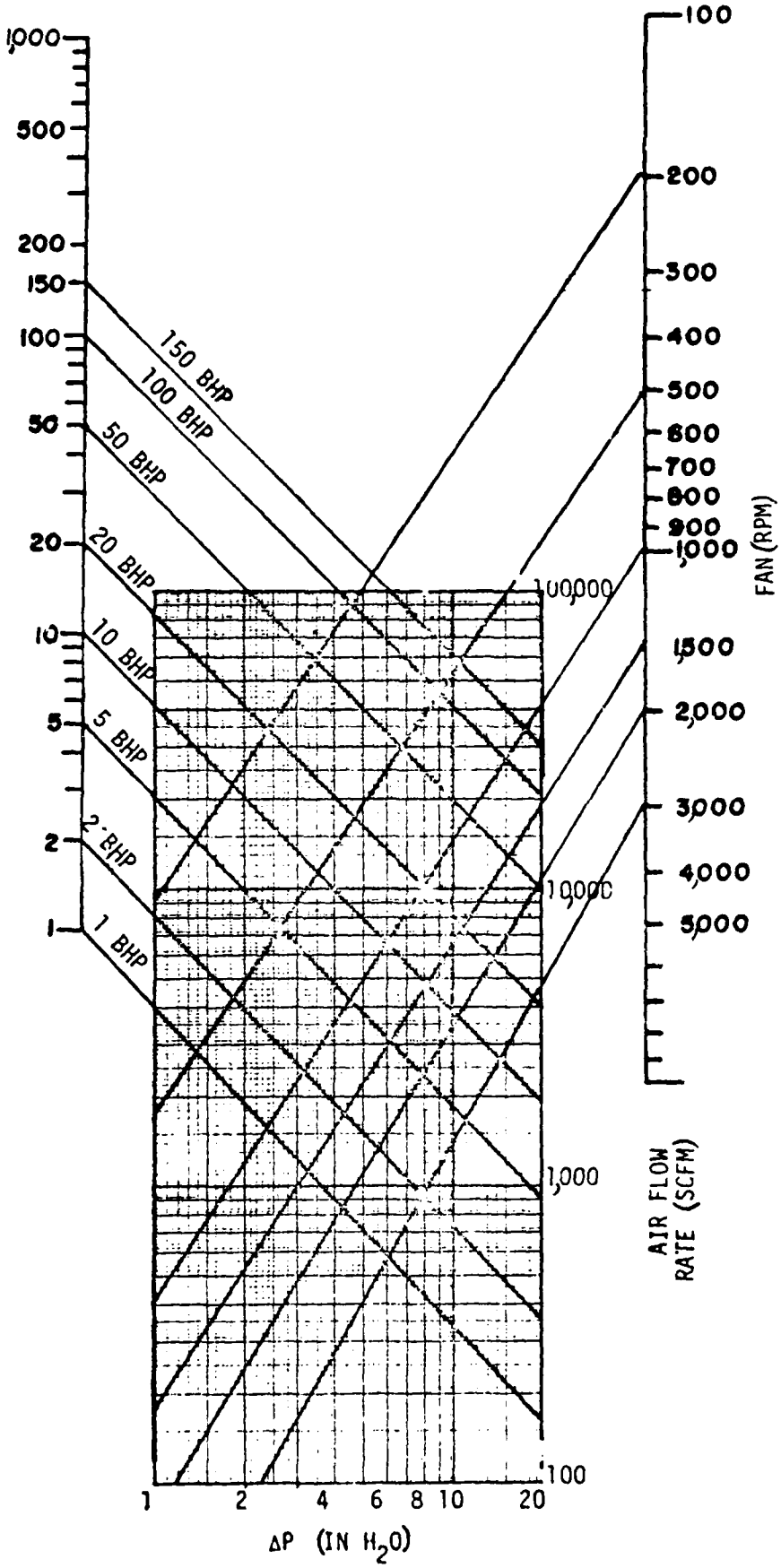


FIGURE 4-11
FAN RPM AND MOTOR BHP FOR RADIAL TIP FANS[35]

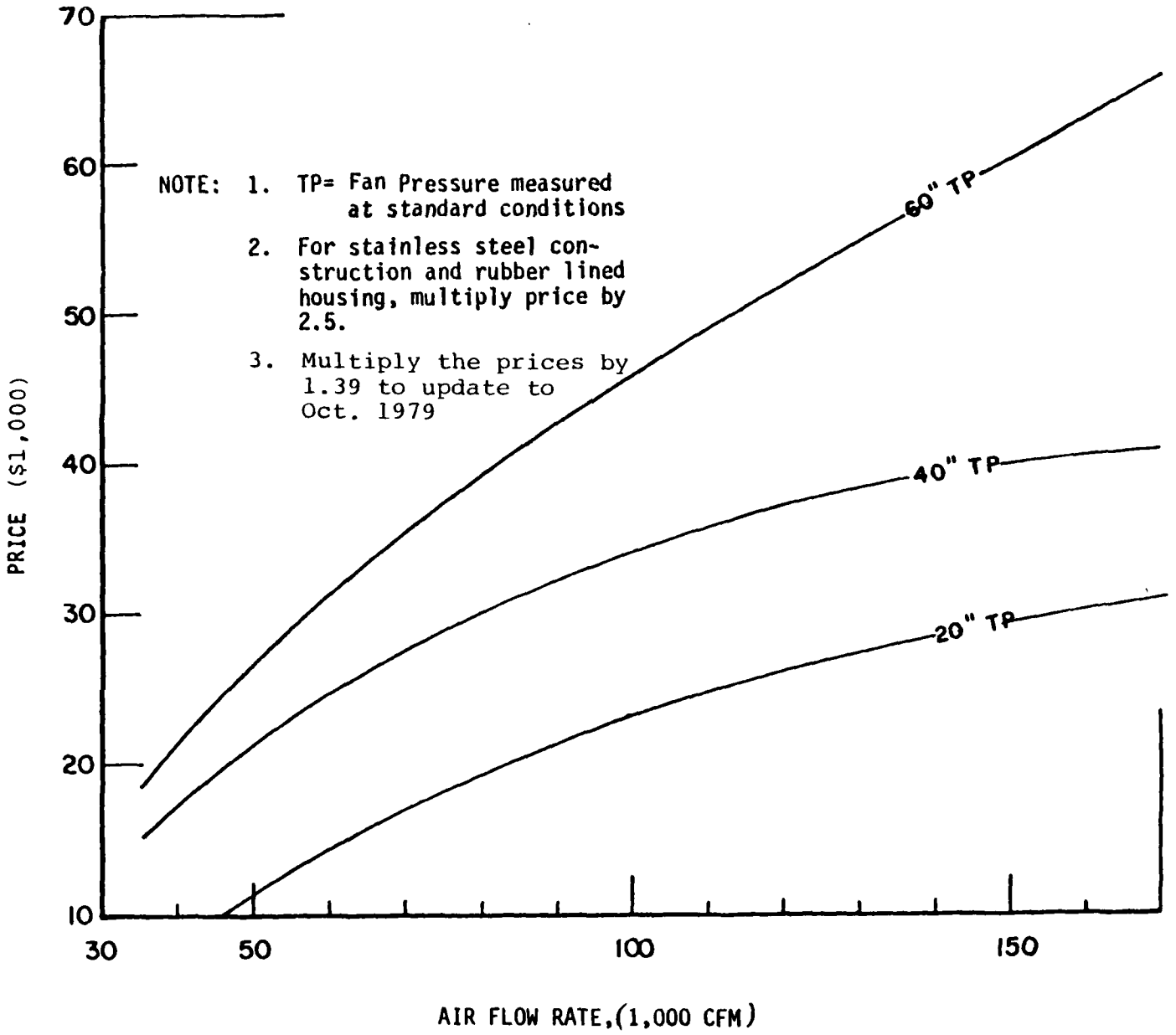


FIGURE 4-12

RADIAL TIP FAN PRICES[35]

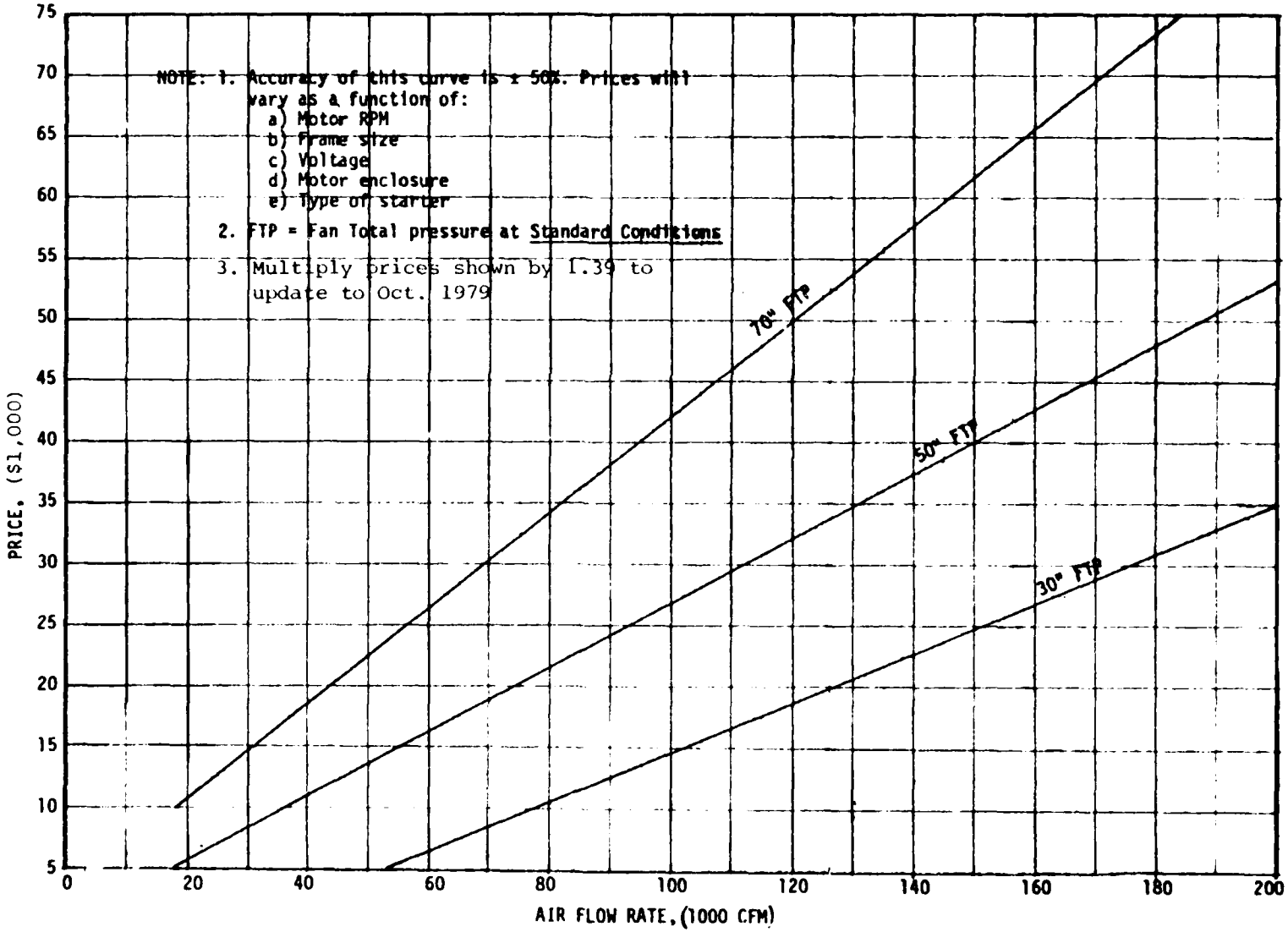


FIGURE 4-13

STARTER AND MOTOR PRICES FOR VENTURI SCRUBBER APPLICATIONS (High Pressure, High BHP)

TABLE 4-10

FAN SIZING FACTORS: AIR DENSITY RATIOS[35]

Unity Basis = Standard Air Density of 0.075 lb/ft^3

At Sea Level (29.92 inches Hg Barometric Pressure), This is equivalent to dry air at $70^{o}F$.

Air Temp of	Altitude in Feet Above Sea Level												
	0	1,000	2,000	3,000	4,000	5,000	6,000	7,000	8,000	9,000	10,000	15,000	20,000
	Barometric Pressure in Inches of Mercury												
	29.92	28.86	27.82	26.82	25.84	14.90	23.98	23.09	22.22	21.39	20.58	16.89	13.75
70	1.000	0.964	0.930	0.896	0.864	0.832	0.801	0.772	0.743	0.714	0.688	0.564	0.460
100	0.946	0.912	0.880	0.848	0.818	0.787	0.758	0.730	0.703	0.676	0.651	0.534	0.435
150	0.869	0.838	0.808	0.770	0.751	0.723	0.696	0.671	0.646	0.620	0.598	0.490	0.400
200	0.803	0.774	0.747	0.720	0.694	0.668	0.643	0.620	0.596	0.573	0.552	0.453	0.369
250	0.747	0.720	0.694	0.669	0.645	0.622	0.598	0.576	0.555	0.533	0.514	0.421	0.344
300	0.697	0.672	0.648	0.624	0.604	0.580	0.558	0.538	0.518	0.498	0.480	0.393	0.321
350	0.654	0.631	0.608	0.586	0.565	0.544	0.524	0.505	0.486	0.467	0.450	0.369	0.301
400	0.616	0.594	0.573	0.552	0.532	0.513	0.493	0.476	0.458	0.440	0.424	0.347	0.283
450	0.582	0.561	0.542	0.522	0.503	0.484	0.466	0.449	0.433	0.416	0.401	0.328	0.268
500	0.552	0.532	0.513	0.495	0.477	0.459	0.442	0.426	0.410	0.394	0.380	0.311	0.254
550	0.525	0.506	0.488	0.470	0.454	0.437	0.421	0.405	0.390	0.375	0.361	0.296	0.242
600	0.500	0.482	0.465	0.448	0.432	0.416	0.400	0.386	0.372	0.352	0.344	0.282	0.230
650	0.477	0.460	0.444	0.427	0.412	0.397	0.382	0.368	0.354	0.341	0.328	0.269	0.219
700	0.457	0.441	0.425	0.410	0.395	0.380	0.366	0.353	0.340	0.326	0.315	0.258	0.210

SOURCE: AMCA STANDARD #402-66
AIR MOVING AND CONDITIONING ASSOCIATION, INC.
205 West touhy Avenue
Park Ridge, Illinois 60068

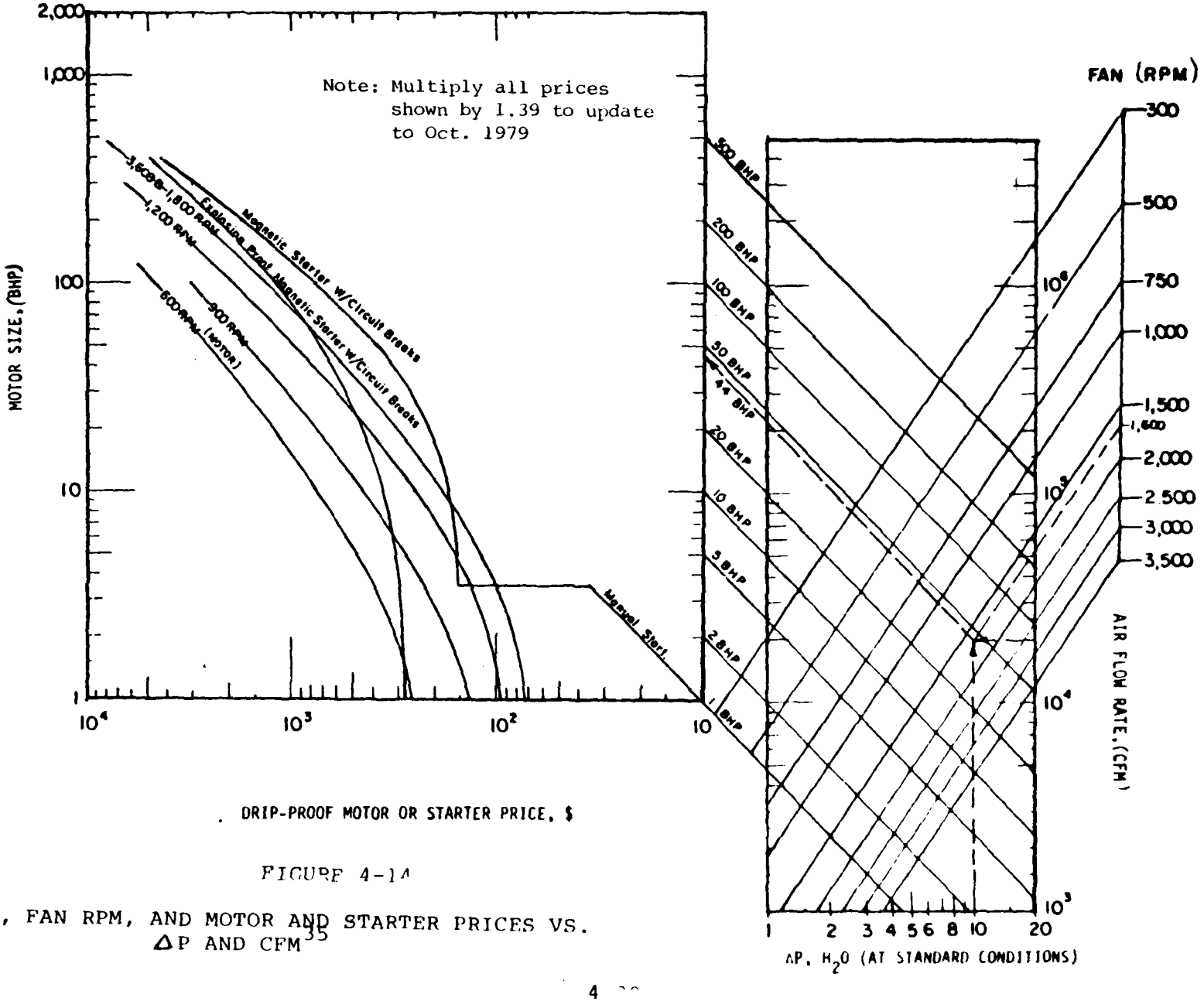


FIGURE 4-14

BHP, FAN RPM, AND MOTOR AND STARTER PRICES VS. ΔP AND CFM[35]

TABLE 4-11

MOTOR AND STARTER PRICE EQUATIONS[35]

rpm	Equation
3,600 1,800	$P = 60 + 11.9\ BHP + 0.00845\ BHP^2$
1,200	$P = 68 + 18.0\ BHP$
900	$P = 100 + 35.0\ BHP - -.07\ BHP^2$
600	$P = 204 + 52.6\ BHP - -.083\ BHP^2$
Mag. Starter	$P = 150 + 2.5\ BHP + 0.04\ BHP^2 - 0.00005\ BHP^3$
Explosion Proof Starter	$P = 270 + 8.5\ BHP + .008\ BHP^2$

TABLE 4-12 [35]

MOTOR RPM SELECTION GUIDE

Motor rpm	Fan rpm Ranges
3,600	2,400-4,000
1,800	1,400-2,400
1,200	1,000-1,400
900	700-1,000
600	700

TABLE 4-13 [35]

PRICING FACTORS FOR OTHER MOTOR TYPES

Horsepower	Totally Enclosed Fan Cooled	Explosion Proof
20	1.3	1.6
20	1.5	1.7

TABLE 4-14 [35]

MOTOR TYPE SELECTION

Drip-proof

In nonhazardous, reasonably clean surroundings free of any abrasive or conducting dust and chemical fumes. Moderate amounts of moisture or dust and falling particles or liquids can be tolerated.

Totally Enclosed Nonventilated or Fan Cooled

In nonhazardous atmospheres containing abrasive or conducting dusts, high concentrations of chemical or oil vapors, and/or where hosing down or severe splashing is encountered.

Totally Enclosed Explosion Proof

Use in hazardous atmospheres containing:

Class 1, Group D, acetone, acrylonitrile, alcohol, ammonia, benzene, benzol, butane ethylen dichloride, gasoline, hexane, lacquer solvent vapors, naptha, natural gas, propane, propylene styrene, vinyl acetate, vinyl chloride, or xylenes

Class II, Group G, flour, starc , or grain dust

Class II, Group F, carbon black coal, or coke dust

Class II, Group E, metal dust including magnesium and alumini or their commerical alloys.

1. Find the bhp from Figure 4-11: 50,000 cfm and 15 inches static pressure = 180 bhp

2. Correct the bhp by multiplying by the fan sizing factor: 180 bhp x 0.672 - 121 bhp, actual.

3. Find motor and starter prices at 121 bhp. The fan rpm does not require adjustment.

An inlet or outlet damper is usually required on fans, and prices for such are presented[35] in Figure 4-15. Note that the static pressure is measured for standard conditions.

V-belt drives may be selected for some applications. Figure 4-16 contains prices for V-belt drives as a function of motor bhp and fan rpm.[35] For direct drives, estimate price at 5 percent of the motor price.

4.5.13 Miscellaneous Equipment

4.5.13.1 Straight Duct

(i) Prices for Carbon Steel Straight Duct ($ per linear Foot) vs. Duct Diameter, D, (inches) and Plate Thickness

A. 1/8-in. Plate

$$\$/ft = -1.7 + 1.28\ (D) \qquad (4.55)$$

B. 3/16-in. Plate

$$\$/ft = -2.0 + 1.63\ (D) \qquad (4.56)$$

C. 1/4-in. Plate

$$\$/ft = -2.4 + 2.10\ (D) \qquad (4.57)$$

D. 3/8-in. Plate

$$\$/ft = -3.5 + 2.81\ (D) \qquad (4.58)$$

E. 1/2-in. Plate

$$\$/ft = -4.6 + 3.56\ (D) \qquad (4.59)$$

Note: 1. Estimates include flange every 40 ft.
2. Equations reflect October 1979 costs.

(ii) Prices for Stainless Steel Straight Duct ($ per linear foot) vs. Duct Diameter, D, (inches) and Plate Thickness

A. 1/8-in. Plate

$$\$/ft = -4.6 + 4.2\ (D) \qquad (4.60)$$

B. 3/16-in. Plate

$$\$/ft = -23.6 + 6.3\ (D) \qquad (4.61)$$

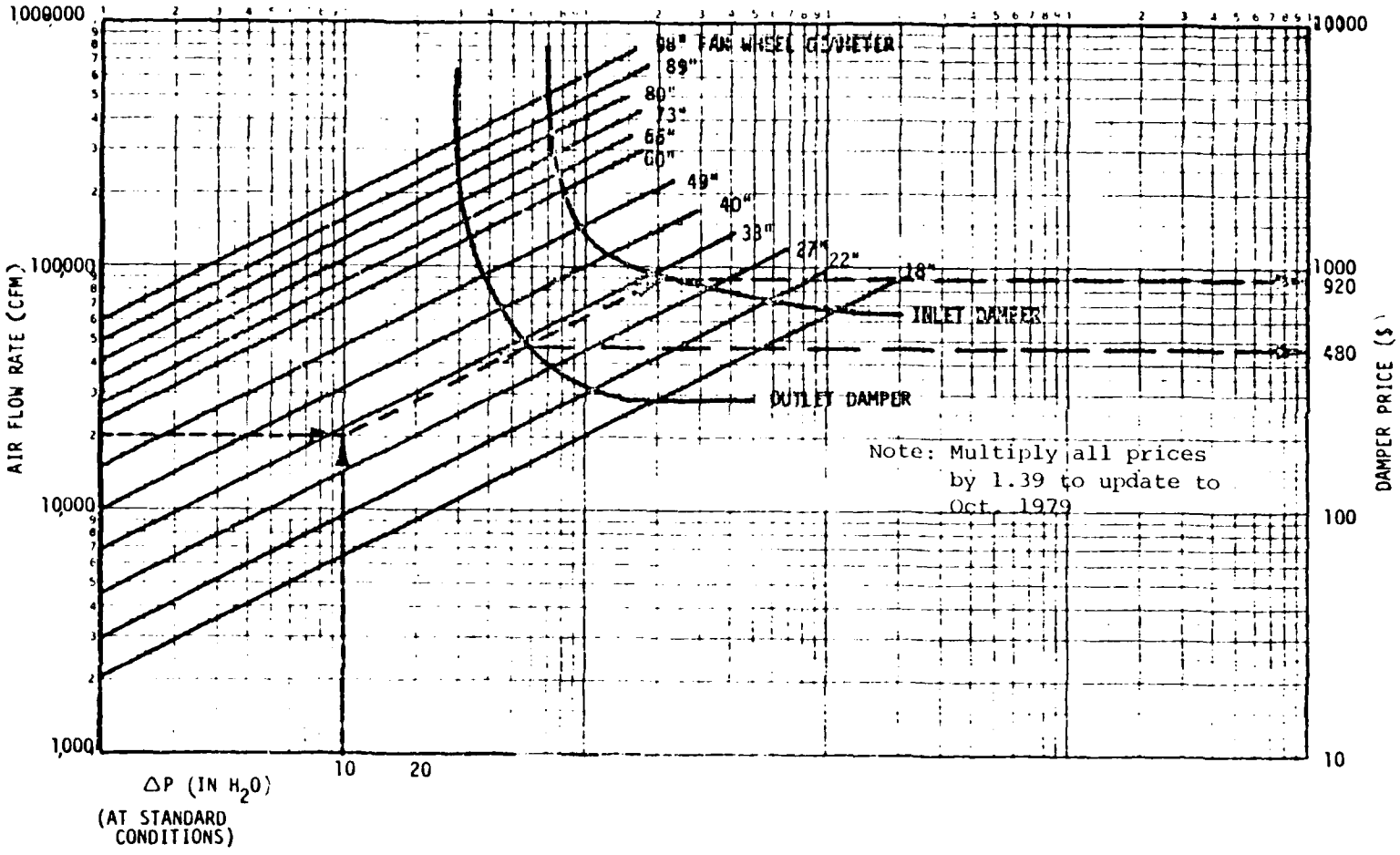


FIGURE 4-15

FAN INLET AND OUTLET DAMPER PRICES AS A FUNCTION OF CFM AND ΔP[35]

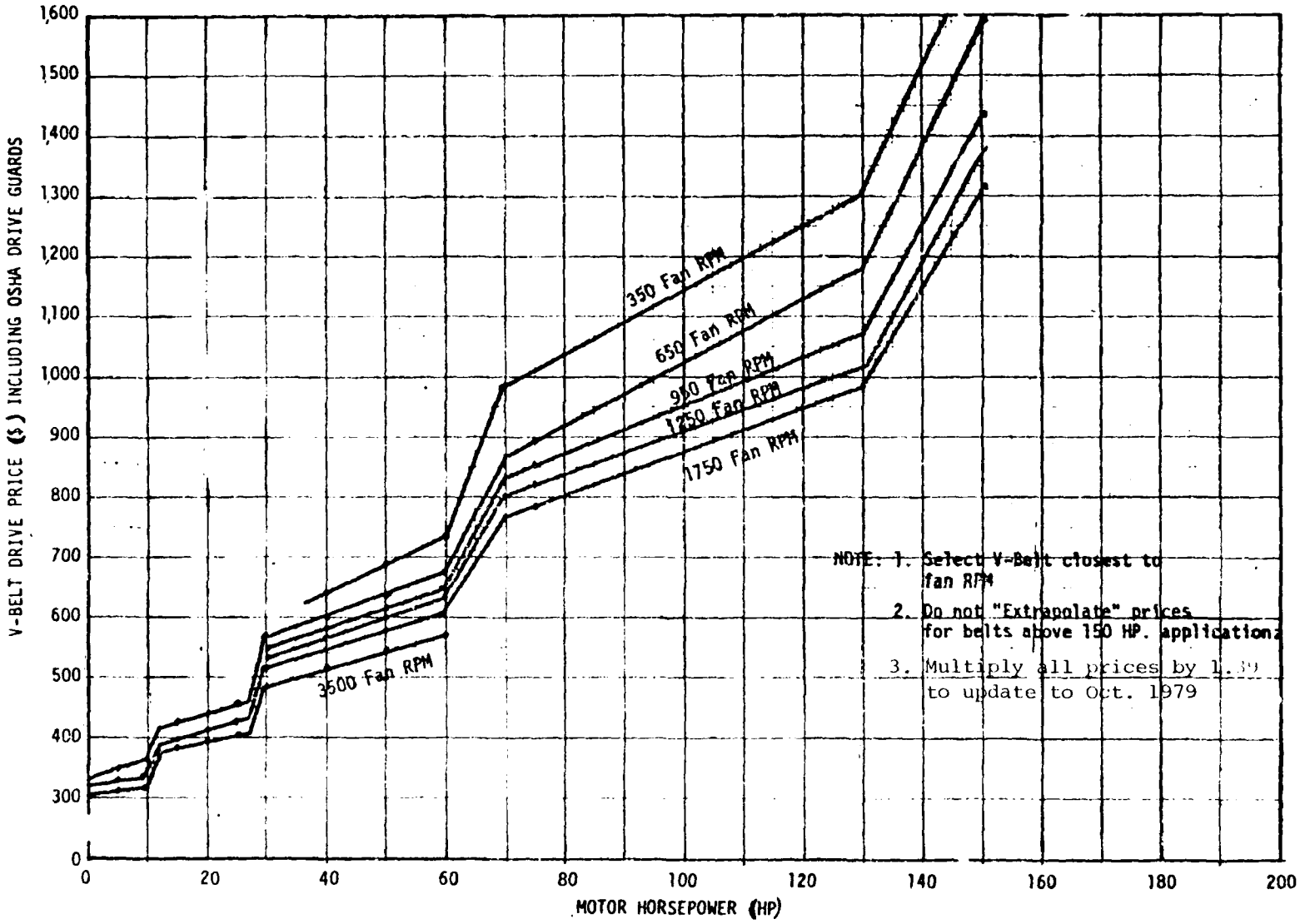


FIGURE 4-16

V-BELT DRIVE PRICES[35]

C. 1/4-in. Plate

$$\$/ft = -29.1 + 8.2\ (D) \qquad (4.62)$$

D. 3/8-in. Plate

$$\$/ft = -45.8 + 11.9\ (D) \qquad (4.63)$$

E. 1/2-in. Plate

$$\$/ft = -59.8 + 15.8\ (D) \qquad (4.64)$$

Note: 1. Estimate includes flange every 40 ft.
2. Equations reflect October 1979 costs.

4.5.13.2 Elbow Duct, Tees, and Transitions

Equations 4.65 through 4.69 and 4.70 through 4.74 contain cost information for carbon steel and stainless steel elbow duct, respectively. Prices are a function of duct diameter and material thickness.

For tees, the price will be one-third the corresponding price of an elbow having the same diameter, thickness, and material. For transitions, the price will be one-half the corresponding elbow price.

(i) Prices for Carbon Steel Elbow Duct ($) vs. Duct Diameter, D, (inches) and Plate Thickness

A. 1/8-in. Plate

$$\$ = 14.32\ (D) + 0.21\ (D^2) \qquad (4.65)$$

B. 3/16-in. Plate

$$\$ = 11.12 + 19.88\ (D) + 0.24\ (D^2) \qquad (4.66)$$

C. 1/4-in. Plate

$$\$ = -9.78 + 21.82\ (D) + 0.28\ (D^2) \qquad (4.67)$$

D. 3/8-in. Plate

$$\$ = -15.29 + 25.00\ (D) + 0.53\ (D^2) \qquad (4.68)$$

E. 1/2-in. Plate

$$\$ = -19.46 + 29.33\ (D) + 0.57\ (D^2) \qquad (4.69)$$

Note: 1. Estimate includes flanges.
2. Equations reflect October 1979 costs.

(ii) Prices for Stainless Steel Elbow Duct ($) vs. Duct Diameter, D, (inches) and Plate Thickness

A. 1/8-in. Plate

$$\$ = 21\ (D) + 0.76\ (D^2) \qquad (4.70)$$

B. 3/16-in. Plate

$$\$ = 135 + 19.18\ (D) + 1.14\ (D^2) \quad (4.71)$$

C. 1/4-in. Plate

$$\$ = 146 + 20.93\ (D) + 1.58\ (D^2) \quad (4.72)$$

D. 3/8-in. Plate

$$\$ = 217 + 25.85\ (D) + 2.30\ (D^2) \quad (4.73)$$

E. 1/2-in. Plate

$$\$ = 296 + 35.17\ (D) + 3.05\ (D^2) \quad (4.74)$$

Note: 1. Estimate includes flanges.
2. Equations reflect October 1979 costs.

4.5.13.3 Dust Removal Equipment

Prices for dust removal equipment ($) vs. length of conveyor, L, (ft) and diameter, heavy duty carbon steel construction.

A. 9-in. Diameter Screw

$$\$ = 575 + 57.1\ (L) \quad (4.75)$$

B. 12-in. Diameter Screw

$$\$ = 655 + 61.5\ (L) \quad (4.76)$$

4.5.13.4 Stacks

(i) Prices for Fabricated Carbon Steel Stacks ($) vs. Stack Height, H, (ft) and Diameter (inches) for 1/4 inch Plate

A. 24-in. Diameter

$$\$ = 1040 + 60\ (H) \quad (4.77)$$

B. 36-in. Diameter

$$\$ = 1040 + 79\ (H) \quad (4.78)$$

C. 48-in. Diameter

$$\$ = 1250 + 96\ (H) \quad (4.79)$$

D. 60-in. Diameter

$$\$ = 1525 + 106\ (H) \quad (4.80)$$

Note: 1. Plate thickness = 1/4 inch.
2. Includes: flange, stack, four cables, clamps, surface coating.
3. Good for stacks up to 100 ft.
4. Equations reflect October 1979 costs.

(ii) <u>Prices for Fabricated Carbon Steel Stacks ($) vs. Stack Height, H, (ft) and Diameter (inches) for 5/16 inch and 3/8 inch Plate Thickness</u>

A. 5/16-in. Plate

A.1 48-in. Diameter

$$\$ = 1{,}668 + 104\ (H) \tag{4.81}$$

A.2 60-in. Diameter

$$\$ = 2{,}016 + 120\ (H) \tag{4.82}$$

B. 3/8-in. Plate

B.1 54-in. Diameter

$$\$ = 2{,}224 + 124\ (H) \tag{4.83}$$

B.2 60-in. Diameter

$$\$ = 2{,}500 + 127\ (H) \tag{4.84}$$

Note: 1. Includes: flange, stack, four cables and surface coating.
2. Good for stacks up to 100 ft.
3. Equations reflect October 1979 costs.

4.6 ESTIMATING OPERATING AND MAINTENANCE COSTS

4.6.1 Particulate Pollutants Control Devices

The operating and maintenance costs for the control equipment for particulate pollutants are discussed here. The operating costs are based on the operating pressure drops and power costs for cyclones, fabric filters, and wet scrubbers. In the case of electrostatic precipitators, only the power cost for ionizing the gas and operating the fan is considered. The maintenance costs are based on the volumetric flow rate of the gas to be cleaned.

4.6.1.1 Cyclone Collectors

The operating cost for most cyclone collectors usually runs about \$0.015 to 0.030 per acfm per year. Power costs are typically rated on a pressure drop of about 0.25 horsepwer per inch of H_2O per 1,000 acfm. The theoretical annual cost of operation and maintenance for centrifugal collectors can be calculated as follows:

$$\$ = \text{acfm} \left[\frac{0.7457\ (\Delta P)\ (t) P_c}{6356\ \eta_f} + M \right] \qquad (4.85)$$

where

acfm = design capacity of collector, actual ft^3/min
ΔP = pressure drop, in. H_2O (typically 2 to 6 in. H_2O)
P_c = power cost, \$/kWh (typically \$0.03-0.04/kWh)
η_f = fan efficiency
M = maintenance cost, \$/acfm, (typically \$0.025-0.050/acfm)
t = annual operating time, hours.

The best cost data, however, are available directly from the manufacturer.

4.6.1.2 Electrostatic Precipitators

The only operating cost considered in the operation of electrostatic precipitators is the power cost for ionizing the gas and operating the fan. As the pressure drop across the equipment is usually less than 0.5 in. of water, the cost of operating the fan is, for all practical purposes, negligible. Operating cost is usually quite low, amounting to about \$0.03 to 0.05/year/acfm capacity. Maintenance cost ranges from \$0.02 to 0.03/year/acfm capacity.

The theoretical cost for operation and maintenance of electrostatic precipitation is:

$$\$ = acfm\ [(j)\ (t)\ (P_c) + M] \qquad (4.86)$$

where

acfm = design capacity of precipitator, actual ft^3/min
j = power requirements, kW/acfm
t = annual operating time, hour
M = maintenance cost, $/acfm
P_c = power costs, $/kWh.

However, the best cost information is available directly from the manufacturer.

4.6.1.3 Fabric Filter Systems

The theoretical operating and maintenance costs for fabric filter systems are based on the design capacity of the system, operating pressure drop of the system, fan efficiency, cost of power, and estimated maintenance charges. Annual maintenance costs may vary from $0.02 to 0.08/acfm and are typically $0.05/acfm. The overall efficiency for fan and motor is typically 0.55. Power costs at present, average about $0.025 to 0.050/kWh.

The theoretical annual operating and maintenance costs associated with fabric filters can be calculated from

$$\$ = acfm\ [1.173\ (10^{-4})\ (\Delta P)\ (tE/\eta_f) + M] \qquad (4.87)$$

where

acfm = design capacity of the fabric filter, actual ft^3/min
ΔP = pressure drop (avg.), in. of water
t = annual operating time, hours
E = power cost, $/kWh
η_f = fan efficiency
M = maintenance costs, $/acfm.

In comparison with other particulate control devices, fabric filters usually have the highest maintenance cost. On the other hand, they are capable of demonstrating the highest collection efficiencies. In general, only electrostatic precipitators require a greater capital investment.

4.6.1.4 Wet Scrubbers

Operating costs for wet scrubbers, especially venturi scrubbers, are relatively high, primarily due to the energy input required to collect the smaller sized particles. Operating costs typically range from $0.35 to

$1.00 per year per acfm, depending on circumstances and the specific type of scrubber. Maintenance costs are usually low, ranging from $0.02 to $0.06 per year per acfm capacity.

The theoretical annual cost of operation and maintenance for wet scrubbers can be expressed as follows:

$$\$ = S \left[0.7457\ HK \left(\frac{P}{6356E} + \frac{Qg}{1722F} + \frac{Qh}{3960F} \right) + WHL + M \right] \quad (4.88)$$

where

S = design capacity of the wet collector, acfm
0.7457 = a constant (1 horsepower = 0.7457 kilowatts)
H = annual operating time, hr
K = power costs, $/kWh
P = pressure drop across fan, in. of water
Q = liquor circulation, gallons per hour/acfm
g = liquor pressure at collector, psig
h = physical height liquor is pumped in circulation system, feet
W = make-up liquor consumption, gallons per hour/acfm
L = liquor cost, $/gallon
M = maintenance cost, $/acfm
E = fan efficiency, assumed as 0.60
F = pump efficiency, assumed as 0.50.

It is recommended that the manufacturer be directly consulted to obtain the best cost information on specific systems.

4.6.2 Gaseous Pollutants Control Devices

It is somewhat difficult to provide a detailed analysis and breakdown of overall operation and maintenance costs for combustion and adsorption units. Cost estimates are best obtained directly from the manufacturer. In general, operating costs for thermal reactors are primarily based on fuel requirements and will vary with operating temperature. These costs can be reduced substantially if heat recovery units are used in the system. Operating costs for catalytic reactors depend both on fuel requirements and catalyst life. The fuel consumption in a catalytic reactor is typically one-third that required for a corresponding thermal reactor. Generally, if the catalyst life is greater than 1.5 years, a catalytic reactor is used rather than a thermal reactor.

The most significant operating costs for adsorption systems are the activated carbon and regenerative steam.

Theoretical equations are given for carbon adsorbers, thermal incinerators and catalytic incinerators in Sections 4.6.2.1, 4.6.2.2, and 4.6.2.3, respectively. These equations have been obtained from empirical and operational experience and often change drastically due to advances in materials and technology, inflation, etc. They should be used only to roughly estimate operating and maintenance costs on a particular system, then verified for accuracy by vendor or manufacturer.

4.6.2.1 Carbon Adsorbers

Equations 4.89 to 4.92 give operating and maintenance costs for carbon adsorbers as a function of inlet gas volume and hydrocarbon concentration.

Carbon Adsorber Operating and Maintenance Costs ($ per hour) vs. Inlet Volume (log acfm) and Hydrocarbon Concentration (HG).

A. 100 ppm HC

$$\$/hr = 10^{(-2.06192 + 0.55732 (\log Q))} \times 1.39^{*} \quad (4.89)$$

B. 1,000 ppm HC

$$\$/hr = 10^{(-2.52906 + 0.72852 (\log Q))} \times 1.39^{*} \quad (4.90)$$

C. 1,600 ppm HC

$$\$/hr = 10^{(-2.80604 + 0.82787 (\log Q))} \times 1.39^{*} \quad (4.91)$$

D. 3,500 ppm HC

$$\$/hr = 10^{(-3.19176 + 0.96307 (\log Q))} \times 1.39^{*} \quad (4.92)$$

Note: 1. Equation based on concentration of toluene in ppm.
2. Costs include all labor and utility costs from collection point to stack.
3. Costs include adsorber replacement.
* October 1979 costs.

4.6.2.2 Thermal Incinerators

Equations 4.93 to 4.96 yield operating and maintenance costs for thermal incinerators with and without heat exchangers. These equations are a function of inlet gas volume and hydrocarbon concentration. The gas volume is measured before entering the heat exchanger for those units employing them.

Thermal Incinerator Operating and Maintenance Costs ($ per hour) vs. Inlet Volume (log acfm) and Hydrocarbon Concentration.

A. With Heat Exchangers

A.1 100 ppm

$$\$/hr = 10^{(-2.63171 + 0.94112 (\log Q))} \times 1.39^{*} \quad (4.93)$$

A.2 1,500 ppm

$$\$/hr = 10^{(-2.66782 + 0.88432 (\log Q))} \times 1.39^{*} \quad (4.94)$$

A.3 3,000 ppm

$$\$/hr = 10^{(-3.1507 + 0.95305 (\log Q))} \times 1.39^{*} \quad (4.95)$$

B. Without Heat Exchangers

B.1 100 ppm

$$\$/hr = 10^{(-2.2297 + 0.86591 (\log Q))} \times 1.39^{*} \quad (4.96)$$

Note: 1. Equations based on ppm concentration of hydrocarbons such as toluene, ketone, and naphtha.
2. Costs include all labor and utility costs from collection point to stack.
* October 1979 costs.

4.6.2.3 Catalytic Incinerators

Catalytic incinerator operating and maintenance costs (with and without heat exchangers) are given in equations 4.97 through 4.100 as a function of inlet volume (log acfm) and hydrocarbon concentration.

A. With Heat Exchangers

A.1 100 ppm

$$\$/hr = 10^{(-2.89536 + 0.91623 (\log Q))} \times 1.39^{*} \quad (4.97)$$

A.2 1,500 ppm

$$\$/hr = 10^{(-3.06752 + 0.91754 (\log Q))} \times 1.39^{*} \quad (4.98)$$

B. Without Heat Exchangers

B.1 100 ppm

$$\$/hr = 10^{(-2.61431 + 0.93898 (\log Q))} \times 1.39^{*} \quad (4.99)$$

B.2 1,500 ppm

$$\$/hr = 10^{(-2.62929 + 0.89972 (\log Q))} \times 1.39^{*} \quad (4.100)$$

Note: 1. Equations based on ppm concentration of hydrocarbon such as toluene, ketones, and naphthas.
2. Costs include all labor and utility costs from collection to stack.
3. Costs include catalyst replacement.
* October 1979 costs.

4.6.3 Maintenance and Installation Cost Factors and Equipment Life Guidelines

Table 4-15 gives[35] installation costs for the different types of control systems and maintenance costs for precipitators, scrubbers, and baghouses, expressed as a percent of purchased equipment cost. Equipment life is also given.

TABLE 4-15

MAINTENANCE AND INSTALLATION COST FACTORS AND EQUIPMENT LIFE GUIDLINES[35]

Maintenance	Low	Average	High	Very High
Electrostatic precipitators	1%	2%	4%	10%
Venturi scrubbers	8%	13%	18%	40%
Fabric filters	1%	2%	5%	7%

	Low	Average	High	Very High
Bag life	4 mo	1.5 yr	5 yr	10 yr

Installation	Low	Average	High	Very High
Electrostatic precipitators	50%	75%	120%	200%
Venturi scrubbers	70%	140%	220%	350%
Fabric filters	40%	75%	120%	170%
Incinerators (wo/HE)*	30%	50%	70%	90%
Incinerators (w/HE)	25%	45%	65%	90%
Adsorbers	30%	50%	70%	90%

Equipment Life	Short	Average	Long
Electrostatic precipitators	5 yr	20 yr	40 yr
Venturi scrubbers	5 yr	10 yr	20 yr
Fabric filters	5 yr	20 yr	40 yr
Thermal incinerators	5 yr	10 yr	20 yr
Catalytic incinerators	5 yr	10 yr	20 yr
Adsorbers	5 yr	10 yr	20 yr

*HE = heat exchangers

Note: Estimate maintenance and installation costs as percent of total equipment purchase price. Also note that a low installation percentage does not imply low maintenance or a short equipment life. These guidelines are estimates of the range of values that have been experienced in the industry. The choice of one over another depends on the application.

Uniform Annual Cost = Equivalent Annual Cost + Operating and Maintenance Cost = $70,368 + $48,960
= $119,328

4.7 LIFE CYCLE COSTING

For an adequate comparison of the costs of alternative air pollution control systems, a procedure is needed for combining the aggregate effects of first cost, operating cost, maintenance cost, and other costs or economic benefits that may arise from owning and operating the system. The procedure presented here is life-cycle cost analysis.

In order to determine the life-cycle cost, multiply the total capital cost for the control system, as calculated in Section 4.2, by the capital recovery factor as shown in Table 4-16. The years shown are the typical service life for the system used.

This equivalent annual cost of the asset added to the annual operating and maintenance costs yields the uniform annual cost for the selected pollution abatement system.

Extended discussion of cost analysis may be found in the book, Principles of Engineering Economy by Grant and Ireson, the Ronald Press Company, New York, 1970. The reader is encouraged to study this text, as there are many important topics and caveats that could not be covered in this brief space. Some special concerns include proper handling of equipment replacement, lease or buy decision, unequal equipment lives, determining the discount rate, and calculating utility costs. However, these subjects are principally the concern of the user of abatement equipment.

Example

A fixed-bed carbon adsorption system is designed to handle 20,000 cfm of air with a pollutant concentration of 1,000 ppm (hexane). Major items include a blower, two carbon beds, a condenser, decanter, interconnecting piping and valves, and automatic controls.

$$\text{ppm} \times \frac{\text{Molecular Weight}}{385.1 \times 10^6} \times 60\ \frac{\text{min}}{\text{hr}} \times Q\ \text{lb/hr hexane}$$

$$1{,}000\ \text{ppm} \times \frac{86}{385.1 \times 10^6} \times 60\ \frac{\text{min}}{\text{hr}} \times 20{,}000\ \text{cfm}$$

$$= 268\ \text{lb/hr hexane}$$

Referring to Table 4-8, the carbon adsorption efficiency of hexane is 6 percent.

TABLE 4-16
CAPITAL RECOVERY FACTORS [36]

Year	3 1/2	4 1/2	6	8	10	12	15	20	25	30
	Rate of Return or Interest Rate, Percent									
1	0.[illegible]2640	0.[illegible]3400	0.54544	0.56077	0.57619	0.59170	0.61512	0.65455	0.69444	0.734[illegible]
4	[illegible]	[illegible]2874	0.23859	0.30192	0.31547	0.32923	0.35027	0.38629	0.42344	0.461[illegible]
6	[illegible]	0.19388	0.20336	0.21632	0.22961	0.24323	0.26424	0.30071	0.33882	0.378[illegible]
8	[illegible]	0.15161	0.16104	0.17401	0.18744	0.20130	0.22285	0.26061	0.30040	0.341[illegible]
10	[illegible]	[illegible]	0.13587	0.14903	0.16275	0.17698	0.19925	0.23852	0.28007	0.3234
12	0.[illegible]348	[illegible]	0.11928	0.13270	0.14676	0.16144	0.18448	0.22526	0.26845	0.3134
14	[illegible]	[illegible]	0.10758	0.12130	0.13575	0.15087	0.17469	0.21689	0.26150	0.3078
16	[illegible]	0.08902	0.09895	0.11298	0.12782	0.14339	0.16795	0.21144	0.25724	0.304[illegible]
18	[illegible]	[illegible]	0.09236	0.10670	0.12193	0.13794	0.16319	0.20781	0.25459	0.3026
20	[illegible]	[illegible]	0.08718	0.10185	0.11746	0.13388	0.15976	0.20536	0.25292	0.301[illegible]
[illegible]	[illegible]	0.06744	0.07823	0.09368	0.11017	0.12750	0.15470	0.20212	0.25095	0.3004
30	[illegible]	0.06159	0.07265	0.08883	0.10608	0.12414	0.15230	0.20085	0.25031	0.3001
35	[illegible]	0.05727	0.06897	0.08530	0.10369	0.12232	0.15113	0.20034	0.25010	0.3000
40	[illegible]	0.05134	0.06646	0.08386	0.10226	0.12130	0.15056	0.20014	0.25006	0.3000

Following the procedure outlined in Section 4.5.6,

Solvent emissions $\frac{(lb)}{hr} \times \frac{100\ lb}{6\ lb}$ = lb carbon required per bed

$$268 \times \frac{100}{6} = 4{,}466 \text{ lb. carbon per bed}$$

Using 4,500 lb carbon per bed
total carbon required = 9,000 lb.

Using Equation 4.41 for custom adsorber equipment cost

$$\$ = 79{,}123 + 6.84\ (C)$$
$$\$ = 79{,}123 + 6.84\ (9{,}000)$$
$$\$ = 141{,}000$$

Next, calculate the total capital cost of the system.

Using Section 4.2,

(1)	Estimated equipment costs	$141,000
(2)	Tax and freight at 7% of (1)	9,870
(3)	Installation costs (use 100% of equipment costs for this example	141,000
(4)	Subtotal (1) + (2) + (3)	$291,870
(5)	Engineering at 10% of (4)	29,187
(6)	Subtotal (4) + (5)	321,057
(7)	Contingencies at 10% of (6)	32,106
(8)	Total estimated capital costs	$353,163

Using Table 4-15, the service life of adsorbers is an average of 10 years.

Assuming that the current interest rate is 15 percent,

Using Table 4-16, the capital recovery factory is 0.19925
Equivalent annual cost of equipment = 353,163 x 0.19925
= $70,368

Using Equation 4.86, calculate the yearly operating and maintenance costs.

$$\$/hr = 10^{(-2.52906 + 0.72852\ (\log Q))} \times 1.39$$

$$\$/hr = 10^{(-2.52906 + 0.72852\ (\log 20{,}000))} \times 1.39$$

$$\$/hr = \$5.59$$

ssuming constant operation,

Operating and Maintenance Costs = $5.49 x 8760
= $48,960

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CHAPTER 5

MATERIALS OF CONSTRUCTION

5.1 INTRODUCTION

In the selection of materials of construction for a particular system, it is important first to take into consideration the characteristics of the system, giving special attention to extraneous factors that may influence corrosion. Since these factors would be peculiar to a particular system, it is impractical here to attempt to offer a set of rules that would cover all situations.

The materials from which the system is to be fabricated are the second important consideration; therefore, knowledge of the characteristics and general behavior of materials when exposed to certain environments is essential.

In the absence of actual corrosion information for a particular set of conditions, a reasonably good selection would be possible from data based on resistance of materials to a very similar environment. These data, however, should be used with some reservations. Good practice calls for applying such data for preliminary screening; materials selected thereby would require further study in the system under consideration.

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5.2 MATERIAL SELECTION

Numerous construction materials are available, both metallic and nonmetallic. Physical properties should be thoughtfully examined before any final selection is made. Table 5.1 offers a general guide to proper selection. It gives corrosion data on many materials that may be encountered in pollution control service.

Selection of the proper materials of construction for a particular application depends on a number of factors. The major ones are:

- Strength requirements
- Temperature
- Corrosiveness
- Scaling potential
- Erosion potential

5.2.1 Strength Requirements

As the size of the individual unit increases, the structural strength requirements also increase. These factors become quite important in weighing the use of fiberglass reinforced polyester and light gauge exotic metals against steel lined with suitable material. The economics of linings compare more favorably as the strength requirement increases.

5.2.2 Temperature Limitations

The temperature limitations of materials of construction[30] are given in Table 5-2.

5.2.3 Materials and Corrosion

Since one single pollution source may produce a variety of corrosive gases, the selection of materials is difficult. Types of corrosion include:

(1) Stress corrosion. This results from a combination of tensile stress and a corrosive material, often a chloride.

(2) Intergranular corrosion. It attacks grain boundaries in stainless steels.

(3) General corrosion. This is the uniform dissolution of metal over its entire surface exposed to a corrosive material.

(4) Pitting. This is a preferential accelerated corrosive attack on a few points such as crevices.

Corrosion accelerates with increased temperature. It occurs not only at lower pH, but can also occur at neutral pH where corrosive salts are present.

5.2.4 Scaling

Some materials resist scaling better than others. Linings such as plastic reinforced with flake glass, e.g., Flakeline 103, have a relatively rough surface that will accelerate scaling. The smoother materials will be more desirable in areas subject to scaling. Where corrosion rates may be acceptable from a structural standpoint, they may not be acceptable because of the rough surface created. This is one reason stainless steel rather than carbon steel is used in moisture eliminators in atmospheres that are not very corrosive.

5.2.5 Erosion

Materials of construction must be carefully examined in light of the erosive nature of continuous flow processes. Venturi scrubbers, for example, cause high abrasion potential because of the high velocity in and downstream of the throat section. Rubber-lined steel and brick lining are often necessary to provide the required abrasion resistance. Where corrosion is not a problem, abrasion liners of steel or rubber belting can be placed in the high-velocity areas. Where corrosion and erosion are both present, the problem deteriorates further. Many materials, such as stainless steel, and corten, rely on a surface film to provide corrosion resistance. This passive film is a thin amorphous layer of metal and oxygen, which clings to the substrate and keeps out the water that will dissolve the iron. Continual abrading of this coating will cause accelerated corrosion.

In addition to high velocity, the size, quantity, and physical properties of particles are also of importance. Hard, sharp particles will cause accelerated erosion. In general, however, abrasion has not been as difficult a problem as corrosion or scaling.

TABLE 5-1

DETAILED CORROSION DATA ON CONSTRUCTION MATERIALS[30]

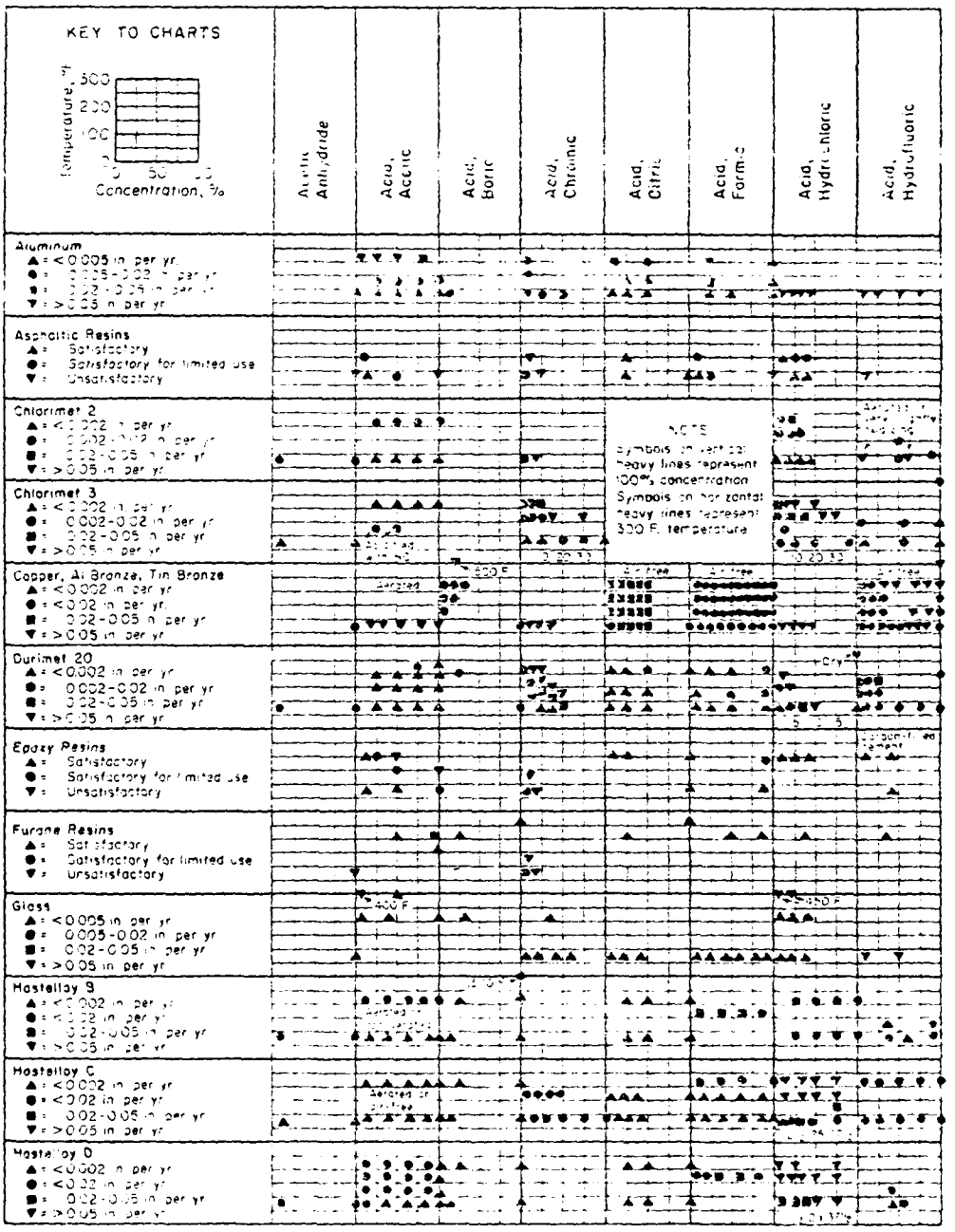
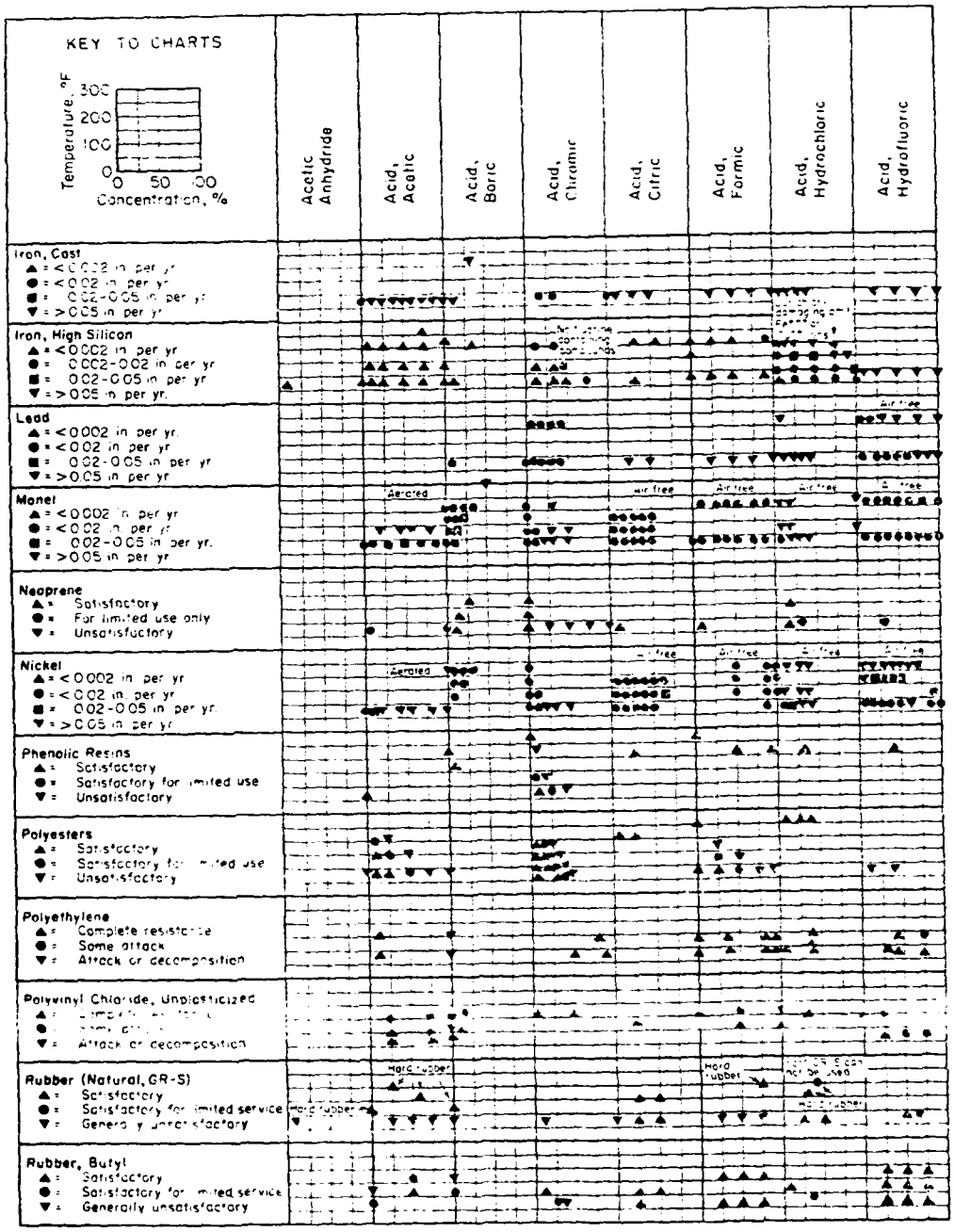


TABLE 5-1 (Cont.)



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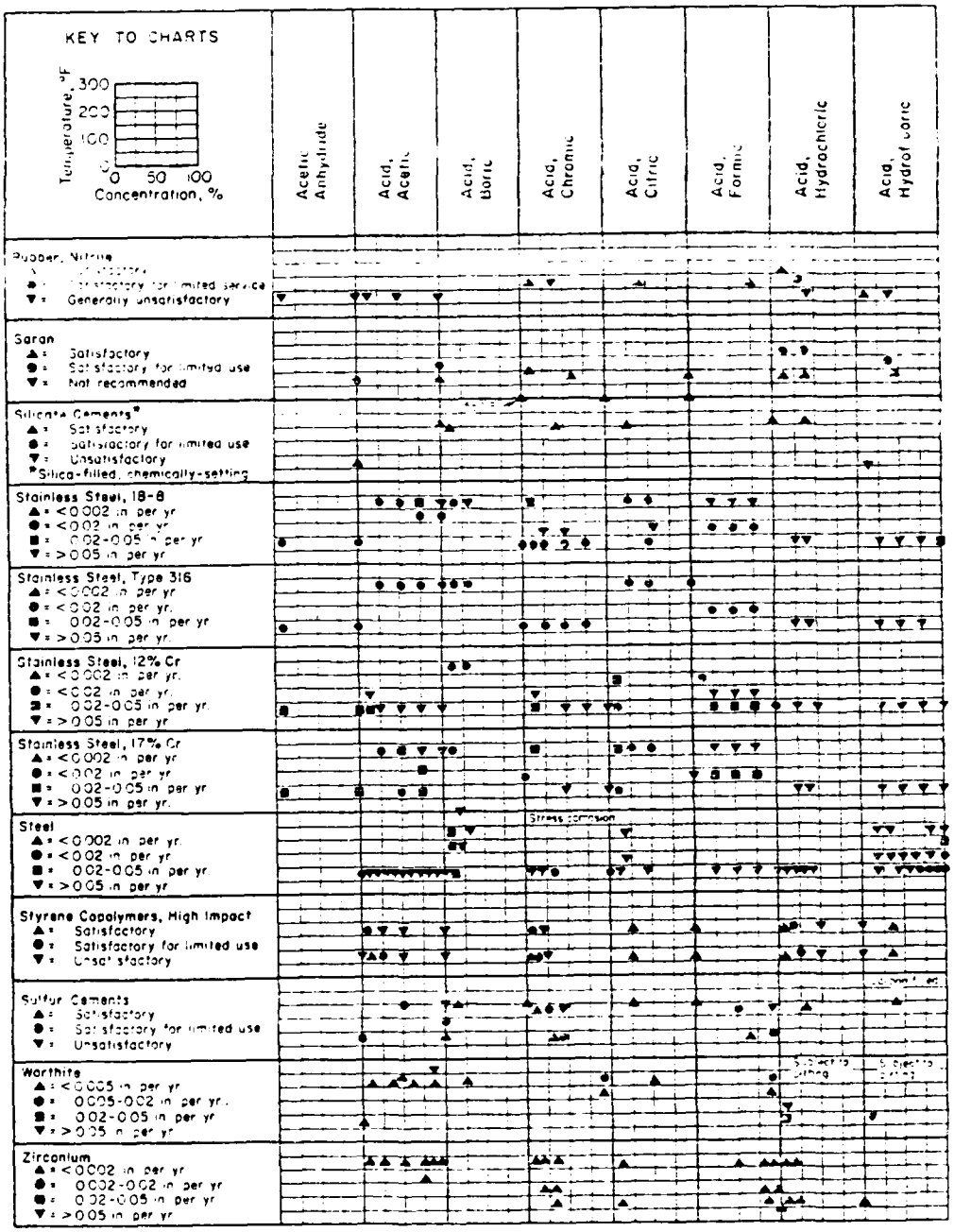

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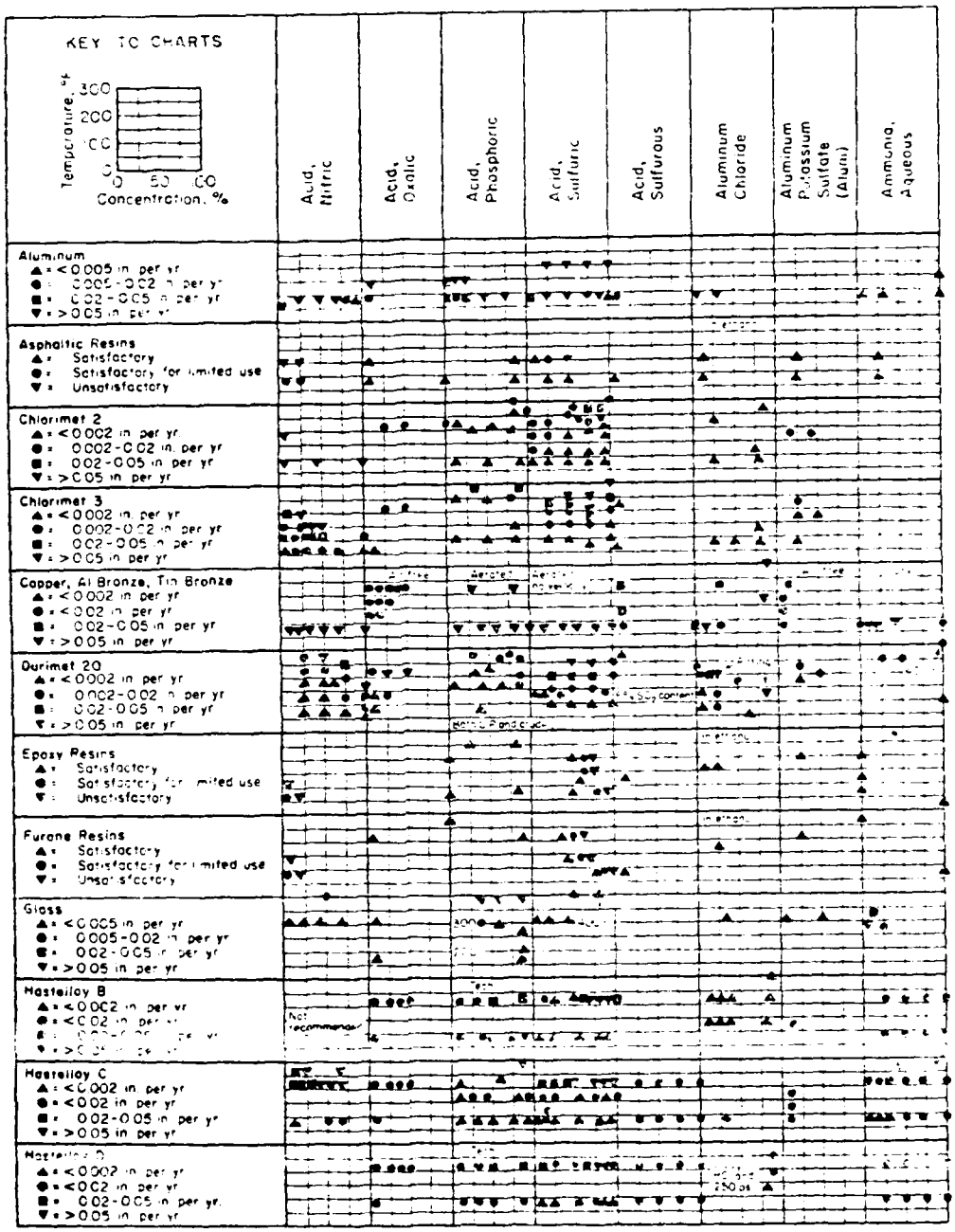

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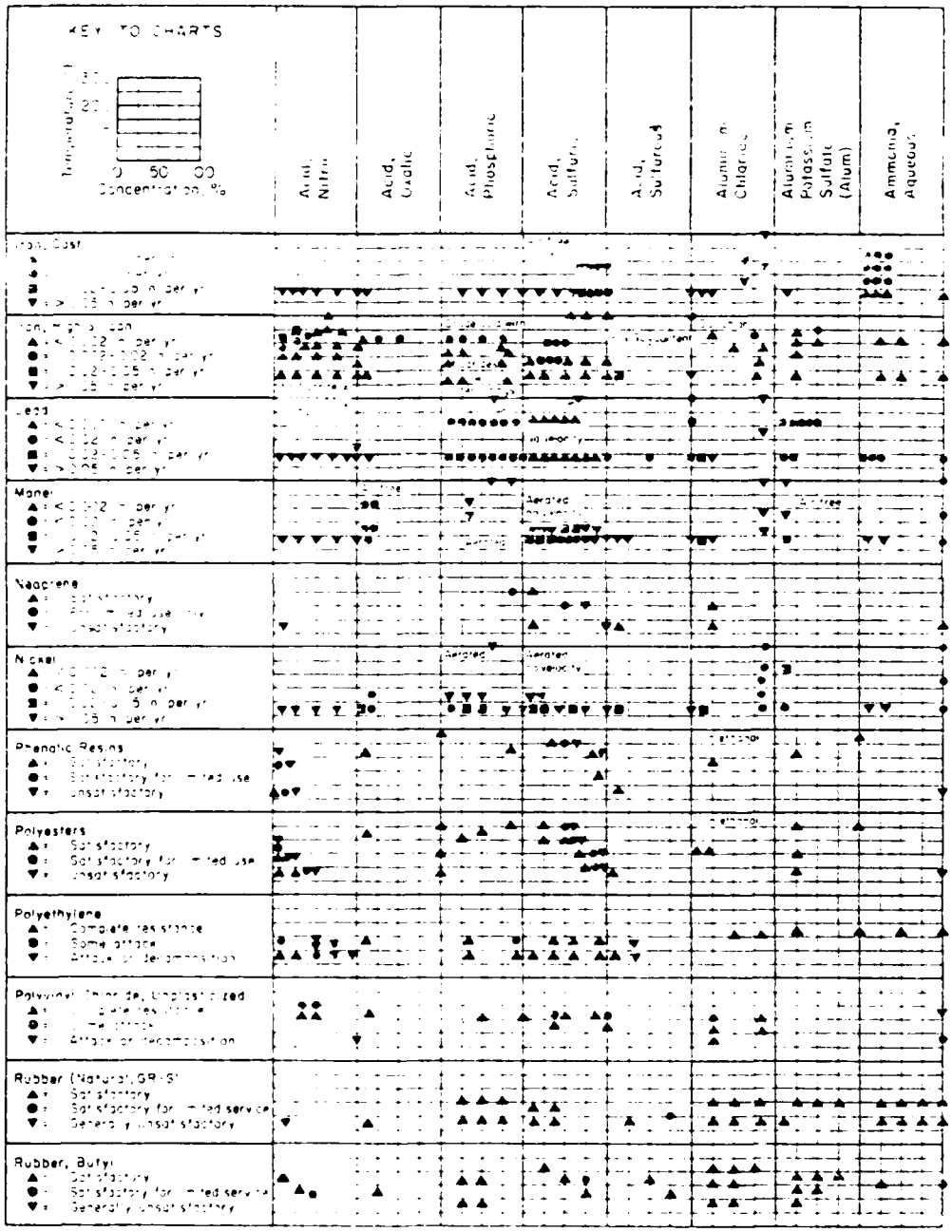

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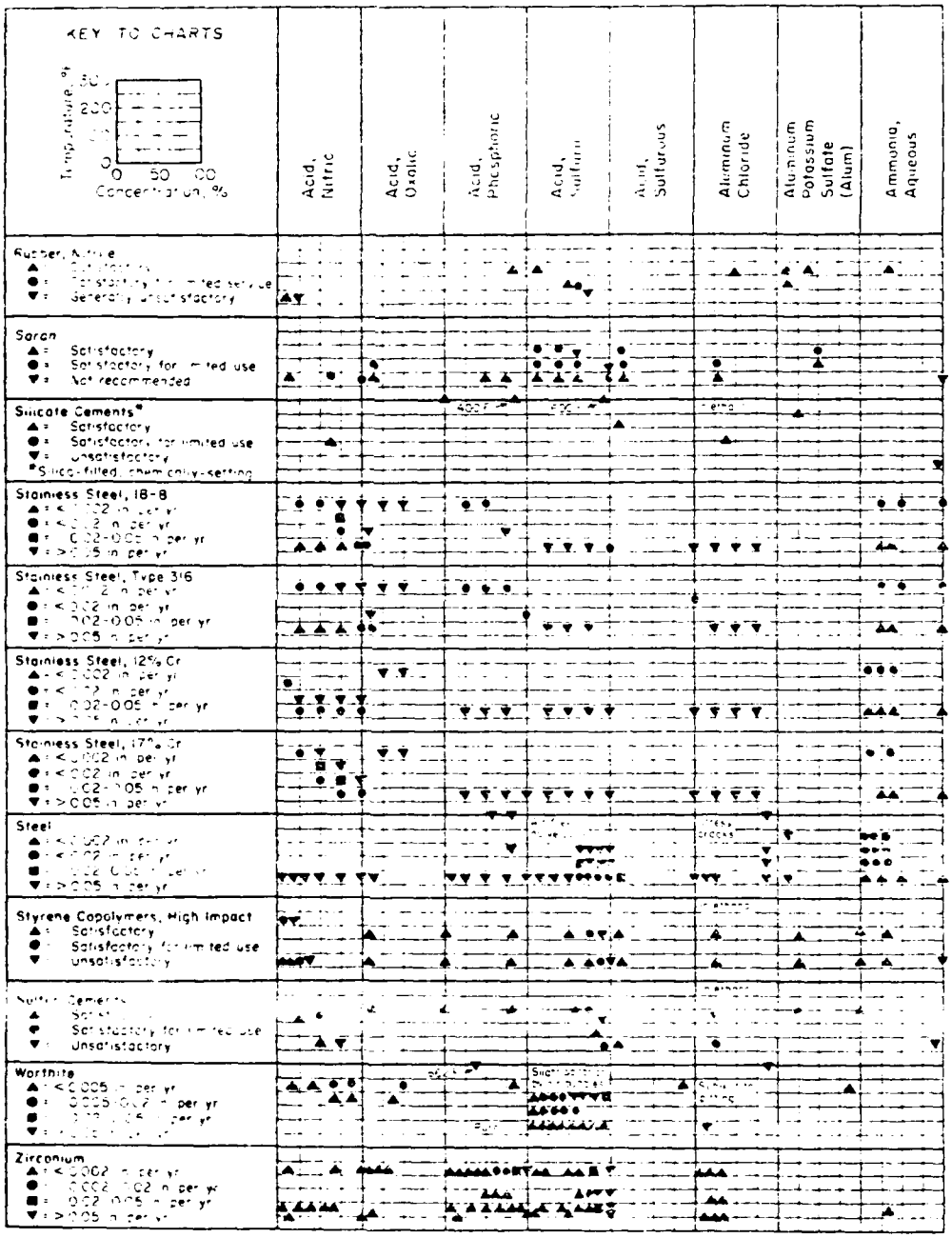

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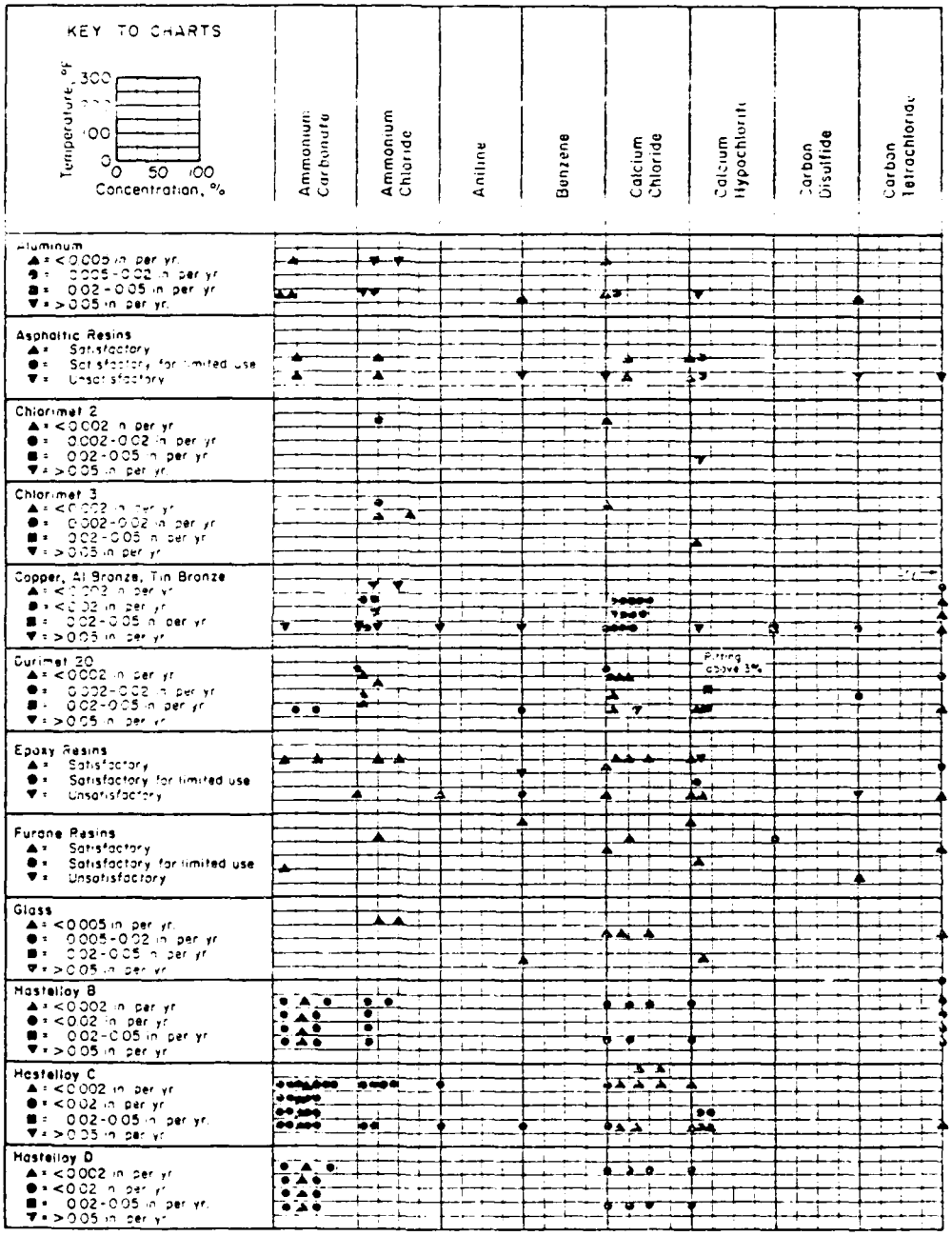

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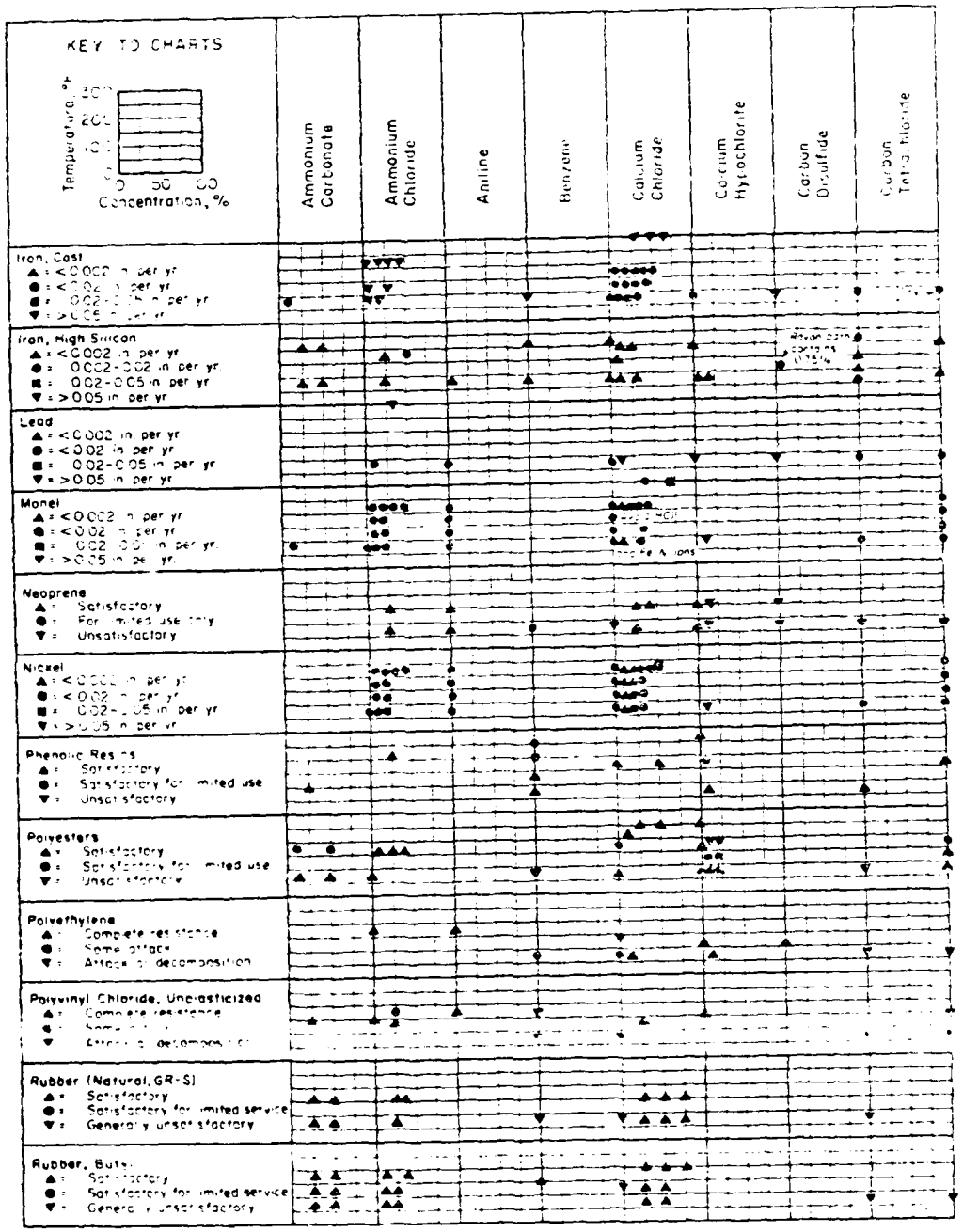


TABLE [illegible]

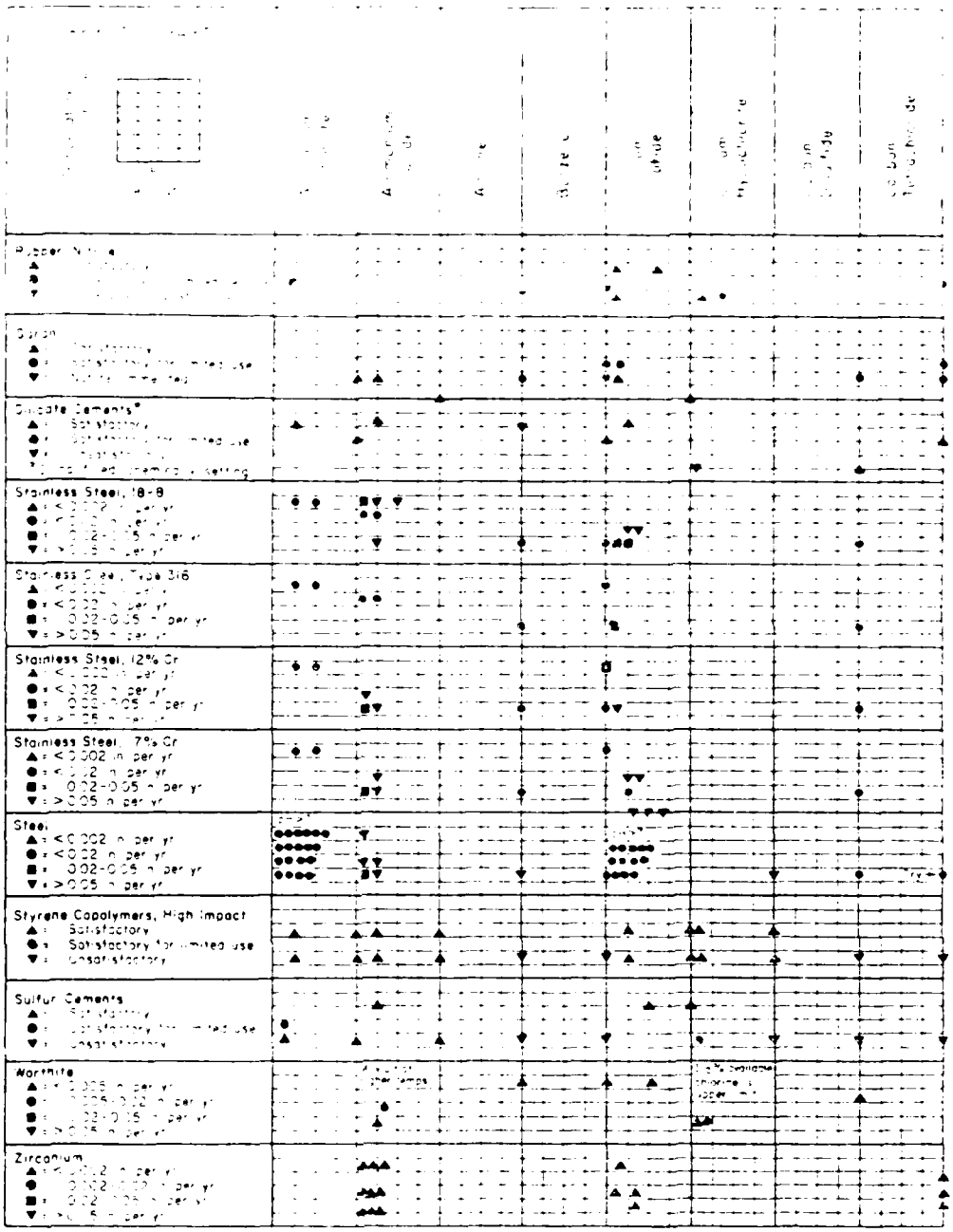
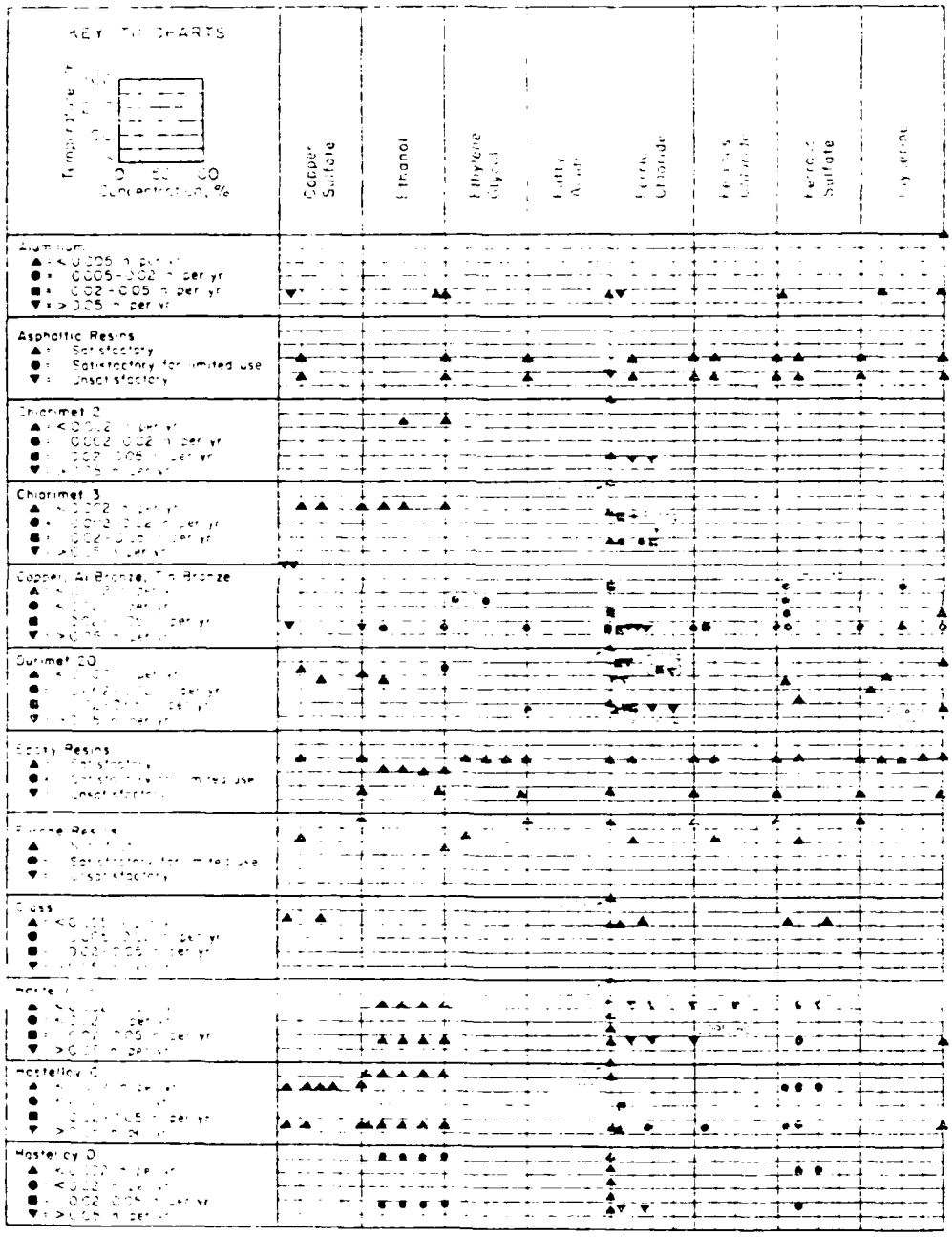

Rubber Nitrile
Satisfactory
Satisfactory for limited use
Unsatisfactory
Silicate Cements*
Satisfactory
Satisfactory for limited use
Unsatisfactory
Stainless Steel, 18-8
Stainless Steel, Type 316
Stainless Steel, 12% Cr
Stainless Steel, 17% Cr
Steel
Styrene Copolymers, High Impact
Satisfactory
Satisfactory for limited use
Unsatisfactory
Sulfur Cements
Satisfactory
Satisfactory for limited use
Unsatisfactory
Worthite
Zirconium

TABLE 5-1 (Cont.)



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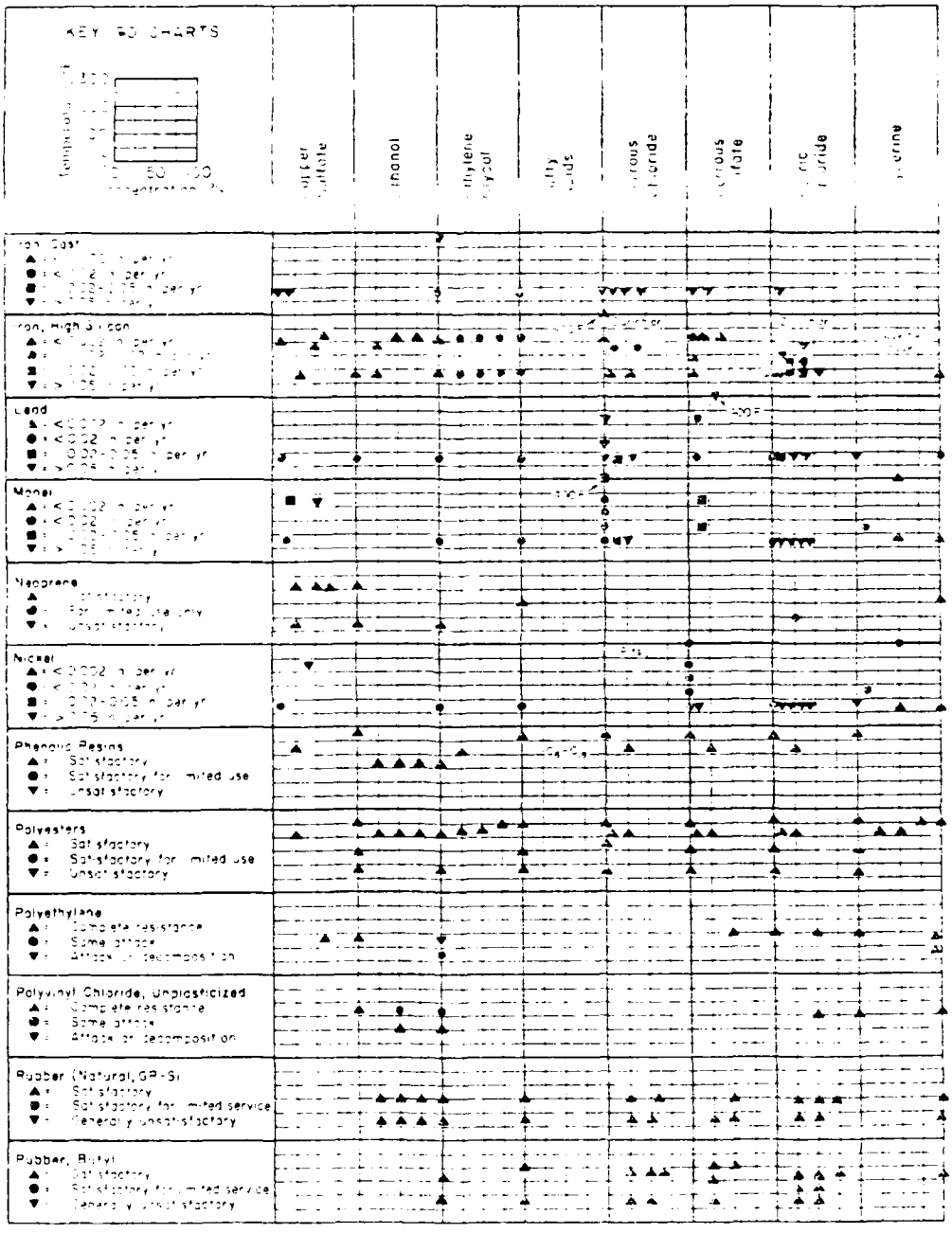

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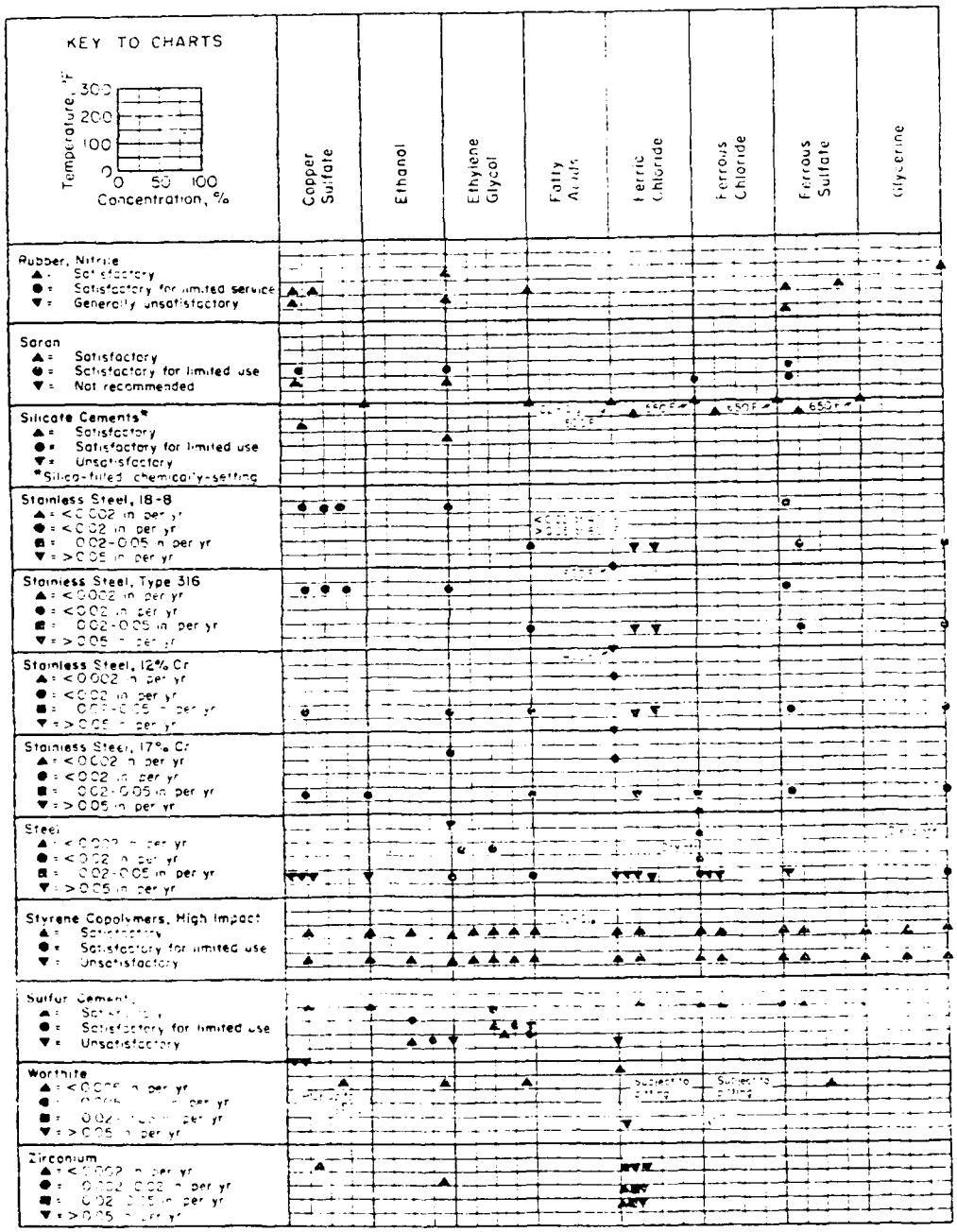

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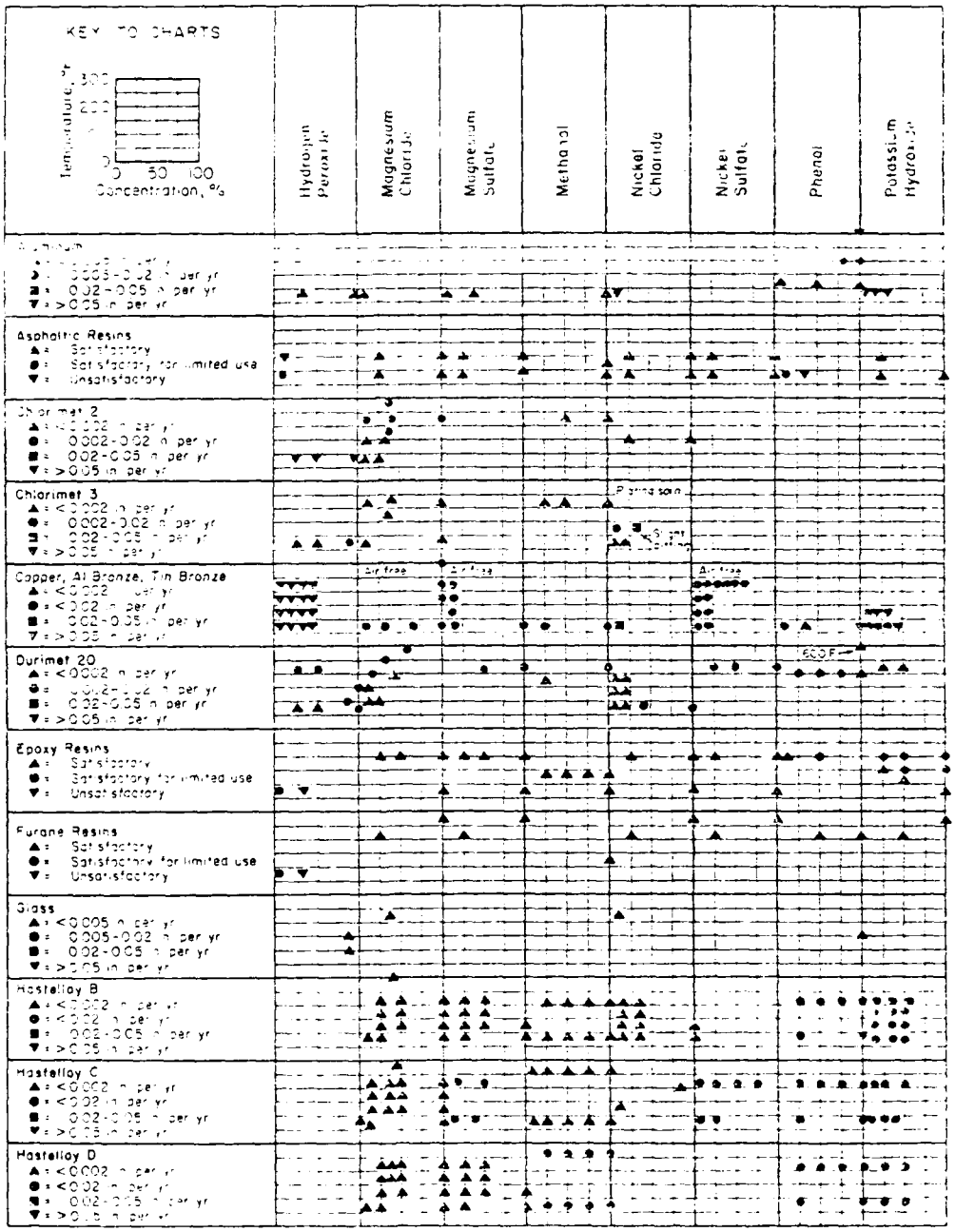

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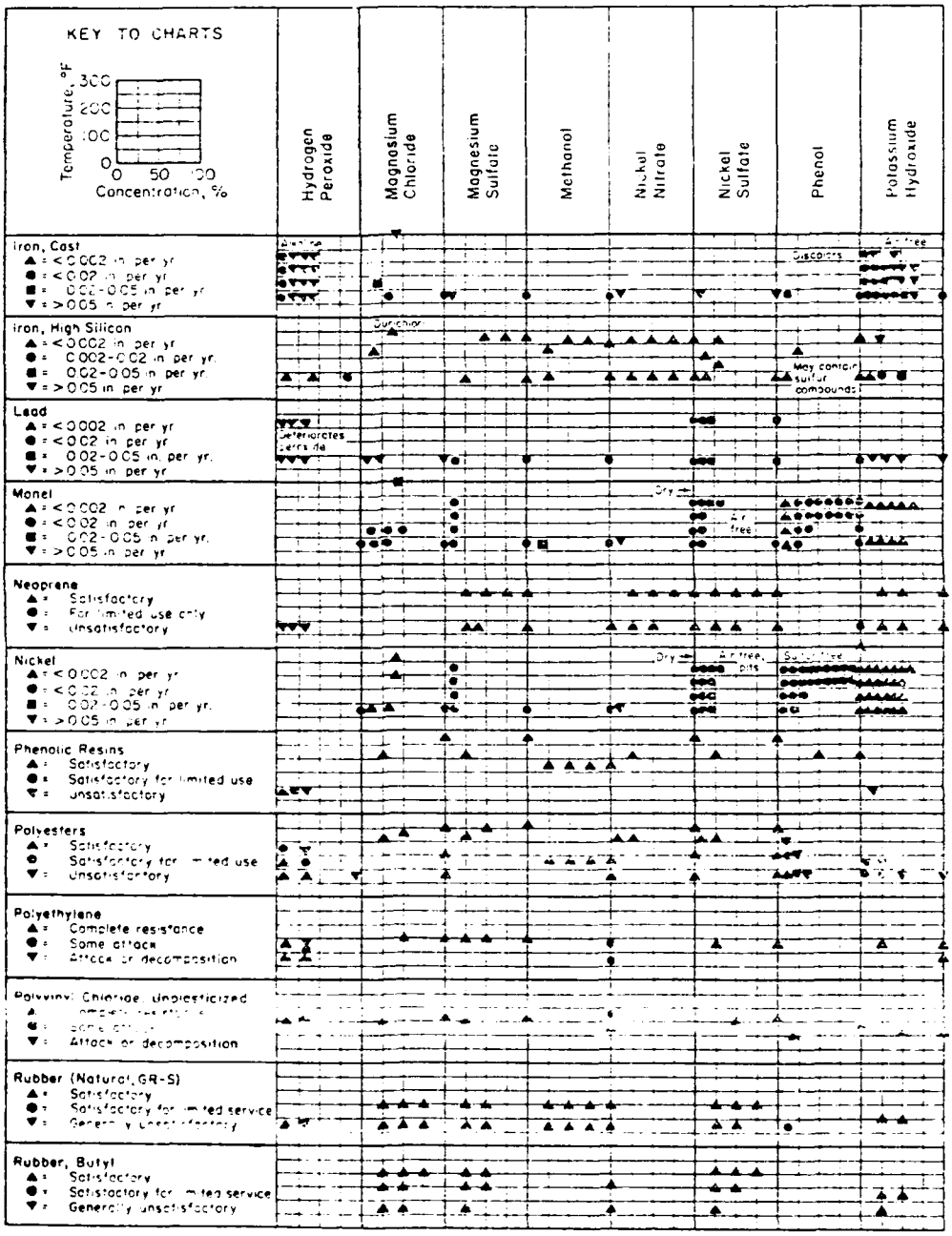

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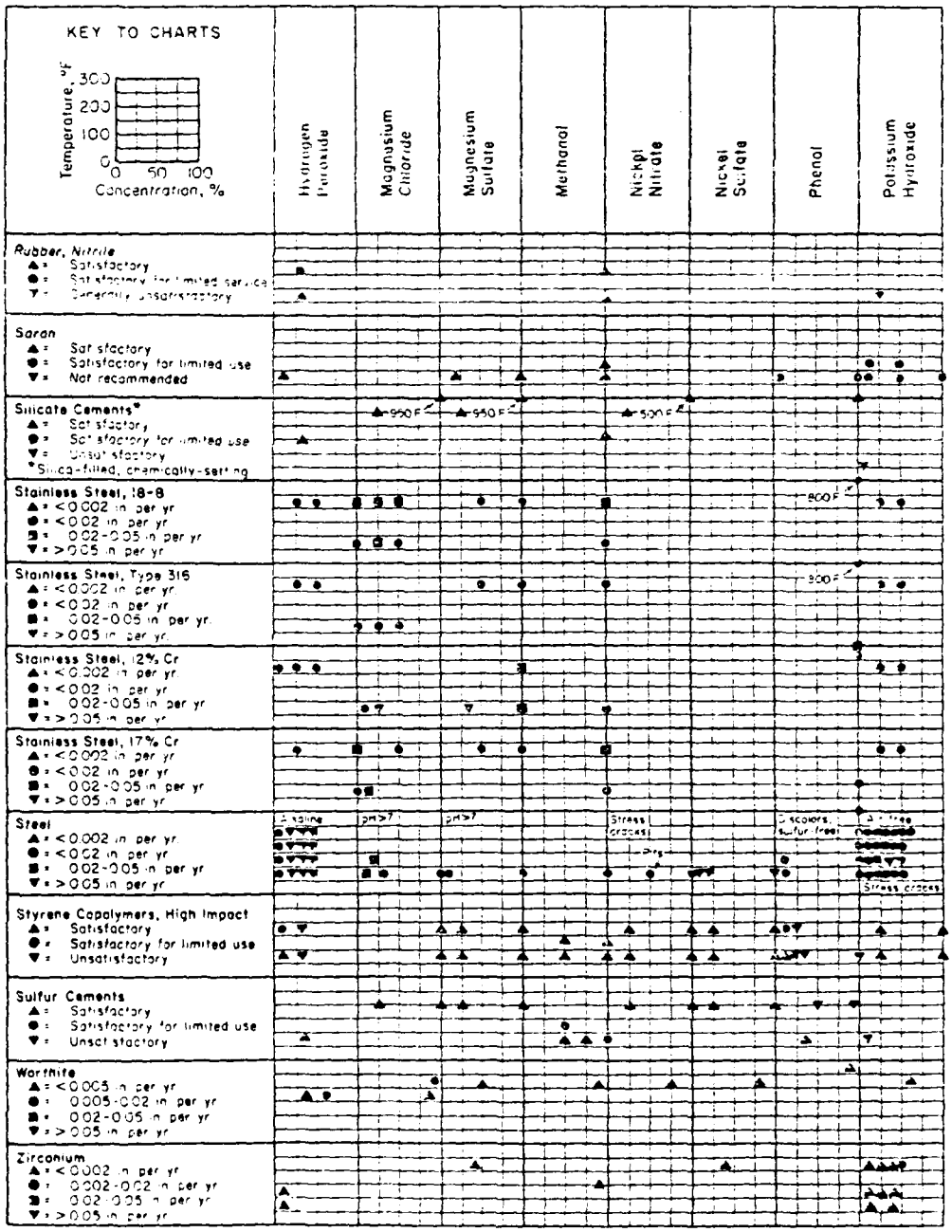


TABLE 5-1 (Cont.)

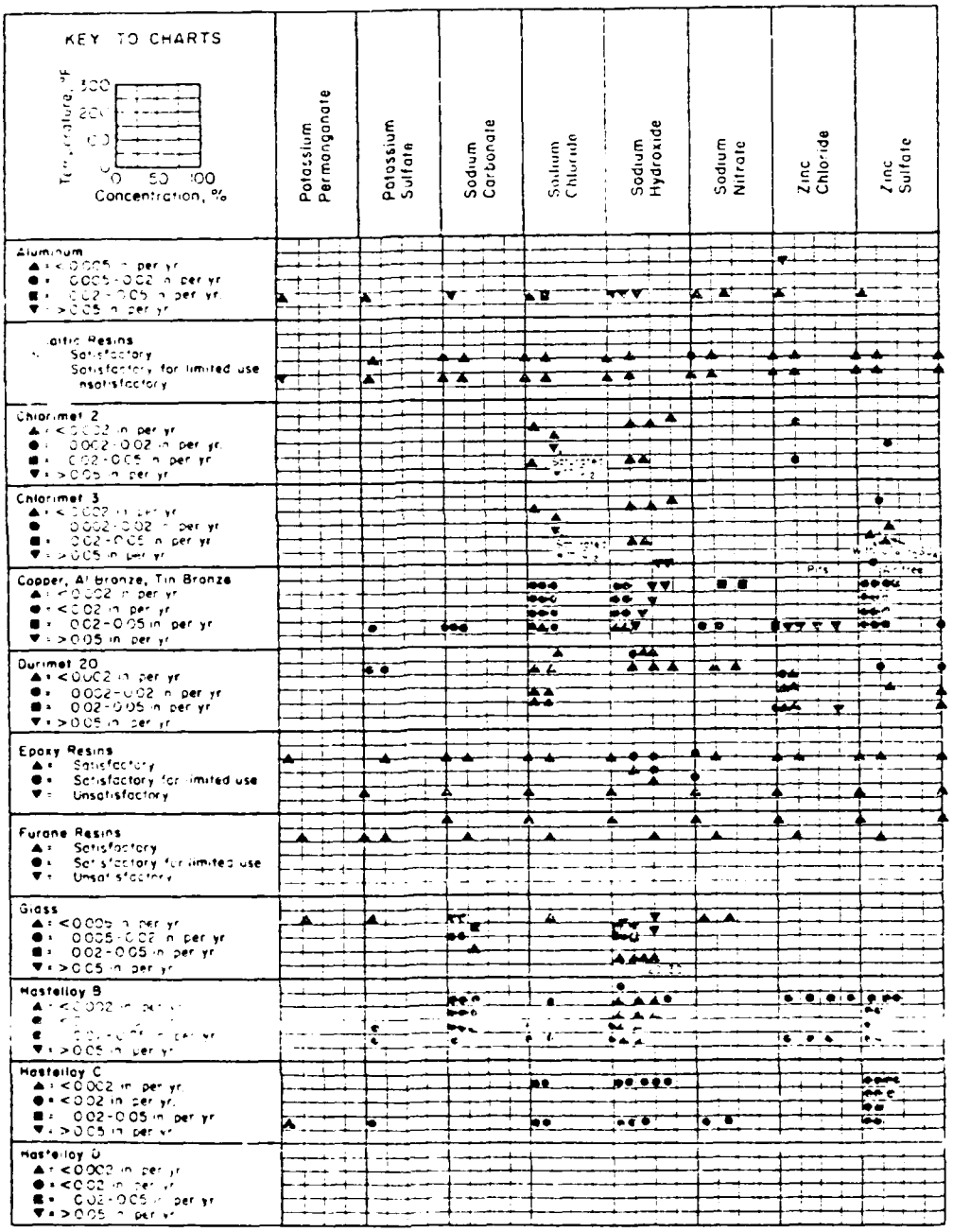
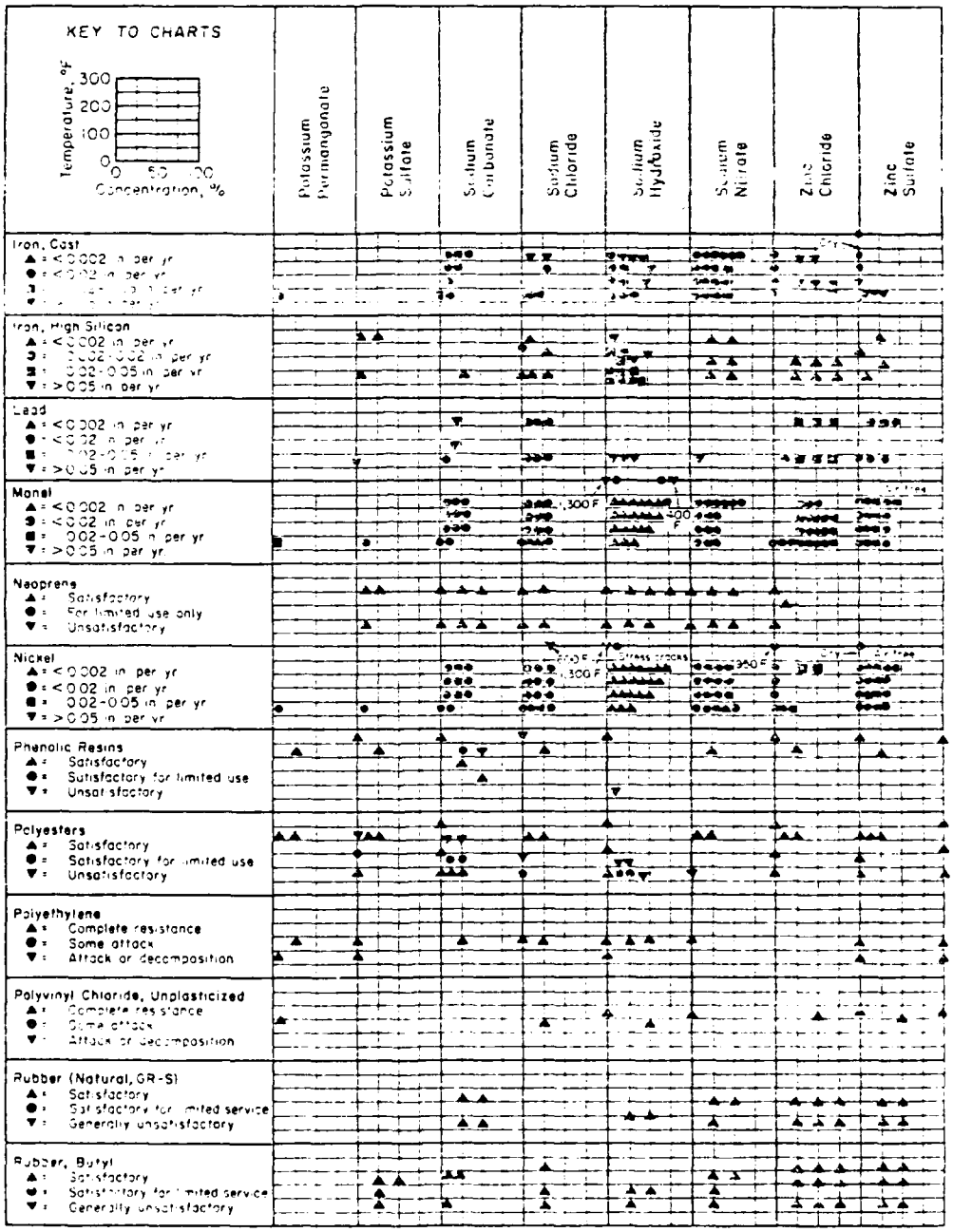


TABLE 5-1 (Cont.)



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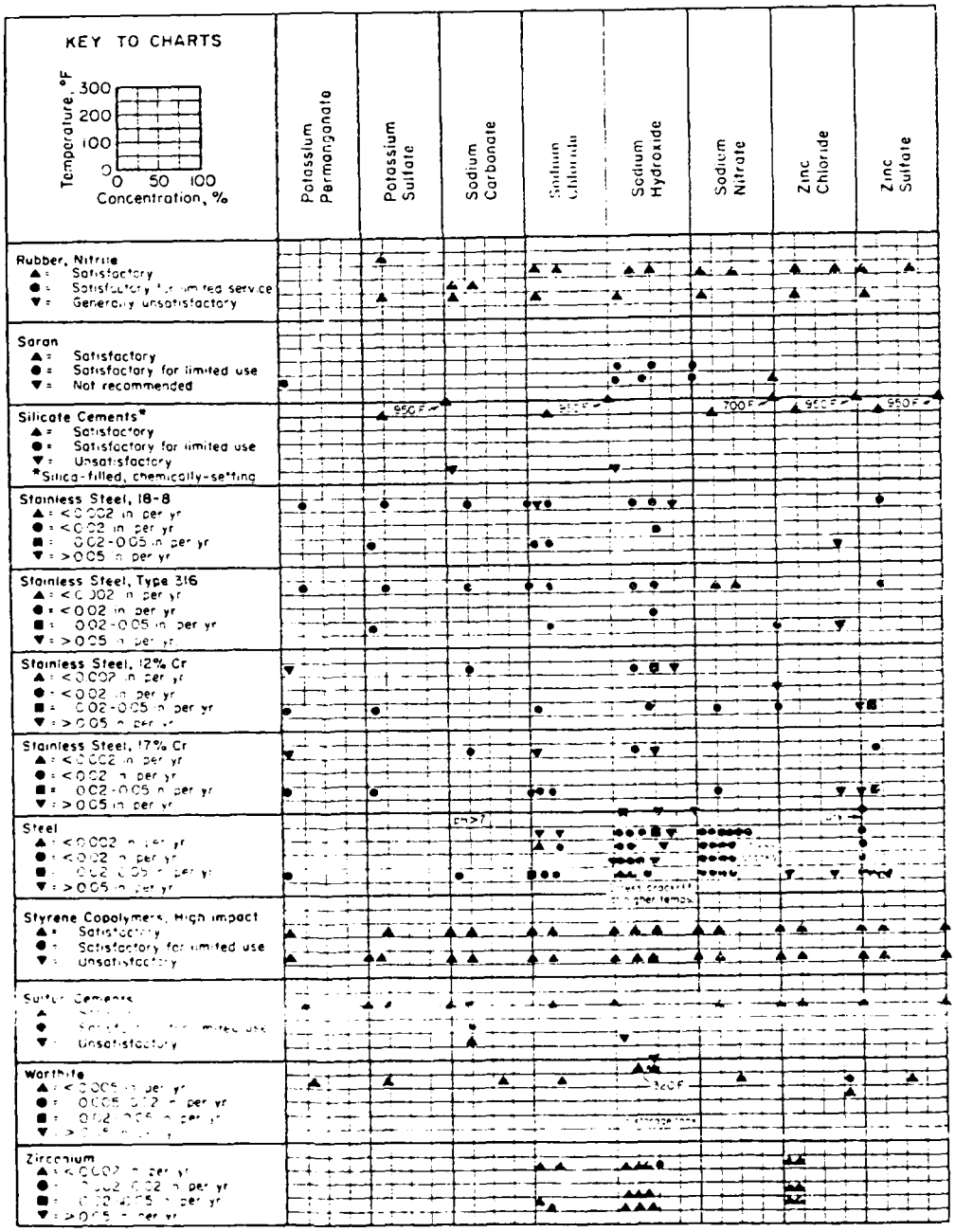

TABLE 5-2

TEMPERATURE LIMITATIONS[30]

Material	Constant Temp. °F	Intermittent Temp. °F
Natural Rubber	150-180	180-225
Neoprene	200	250
Butyl	200	250
Hypalon	200	250
PVC	150	170
Polypropylene	230	250
Flakeline 103	200	
Flakeline 200	160-180	200-230
Reinforced Plastics		
Duracor 1000	200	
2000	250	
4000	230	
6000	270	
Epoxy	250	
Phenolic	300	
Acid Brick		
H brick (fire clay)	2,500	
L Brick (shale)	1,100	
Carbon brick	2,000	
Bonding Cements		
Corobond-furan based resin	375	
Potassium silicate with inorganic setting agent	2,000	
Sodium silicate with an inorganic setting agent	2,000	

TABLE 5-2

TEMPERATURE LIMITATIONS[30]

Material	Constant Temp. °F	Intermittent Temp. °F
Natural rubber	150-180	180-225
Neoprene	200	250
Butyl	200	250
Hypalon	200	250
PVC	150	170
Polypropylene	230	250
Flakeline 103	200	
Flakeline 200	160-180	200-230
Reinforced Plastics		
Duracor 1000	200	
2000	250	
4000	230	
6000	270	
Epoxy	250	
Phenolic	300	
Acid Brick		
H brick (fire clay)	2500	
L Brick (shale)	1100	
Carbon brick	2000	
Bonding Cements		
Corobond-furan based resin	375	
Potassium silicate with inorganic setting agent	2000	
Sodium silicate with an inorganic setting agent	2000	

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CHAPTER 6
OVERVIEW OF A PILOT CASE STUDY OF A STOKER-FIRED BOILER

6.1 INTRODUCTION

This study takes the engineer through the design review procedure from definition of the problem to the selection of a control method. This example includes (1) the characterization of process variables, (2) emission profile, and (3) examination of emission regulations and determination of control requirements. Several control methods are examined in order to determine the optimum from both performance and economic considerations. The design review procedure for each applicable control method follows the outline in Figure S-1. The size, design, and performance criteria are incorporated into the cost analysis, which normally is a primary determining factor for selection.

The application concerns the boiler plant at the Anniston Army Depot, Anniston, Alabama. The smoke from the stacks of the facility is objectionable to nearby residents. Complaints are registered with the base commander and the local governmental authorities. To relieve citizens' complaints and comply with the applicable federal, state, and local air pollution codes, the base commander requests the Army Environmental Hygiene Agency (AEHA), Air Pollution Engineering Division (APED), to assess the problem and recommend a solution.

The data sheet for the case study follows.

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DATA SHEET

1. Problem Identification:

Smoke emanating from boiler stacks, Building 401, Anniston Army Depot, Anniston, Calhoun County, Alabama. Citizens complain to base commander and local governmental authorities about the obvious smoke. The base plant is located 10 miles west of Anniston, Alabama. Heating plant contains four spreader stoker coal-fired boilers with dump grates and fly-ash reinjection (three installed 1954, one installed 1961) and one spreader stoker with traveling grate and no fly-ash reinjection (installed 1977).

2. Process Description:

Saturated steam production

Use:	Process and space heat
Pressure:	145 psig
Nominal Capacity:	30,000 lb per hour
Furnace Type:	Water tube, no economizer, no superheater
Air Preheater:	None
Primary Collector:	Multi-cyclone, small diameter tubes, high efficiency
Secondary Collector:	None
Stack:	42 inches ID, 75 feet tall, one per boiler
Fans:	Induced draft between multi-cyclone and stack, 50 bhp per boiler
Soot Blow:	Once per 8-hour shift, 1.5 hours per blow, furnace tubes only

3. Process Characterization:

Steam Production Rate:

Maximum:	30,000 lb per hour
Minimum:	17,000 lb per hour
Average:	20,000 lb per hour

Fuel Rate:

Maximum:	3,300 lb per hour
Minimum:	1,500 lb per hour
Average:	2,000 lb per hour

Fuel Analysis (as fired):

Heat Content: 12,800 Btu/lb

Moisture:	5.8%
Volatile Matter:	35-45%
Fixed Carbon:	45-55%
Ash:	8.0%
Sulfur:	1.5%
Hydrogen:	5.5%
Nitrogen:	1.6%
Oxygen:	10.5%

Alternate Fuels:

None

Operating Schedule:
24 hours per day, 7 days per week,

Range of Capacity:
17,000-30-000 lb steam per hour
during 24-hour day

Availability: 300 days per year per boiler

4. Emission Characterization:

Contaminant Emissions: Emission Factors (Maximum Normal Load)

SO_2:	94 lb/hr
CO:	1.65 lb/hr
NO_x (as NO_2):	24.8 lb/hr
HC (CH_4):	0.5 lb/hr
Particulate (before collector):	264 lb/hr
Particulate (after collector, based on 75-90% collection efficiency)	26.4-66.0 lb/hr
Heat Input to Boiler:	3,300 lb/hr x 12,800 Btu/lb = 42.24 MMBtu/hr

Contaminant Emissions Calculated as lb./MMBtu: Emission Factors

SO_2:	2.2 lb/MMBtu
CO:	0.04 lb/MMBtu
NO_x (as NO_2):	0.59 lb/MMBtu
HC (as CH_4):	0.01 lb/MMBtu
Particulate (after collector)	0.625-1.56 lb/MMBtu

Contaminant Emissions: Source Test Data from AEHA Test Report No.: 21-0017-78 (Maximum Normal Load)

SO_2: 2.04 lb/MM Btu
Particulate (after collector): 1.50 lb/MM Btu
Particulate(after collector-soot blow): 7.73 lb/MMBtu

Particulate Size Distribution: Estimated from graph in Chapter 3 (before collector)

45% <40 um, 30%< 20 um, 20% < 10 um, 10%< 5 um, 2%< 2.5 um

Carrier Gas Emissions: Source Test Data from AEHA Test Report (Maximum Normal Load)

Flow Rate: 11,956 scfm
22,874 acfm

Temperature: 483°F at Stack
563°F at Boiler Exit

Pressure: -0.4 in. water gauge

Composition: CO_2 = 9.4%, O_2 = 10.8%, N_2 = 79.8%

Carrier Gas Emissions: Source Test Data from AEHA Test Report (Minimum Normal Load)

Flow Rate: 10,677 scfm
16,962 acfm

Temperature: 370°F at Stack
450°F at Boiler Exit

Pressure: -0.48 in. water gauge
Composition: CO_2 = 7.2%, O_2=13.2%, N_2 = 79.6%

5. Emission Regulations:

Alabama State Regulations, Indirect-Fired Heat Exchangers

Calhoun County Class Designation for Prevention of Significant Deterioration:

SO_2 - Class II
Particulate - Class I

Allowable Emission Rate:

SO_2 - Class II: 4.0 lb/MMBtu
Particulate - Class I: $E = 1.38(H)^{-0.44}$
where E = emission rate, lb/MMBtu
H = heat input, MMBtu/hr

$E = 1.38(42.24)^{-0.44} = 0.27$ lb/MMBtu

Comparison of allowable and actual emissions:

	Emission Rate (lb/MM Btu)	
	Actual	Allowable
SO_2		
Estimated	2.2	4.0
Source Test	2.0	4.0
Particulate		
Source Test		
Maximum Normal	1.50	0.27
Soot Blow	7.73	0.27
Estimated		
Maximum Load	0.625-1.56	0.27

Compliance: SO_2
Noncompliance: Particulate under all conditions

6. Control Requirements: (Maximum Load) Particulate

With existing multi-cyclone,

Estimated collection efficiency: 56.8-82.7%
Source test: 82.0%
Source test (soot blow): 96.5%
Without existing multi-cyclone (estimated): 95.7%

6.2 HOW TO USE DESIGN REVIEW PROCEDURE MANUAL

Baseline Data

- Emission Source (e.g., boiler) (as described in Chapter 1).
- Source Characteristics (i.e., stoker-fired, % sulfur coal, % ash, Btu heat inputs, etc.) (as described in Chapter 1).
- Applicable Air Pollution Codes (most stringent of federal, state, or local) (as outlined in Appendix A).

- Emission Source (Chapter 1):
  Boiler plant, Anniston Army Depot

- Source Characteristics (Appendix A, Chapter 1, and AEHA Test Report):

  5 Spreader Stoker Coal-fired Boilers:

  17,000-30,000 Lb/Hr steam rate/boiler
  1,500- 3,000 Lb/Hr fuel rate/boiler

  Fuel Analysis:
  Coal
  Heating Value: 12,800 Btu/lb
  H_2O : 1.5%
  Ash: 8.0%
  Soot Blow: 3 times/day, 1.5 hours
  Stack Size: 41.5 inches diameter
  76 feet height

- Applicable Air Pollution Codes (Appendix A) (Alabama State EPA):

Contaminant	Regulation	Compliance
Oxides of Nitrogen, NO_x	None	Yes
Sulfur Dioxide, SO_2	4.0 lb/MMBtu	Yes
Particulate	0.27 lb/MMBtu	No

Identification of Operating Conditions into Pollution Control System

(flow rate, temperature, pressure, particulate size distribution, gaseous contaminant loading, particulate loading, gas stream composition, etc.) (as described in Chapter 1).

From AEHA Test Report:

Flow Rate: 22,874 acfm, 11,956 scfm
Temperature: 483°F
Pressure: 30 inches Hg.
Particulate Loading: 7.73 lb/MMBtu
Particulate Size Distribution:

Micron Size	%
Over 40	47
20-40	22
10-20	15
5-10	10
2.5-5	5
Less than 2.5	1
	100

Comparison
Regulations vs. Emissions

In Compliance	Out of Compliance
o Sulfur Dioxide	Particulates
o Oxides of Nitrogen	

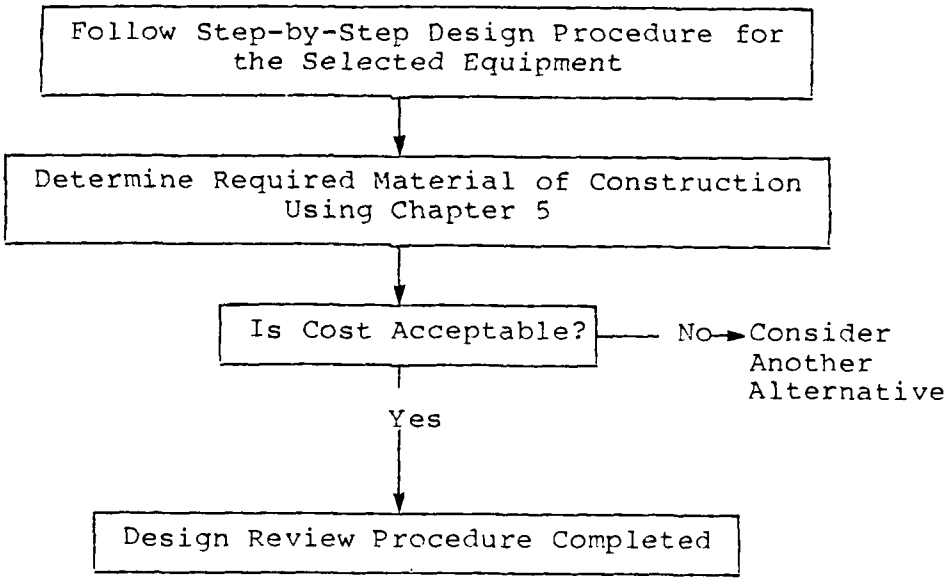
Selection of a Control Equipment from
Available Alternatives

Use Chapter 2 for
Particulate Pollutants

Since particulates are out of compliance, Chapter 2 suggests the following alternatives:

1. Centrifugal separators (Section 2.2)
2. Wet scrubbers (Section 2.3)
3. Electrostatic precipitators (Section 2.4)
4. Fabric filters (Section 2.5)
5. Combinations
6. Process modification.

Existing centrifugal separators do not meet performance requirements as they are presently being used. Wet scrubbers, electrostatic precipitators, fabric filters, and process modification require evaluation.



Following the above steps, the engineer has evaluated six separate control methods:

1. Mechanical collector
2. Wet scrubber
3. Electrostatic precipitator
4. Fabric filter
5. Combination
6. Process modification

The following are the results:

Equipment	Model/Size	Materials Of Construction	Costs Capital	Costs Operating	Acceptable
Mechanical Collector	Existing	Mild Steel	---	---	No
Wet Scrubber	Venturi/25 psi, ΔP	Stainless Steel	$ 45,473	$ 46,360/ year	Yes
Electro-static Pre-cipitator	Plate/2600 ft^2	Mild Steel	$176,750	$ 4,700/ year	Yes
Fabric Filter	Reverse Air/15,457 ft^2 BAG	Mild Steel	$124,284	$ 8,989/ year	Yes
Combinations	(Not Applicable)				
Process Modification	Low Ash Coal Prep.	----	---	$426,174/ year	No

Alternatives 2, 3, and 4 are applicable and now require authorization of procurement, which will be based on factors of overriding concern. Such factors of decision are:

- Expeditious need
- Duration of need
- Funds available
- Future regulations.

6.3 PROBLEM IDENTIFICATION

The APED engineer first defines the problem; in this case the base commander has defined the problem as smoke emanating from the five stacks of the heating plant, Building 401, housing five spreader stoker coal fired boilers. The engineer begins by collecting information about the plant. Appendix A of this manual contains the available sources of information to be used when performing a process characterization and emission assessment. Chapter 1 contains specific information about the process and emission characteristics for a general case as well as a section specific for boilers. These chapters guide the engineer in this task. A summary of the collected information pertinent to this procedure is presented at the end of this discussion.

The engineer, from specific AEHA test reports, finds a description of the plant; the heating plant contains four spreader stoker boilers with dump grates and fly-ash reinjection (three installed in 1954 and one installed in 1961) and one spreader stoker boiler with traveling grate and no fly ash reinjection (installed 1977). All five of the boilers are rated at 30,000 lb steam per hour, 145 psig. The engineer collects the layout drawings of the plant.

6.4 PROCESS CHARACTERIZATION

The engineer contacts the boiler plant operations personnel and finds that the boilers are able to operate 24 hours per day, 7 days per week. All boilers are not needed most of the time and the number of boilers on-line varies with production and the heating needs. The output varies from 17,000 to 30,000 lb steam per hour for each boiler. Average production rate over the 24 hours is 20,000 lb steam per hour. Corresponding fuel rates are 1,500 lb per hour minimum to 3,300 lb per hour maximum, with average fuel rate of 2,000 lb per hour. In addition, the AEHA test reports a fuel analysis: heating value - 12,800 Btu/lb, moisture = 5.8%, ash = 8.0%, sulfur = 1.5%. The plant operations verify that the same fuel is being used. If the engineer must calculate theoretical gas volumes based on fuel analysis, the fixed carbon and volatile matter must be analyzed or estimated. A table of approximate values can be found under Coal Classification, Boiler and Heating Plants, Chapter 1.

Documenting the process flow is best done on a site visit, observing the equipment and talking to the operators. In lieu of a plant trip, the engineer can obtain the information from various sources: plan drawings of the facility, operating permit applications, test reports, construction reports, direct communication with the plant supervisor, etc. The gathered information shows that each operating boiler generates steam for process and space heating in a water tube furnace. No economizer or air preheater is used, and a multi-cyclone is utilized on the boiler exhaust to collect fly-ash under negative pressure. Downstream of the mechanical collector, the induced-draft fan provides the prime gas-moving force. The induced-draft fan is powered by a 50-bhp motor and gas throughput is controlled by a damper on the outlet side of the fan. Each boiler exhausts gas to a stack 41.5 inches ID and 76 feet tall.

Boiler soot blowing is performed once during each shift (three times per day) and lasts for a duration of 1.5 hours each blow.

6.5 EMISSION CHARACTERIZATION

The engineer is now ready for the emission assessment section of the review procedure. In Chapter 1 of this report, the emission factors for coal-fired stoker boilers are presented. Other sources of information can be found in Appendix A, and include emission test reports. In this case, there is an AEHA test report that presents the results of actual emission tests while the boiler was firing the currently used coal. The emission assessment presented here is a profile based on one boiler only. This is for simplification of the review procedure; in actuality, the considerations should be extended to all boilers and to the feasibility of collecting the gas streams from all boilers to one control device (see Appendix A).

Emission factors from Chapter 1 are used to quantify the uncontrolled gaseous and particulate emissions. These are the emissions that would result if no control device existed on the boiler system. These emission rates are (maximum load: 3,300 lb/hr coal feed):

SO_2:	94 lb/hr
CO:	1.65 lb/hr
NO_x (as NO_2):	24.8 lb/hr
HC (as CH_4):	0.5 lb/hr
Particulate:	264 lb/hr

Using the maximum fuel rate of 3,300 lb/hour and heat content of the fuel 12,800 Btu/lb, the engineer calculates maximum heat input to the boiler of 42.24 million Btu per hour. The calculated emissions in units of lb per million Btu (units of allowable emissions) are as follows:

SO_2:	2.20 lb/million Btu
CO:	0.04 lb/million Btu
NO_x (as NO_2):	0.59 lb/million Btu
HC (as CH_4):	0.01 lb/million Btu
Particulate:	6.25 lb/million Btu

Since the existing particulate collector on the boiler is a multi-cyclone, with small diameter, high-efficiency tubes, the engineer searches the mechanical collector section of Chapter 2 and finds the average particulate collection efficiency is 75 to 90 percent. This results in an estimated particulate emission rate of 26.4 to 66.0 lb/hr or 0.625 to 1.56 lb/million Btu.

As a check on the emission rates, the engineer examines the AEHA test report and compares the estimated and measured emissions. The source test results show

particulate emissions, 1.50 lb per million Btu, and SO_2 emissions, 2.04 lb per million Btu, which compare favorably with the estimates. The engineer also finds in this test report, a source test that was performed during soot blow. The particulate results are 7.73 lb per million Btu and 185 lb per hour.

No particle size data are available from the emission tests, so the engineer examines the average particle size data presented graphically in Chapter 1, Boilers and Heating Plants: 45%<40μ, 30%<20μ, 20%<10μ, 10%<5μ, and 2%<2.5μ. This distribution is representative of particles leaving the boiler, prior to the collection device. Average fractional efficiency information in Chapter 2 can be used to estimate the emitted particle size distribution; however, this is of dubious value. When it is necessary to design a control system, particle size tests should be performed.

The second important emission characteristic to be assessed is the flue gas flow rate. From the AEHA test report, it is learned that the measured volumetric flow rate is 11,956 scfm (standard cubic feet per minute at 70°F, 29.92 inches mercury, dry) at maximum load. After correcting for moisture content (7.0%) and temperature (483°F), the actual flow rate is 22,874 acfm. As a check, the engineer can use theoretical gas volume calculations with fuel analysis, gas oxygen analysis, and fuel rate to arrive at 10,718 scfm and 28,621 acfm. Oxygen and carbon dioxide contents were 10.8 and 9.4 percent respectively.

A minimum load test result from the AEHA test report shows 10,677 scfm and 16,962 acfm.

6.6 REVIEW OF ALLOWABLE EMISSION REGULATIONS

The engineer must examine the applicable emission control regulations in order to determine whether or not the boilers are in compliance. The regulations are available at AEHA; however, the engineer must confirm that regulations are current by making a telephone call to the state agency. This telephone call also confirms that the state agency has jurisdiction over the area and not the local agencies. Appendix A of this report presents the details of examining applicable regulations.

Calhoun County, the county in which Anniston Army Depot is situated, is designated Class II for SO_2 and Class I for particulate. Class II SO_2 allowable emission for boilers is 4.0 lb per million Btu, which means that the actual 2.0 to 2.2 lb per million Btu emission rate is well within the maximum limits. No NO_x emissions are presented for boilers less than 250 million Btu per hour of heat input. (A special case may be where the streams of all boilers are gathered and emitted at one point; in this case, the total heat input to the stack may be greater than 250 million Btu per hour.)

Particulate emissions are limited on a sliding scale depending on the boiler heat input according to the following formulas:

$$E = 1.38\ (H)^{-0.44} \text{ Class I Counties}$$
$$E = 3.109\ (H)^{-0.589} \text{ Class II Counties}$$

where E = allowable emission rate in lb per million Btu
and H = heat input to boiler in million Btu per hour

Since Calhoun County is Class I for particulate emissions, the first formula is used to determine the allowable limits:

$$E = 1.38\ (42.24)^{-0.44} = 0.27 \text{ lb per million Btu}$$

This value is less than the actual emission rate of 1.50 lb per million Btu, as determined by emission tests, demonstrating noncompliance.

The boilers are not in compliance with the particulate regulations. The engineer must calculate the degree of control efficiency necessary to meet the applicable code and eliminate the visible smoke from the stack effluents. The control requirement is calculated as follows:

$$\%\ \text{efficiency} = (1-(\text{allowable}/\text{actual})) \times 100$$

The percent efficiency is calculated under all conditions:

1. With existing multi-cyclone
   a. Estimated maximum normal load (75% collection efficiency of multi-cyclone) - 82.7%

   b. Estimated maximum normal load (90% collection efficiency of multi-cyclone) - 56.8%

   c. Measured maximum normal load - 82.0%

   d. Measured maximum load while blowing soot - 96.5%

2. Without existing multi-cyclone
   a. Estimated maximum normal load (without soot blowing) - 95.7%

In order to ensure meeting existing and future emission codes, and to eliminate visible smoke while blowing soot, the engineer decides to obtain 98 percent collection efficiency with the multi-cyclone still in operation.

6.7 EVALUATION OF CONTROL METHODS

After establishing the degree of control necessary to meet the regulatory requirements, the engineer must evaluate the different methods of control. Chapter 1, Boiler and Heating Plants section, and Chapter 2 on particulate collection equipment, guide the engineer to decide that three types of control equipment can be designed for 98 to 99 percent collection efficiency. These are (1) electrostatic precipitators, (2) fabric filters, and (3) venturi scrubbers. In addition, the coal supplier may be able to supply an expensive low-ash coal that may meet the requirements of the emission regulations without installing the expensive control equipment. The next step in the design review procedure are to size the individual piece of equipment, develop basic design criteria, and estimate the approximate cost. These are covered in the following sections.

6.7.1 Wet Scrubber

1. Calculate the number of transfer units needed to remove 98 percent of the particulate material.

   $N_t = (1/(1-\eta))$, from Chapter 2

   where N_t = number of transfer units
   η = collection efficiency = 98%
   $N_t = \ln (1/(1-0.98)) = \ln 50$
   $N_t = 3.9$

2. Choose a gas pressure drop. Since this is a medium-to-high efficiency scrubber, choose 25 inches of water.

3. Calculate contacting power using equations in Chapter 2.

   $$P_G = 0.1573\ (\Delta p)$$

   where P_G = contacting power based on gas stream energy input, hp/1,000 acfm

   Δp = gas pressure drop across scrubber, inches of water

   $$P_G = 0.1573\ (25) = 3.933 \text{ hp/1,000 acfm}$$

Total gas input power = 3.933 hp/1,000 acfm x 22,900 acfm

= 90 hp

Also, $P_L = 0.583\ p_1\ (L/G)$

where P_L = contacting power based on liquid steam energy input, hp/1,000 acfm

p_1 = liquid inlet pressure, psi
L = liquid feed rate, gpm
G = gas flow rate, acfm

Since venturi scrubbers normally operate at 6 to 10 gallons/1,000 acfm (from Chapter 2), the L/G in this equation is 0.006 to 0.010; choose 0.008 gallons/acfm.

At 20 psi inlet pressure,

$$P_L = 0.583\ (20)\ (.008) = 0.093 \text{ hp/1,000 acfm}$$

$$P_T = P_L + P_G$$

where P_T = total contacting power, hp/1,000 acfm

$$P_T = 3.933 + 0.093 = 4.026 \text{ hp/1,000 acfm}$$

4. Check number of transfer units, using and from table in Chapter 2.

$$N_t = \alpha P_T^{\beta}$$
$$= 3.928\ (4.026)^{0.189} = 5.1$$

Note: This is an estimate based on empirical values. Since this value is higher than the N_t = 3.9, calculated in Step 1, it is prudent to assume that the 25-inch water pressure drop is more than adequate to control fly-ash emissions to acceptable limits.

5. Calculate capital equipment cost for venturi scrubber sized for 22,900 acfm. Refer to Chapter 4 for details. Since 1/8-inch plate is adequate for 22,900 acfm and 25-inch pressure drop, the basic equation is used (Capital cost in dollars).

$$S = 8{,}095 + 463\ (1{,}000 \text{ acfm}) - 0.97\ (1{,}000 \text{ acfm})^2$$
$$S = 8{,}095 + 463\ (22.9) - 0.97\ (22.9)^2$$
$$S = 18{,}189$$

6. Choose the materials of construction. From Chapter 5, using wet sulfur dioxide as reference, carbon steel is unsatisfactory, 304 stainless steel is acceptable for limited service, and 316 stainless steel is recommended. From Chapter 4, the adjustment factor for 316 stainless steel is 2.5.

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Cost = \$18,189 x 2.5 = \$45,473

Note: This figure does not include the increased fan capacity that would be required for a 25-inch water pressure drop.

7. Calculate the operation and maintenance cost. Refer to Chapter 4 for details.

$$\$ = S \left[0.7457\ HK \frac{P}{6356E} + \frac{Qg}{1722F} + \frac{Qh}{3960F} + WHL + M \right.$$

$$\$ = 22{,}900 \left[0.7457(8760)(0.4)\left(\frac{25}{6356(0.60)} + \frac{(0.008)20}{1722(0.5)} + \frac{(0.008)100}{3960(0.5)} \right) + (0.003)(8760)(3.5 \times 10^{-4}) + 0.15 \right]$$

$$= \$46{,}360$$

8. Calculate the total capital cost of a wet scrubber - following the itemized list from Section 4.2 of the manual:

(1)	Equipment Costs	\$ 45,473
(2)	Tax and Freight at 7% of (1)	3,183
(3)	Installation Costs (use 100% equip. cost for this example)	45,473
(4)	Subtotal (1) + (2) + (3)	94,129
(5)	Engineering at 10% of (4)	9,413
(6)	Subtotal (4) + (5)	103,542
(7)	Contingencies at 10% of (6)	10,354
(8)	Total Capital Costs (6) + (7)	\$113,896

9. Calculate the uniform annual cost for the wet scrubber assuming that the present rate of return required on funds spent is 12 percent and the service life is assumed to be 10 years.

From Table 4-16, the capital recovery factor (CRF) is 0.17698

Equivalent Annual Cost = Total Capital Cost x CRF
= \$113,896 x 0.17698
= \$20,157

Uniform Annual Cost = Equivalent Annual Cost + Operating and Maintenance Cost
Uniform Annual Cost = \$20,157 + \$46,360
= \$66,517

6.7.2 Electrostatic Precipitator

1. Estimate resistivity from graph of resistivity versus temperature at 1.5 percent sulfur coal. The estimated resistivity is 3×10^8 ohm-cm (Chapter 2).

2. Estimate precipitation rate parameter from graph of resistivity versus precipitation rate parameter. The estimated rate parameter is 0.574 ft/sec (curve must be extrapolated in this case).

3. Apply the Deutsch equation using the available data:

$$\eta = 1 - e^{(-AW/Q)}$$

where Q = Gas flow rate, acfm
η = Desired efficiency, fraction
W = Precipitation rate parameter, ft/sec
A = Net plate area, ft^2

Therefore,

$$A = \frac{Q}{W} \ln\left(\frac{1}{1-\eta}\right)$$

A 98 percent collection efficiency is used, assuming the existing collector is not removed.

or $$A = \frac{22{,}900 \text{ ft}^3/\text{min}}{0.574 \text{ ft/sec} \times 60 \text{ sec/min}} \ln\left(\frac{1}{1-.98}\right)$$

$$A = 2{,}600 \text{ ft}^2. \text{ precipitator plate area}$$

4. Check degree of sectionalization (See graph of efficiency versus bus sections/100,000 acfm. Since this is a small precipitator with less than 100,000 acfm, one bus section is adequate.

5. Determine electrical power required to operate precipitator. From Figure 2-29, the power required is 120 watts/1,000 acfm, or 0.12 kW/1,000 acfm. For 22,900 acfm, total power = 22.9 x 120 = 2,748 watts exclusive of gas moving power.

6. Estimate capital equipment cost of insulated precipitator (from Chapter 4).

$$\begin{aligned} \$ &= 163{,}750 + 5.0\ (A) \\ &= 163{,}750 + 5.0\ (2{,}600 \text{ ft}^2) \\ &= 176{,}750 \end{aligned}$$

7. Estimate operation and maintenance (O&M) costs (refer to Chapter 4, Electrostatic Precipitators Section).

$$\$ = acfm\ [JtP_c + M]$$
$$= 22{,}900\ [(5 \times 10^{-4})\ (8760)\ (00.4) + 0.03]$$

O&M Cost = $4,700

8. Calculate the total capital cost of an electrostatic precipitator, following the itemized list from Section 4.2 of the manual.

(1)	Equipment Costs	176,750
(2)	Tax and Freight at 7% of (1)	12,373
(3)	Installation Costs (use 100% equip. cost for this example)	176,750
(4)	Subtotal (1) + (2) + (3)	365,873
(5)	Engineering at 10% of (4)	36,587
(6)	Subtotal (4) + (5)	402,460
(7)	Contingencies at 10% of (6)	40,246
(8)	Total Capital Costs (6) + (7)	$442,706

9. Calculate the uniform annual cost for an electrostatic precipitator assuming again that the present rate of return is 12 percent and the service life is 20 years.

From Table 4-16, the capital recovery factor is 0.13388.

Equivalent Annual Cost = Total Capital Cost x CRF
= $442,706 x 0.13388
= $ 59,269

Uniform Annual Cost = Equiv. Annual Cost + O&M Costs
Uniform Annual Cost = $59,269 + $4,700
= $63,969

6.7.3 Fabric Filter

1. Fabric filter must operate continuously; therefore, a compartmented unit is used (see Chapter 2).

2. A suction unit is used so that the fan erosion is minimized (see Chapter 2).

3. Establish fabric type (see Chapters 1 and 2). Glass fiber bags are best for high temperature (500°F). The glass fiber material has poor flexural characteristics, therefore a gentle cleaning method should be used, e.g., reverse flow.

4. Select air-to-cloth ratio. From Chapters 1 and 2, air-to-cloth ratios of 2.0 to 2.3 to 1 are usually used for coal fly ash and glass fiber bags. Choose 2.0 to 1, which is a conservative value.

5. Choose pressure drop. Pressure drop ranges from 3 to 5 inches of water, maximum 6 inches of water (from Chapter 2).

6. Calculate net cloth area.

$$A_n = \frac{22{,}900\ ft^3/min}{2.0\ ft^3/min\ ft^2} = 11{,}450\ ft^2$$

Adjust to gross cloth area. Adjustment factor = 1.5 from Chapter 4. Since a conservative air-to-cloth (A/C) ratio was used, a factor of 1.35 is chosen.

$$A_g = 11{,}450 \times 1.35 = 15{,}457\ ft^2$$

7. Establish bag size and calculate number of bags needed. From Chapter 2 0.5 ft diameter x 12 feet high bag size is selected.

$$\text{Area of each bag} = 2\ \pi rh = 18.85\ ft^2/bag$$
$$\text{Number of bags} = \frac{15{,}457\ ft^2}{18.85\ ft^2} = 820\ \text{bags}$$

8. Choose number of compartments. Use information in Chapter 2 to guide selection. Ten compartments should be adequate to provide enough area when one compartment is out of service for cleaning.

820 bags/10 compartments = 82 bags/compartment.

Since bags are in approximately square configuration within each compartment, nearest multiplier is 9 x 9 = 81 bags/compartment. New cloth area is 18.85 ft^2/bag x 81 bags/compartment x 10 compartments = 15,268 ft^2. With one compartment out of service for cleaning, cloth area is 0.9 x 15,268 ft^2 = 13,741 ft^2. New A/C ratio is 22,900/13,741 = 1.67.

9. Through optimization of the cloth area, an A/C ratio closer to 2.0 can be found with one compartment out of service: 22,900/2.0 = 11,450 ft^2. With all compartments in service, 11,450/0.9 = 12,722 ft^2. New number of bags is 12,722 ft^2/18.85 ft^2/bag = 675 bags. With 10 compartments, nearest multiplier is 8 x 9 = 72 bags/compartment. Total number of bags = 720. New cloth area is 720 bags x 18.85 X ft^2/bag = 13,572 ft^2. New A/C ratio is 22,900/0.9 (13,572) = 1.87.

Since this is closer to objective of 2.0, this conservative cloth area will be used in estimating capital equipment cost.

10. Estimate bag costs. Refer to cost table in Chapter 4 for glass bags and reverse flow cleaning. Cost = $\$0.52/ft^2$.

$$\text{Bag Cost} = 13{,}572\ ft^2 \times \$0.52/ft^2 = \$7{,}057$$

Estimate baghouse structure cost. Refer to Chapter 4 for costs of continuous pressure, reverse air fabric filter.

$$\text{Cost} = \$29{,}210 + 3.40\ (\text{net cloth area})$$
$$\text{Cost} = \$29{,}210 + 3.40\ (13{,}572) = \$75{,}355$$

For suction, add-on cost = \$1,925 + 0.37 (cloth area)
= \$6,950

For insulation, add-on cost = \$12,010 + 1.83 (net cloth area)
+ 1.83 (cloth area)
= \$36,850

Calculate total capital equipment cost:

Total Cost = Bag cost + structure cost + suction add-on cost + insulation add-on cost
= \$7,057 + \$75,355 + \$6,950 + \$36,850
= \$126,212

11. Estimate operating and maintenance costs. Refer to Chapter 4 for cost estimating formulas.

$$\text{O\&M Costs}(\$) = acfm[1.173 \times 10^{-4}\ (\Delta P)\ (tE/\eta_f) + M]$$
$$= 22{,}900[1.73 \times 10^{-4}\ (5)\ (8760)\ (0.04)/(0.6) + 0.05]$$
$$= \$12{,}713$$

12. Calculate the total capital cost of a baghouse-fabric filter - following the itemized list from Section 4.2 of the manual.

(1)	Equipment Costs	\$126,212
(2)	Tax and freight at 7% of (1)	8,835
(3)	Installation Costs (use 100% of equip. cost for this example)	126,212
(4)	Subtotal (1) + (2) + (3)	261,259
(5)	Engineering at 10% of (4)	26,126
(6)	Subtotal (4) + (5)	287,385
(7)	Contingencies at 10% of (6)	28,739
(8)	Total Capital Costs (6) + (7)	\$316,124

13. Calculate the uniform annual cost for a fabric filter, assuming again that the present rate of return is 12 percent and the service life is 20 years.

From Table 4-16, the capital recovery factor = 0.13388.

Equivalent Annual Cost = Capital Cost x CRF
= $316,124 x 0.13388
= $42,323

Uniform Annual Cost = Equiv. Annual Cost + O&M Costs
Uniform Annual Cost = $42,323 + $12,713
= $55,036

6.7.4 Process Modification

Low ash coal is available from the supplier. This coal may reduce particulate emissions to the required levels without any further equipment modifications or installations. A random sampling of coal has yielded the following laboratory analyses:

COAL ANALYSIS (As Received)

Test No.	Test No. 738	9416	1453	1403	273	273A	Avg.
Proximate							
Heat Value (Btu/lb)	13,820	13,870	13,860	13,680	13,680	13,930	13,810
Moisture (%)	4.9	4.9	4.6	5.3	5.3	4.6	4.9
Fixed Carbon (%)	57.2	57.5	57.6	57.3	57.1	57.7	57.4
Volatile Matter (%)	35.0	34.8	34.8	34.2	34.4	34.9	34.7
Ash (%)	2.9	2.8	3.0	3.2	3.2	2.8	3.0
Ultimate							
Hydrogen (%)	5.9	6.0	5.7	5.8	6.0	5.7	5.9
Carbon (%)	77.2	77.2	76.8	76.2	74.8	77.8	76.7
Nitrogen (%)	1.9	1.9	1.9	1.9	1.9	1.9	1.9
Sulfur (%)	0.9	0.9	0.8	0.8	0.8	0.8	0.8
Oxygen (%)	11.2	11.2	11.8	12.0	13.4	11.1	11.8

Assuming the same heat input rate from the previous data (42 million Btu/hr), the coal rate is: (42 MMBtu/hr)/(13,810 Btu/lb)
= 3,040 lb/hr
= 1.52 ton/hr.

1. Determine the emission rate of particulate when burning the new coal. Refer to Chapter 1 for emission factors.

   (i) Uncontrolled Particulate Emission
   = 20 lb/ton x 3.0% ash x 1.52 ton/hr
   = 91.2 lb/hr

   (ii) Controlled emission rate (assuming 75% collection efficiency)
   = 91.2 lb/hr x 0.75
   = 22.8 lb/hr
   or (22.8 lb/hr)/(42 MMBtu/hr)
   = 0.54 lb/MMBtu
   For 90% collection efficiency, controlled emission rate is:
   9.12 lb/hr or 0.22 lb/MMBtu

Since the allowable emission rate under these conditions is 0.27 lb/MMBtu, this new coal will be effective as a control measure only if existing multi-cyclone operates at maximum performance. A test firing of the new coal is indicated.

Since the emission rate of sulfur dioxide is in compliance with the original coal, there is no need to calculate the sulfur dioxide emission rate while firing the new coal.

2. Calculate the cost of firing the new coal. The new coal is available at $70/ton, while the old coal is available at $35/ton. The yearly cost of firing the old coal (assuming constant maximum load to keep an equal base for comparison purposes) is:

$35/ton x 1.65 ton/hr x 8,760 hr/year
= $505,890/yr

The yearly cost of firing the new coal is:

$70/ton x 1.52 ton/hr x 8,700 hr/year
= $932,064

The difference then is:

$932,064 - $505,890 = $426,174

6.8 SUMMARY OF CONTROL METHODS

1. Wet Scrubber

   a. Unit is designed for 98 percent collection efficiency. Existing multi-cyclone is left intact, to ensure adequate total collection to meet existing codes under all conditions.

   b. The unit is designed for 3.9 transfer units and 25 inch water gauge pressure drop; total contacting power is 4.026 hp/1,000 acfm; unit is made from 1/8 inch 316 stainless steel to prevent corrosion from wet sulfur dioxide.

   c. Estimated capital equipment cost is $45,473 not including cost of extra fan capacity to boost gas pressure drop. The total capital cost is $113,896.

   d. Estimated operating and maintenance cost is $40,748/year based on continuous operation.

2. Electrostatic Precipitators

   a. Designed for 98 percent collection efficiency, which is adequate to meet existing codes under all conditions, including soot blow. Existing multi-cyclone is not removed in this case, since electrostatic precipitator has low pressure drop.

   b. Unit is designed with 2,600 ft^2 plate area, one bus section, and 2,748 watts. Selected precipitation rate parameter is 0.574 ft/sec.

   c. Estimated equipment cost is $176,750 for an insulated precipitator. Total capital cost is $442,706.

   d. Estimated operating and maintenance cost is $4,246/year based on continuous operation.

3. Baghouse

   a. Operates at 99+ percent collection efficiency; will comply with emission codes even if existing multi-cyclone collector is removed. Removal of multi-cyclone reduces fan load.

   b. Unit with glass fiber bags, reverse-flow cleaning, insulation, and suction operation will cost $126,212 basic equipment cost (no contingencies, taxes, shipping, or installation included).

c. Unit is sized for 22,900 acfm with 13,572 ft^2 of net cloth area.

d. Since the baghouse will collect submicron size particles very efficiently, the unit will provide compliance for all possible future codes, including opacity and inhalable particulate codes.

e. Estimated capital equipment cost is $126,212. Total capital cost is $316,124.

f. The baghouse estimated operating and maintenance cost is $12,713/year based on continuous operation.

4. Process Modification

a. Estimated emissions from the firing of low-ash coal (3.0%) range from 0.22 to 0.54 lb/million Btu, depending upon performance of existing multi-cyclone collector. A test firing of the coal with simultaneous emission tests would be necessary before deciding if the low-ash coal can meet the 0.27 lb/million Btu emission code.

b. Estimated additional costs of the coal are $426,174. No capital equipment costs are incurred.

6.9 SUMMARY--CONTROL ALTERNATIVE COSTS

Control Method	Equipment	Operation and Maintenance
Wet Scrubber	$ 45,473	$ 46,460
Electrostatic Precipitator	$176,750	$ 4,700
Baghouse	$126,212	$ 12,713
Process Modification	$ 0	$426,174

Refer to Chapter 4 for information on tax, freight, engineering, and contingencies. Installation is estimated equal to base equipment costs. These total capital costs are:

Wet Scrubber	$113,896
ESP	$442,706
Baghouse	$316,124

Refer to Chapter 4 for information concerning capital recovery factor. For the wet scrubber a 10-year service life is used at 12 percent interest, and therefore the capital recovery factor is 0.17698. Both the ESP and baghouse can have a service life of 20 years; therefore, the capital recovery factor used is 0.13388 (again at 12 percent interest). The yearly cost of capital equipment is as follows:

Wet Scrubber	$20,157
ESP	$59,269
Baghouse	$42,323

When this cost is added to the yearly operation and maintenance cost, the uniform annual cost is:

Wet Scrubber	$ 66,517
ESP	$ 63,969
Baghouse	$ 55,036
Process Modification	$426,174

6.10 CONCLUSIONS

When just the equipment costs are compared, the wet scrubber is the most attractive. However, when the long-term costs are compared, the alternative baghouse filter is the lowest cost equipment for the application. Not only does the baghouse seem to be the most attractive choice based on capital expenditures, but it also can ensure compliance with all existing and possible future regulations.

The ESP can meet all existing codes, but may not meet possible future regulations, and, of course, is more costly to purchase and operate than the baghouse.

The wet scrubber is the most unattractive solution: not only does it have the highest uniform annual cost, but it also creates a potentially hazardous waste liquid that must be disposed of.

Comparison of Control Methods

	Meets Reg.	Equip. Cost	Oper. Cost	Fut. Reg.	Uniform Annual Cost
Baghouse	Yes	High	Low	Yes	Lowest
ESP	Yes	High	Low	?	High
Wet Scrubber	No	Low	High	No	Highest
Process Change	Yes	--	High	?	

Based on this evaluation, the baghouse could be selected for long-term operation.

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