Section 5
SO_2 and Acid Gas Controls

Chapter 1

Wet and Dry Scrubbers for Acid Gas Control

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April 2021

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1.1 Introduction

Gas absorbers (or scrubbers) are used extensively in industry for separation and purification of gas streams, as product recovery devices, and as pollution control devices. In the petrochemical, pharmaceutical and biotechnology industries, gas absorbers are used for product recovery and purification. When used for air pollution control, gas absorbers are used to remove water soluble contaminants, such as sulfur dioxide (SO_2), acid gases such as hydrogen chloride (HCl), and hazardous air pollutants (e.g., mercury -Hg), from air streams. [l, 2] Gas absorbers are commonly used to control SO_2 emissions from stationary coal- and oil-fired combustion units (e.g., as electric utility and large industrial boilers). They are also used to control emissions from municipal and medical waste incinerators and a wide range of industrial processes, including cement and lime kilns, metal smelters, petroleum refineries, glass furnaces, and sulfuric acid plants. [3]

Gas absorbers are generally referred to as scrubbers due to the mechanisms by which gas absorption takes place. However, the term scrubber is often used very broadly to refer to a wide range of different control devices, such as those used to control particulate matter emissions (e.g., venturi scrubber).[1] In this chapter, the term scrubber is used to refer to control devices that use gas absorption to remove gases from waste gas streams. When used to remove SO_2 from flue gas, gas absorbers are commonly called flue gas desulfurization (FGD) systems; when used to control HCl and other acidic gases, they are called acid gas scrubbers.

This chapter focuses on the application of gas absorption for air pollution control on SO_2 and acid gas streams with typical pollutant concentrations ranging from 250 to 10,000 parts per million by volume (ppmv). Section 1.2 focuses on gas absorbers used to control SO_2 emissions and describes the various types of FGD systems available and presents methods for estimating the capital and operating costs for wet and dry/semi-dry FGD systems.

Section 1.3 focuses on wet packed tower scrubbers and presents a detailed methodology for determining the design parameters and estimating the capital and operating costs for typical packed tower scrubbers. The methods in Section 1.3 can be used to design and estimate costs for packed tower scrubbers used to control acid gases, such as HCl and hydrofluoric acid (HF). However, the methods outlined in Section 1.3 can also be used as an alternative to the methods outlined in Section 1.2.3 for designing and costing packed tower FGD systems.

The cost methodologies presented provide study-level estimates of capital and annual costs, consistent with the accuracy of estimates for other control technologies included in the Control Cost Manual. These methodologies can be used to compare the approximate costs of different scrubber designs. Actual costs may differ from those estimated using these methodologies due to site-specific factors and type of contracting agreements. As with other control technologies included in the Control Cost Manual, where more accurate cost estimates are needed, we recommend capital and operating costs be determined based on detailed design specifications and extensive quotes from suppliers.

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[1] For information on wet scrubbers used to control particulate emissions, including venturi scrubbers, see Section 6, Chapter 2 (Wet Scrubbers for Particulate Matter) of this Manual.

1.1.1 Process Description

Acid gas scrubbers are designed to bring gas mixtures into contact with a sorbent so that one or more soluble components of the gas interacts with the sorbent. The absorption process can be categorized as physical or chemical. Physical absorption occurs when the absorbed compound dissolves in a liquid sorbent; chemical absorption occurs when the absorbed compound and the sorbent react. Sorbents are generally liquid solvents such as water, mineral oils, nonvolatile hydrocarbon oils, and aqueous solutions. However, some scrubbers use dry or semi-dry absorbents. For example, some scrubbers use lime mixed with a small amount of water to create a slurry. [1]

Absorption is a mass transfer operation in which one or more soluble components of a gas mixture are dissolved in a liquid that has low volatility under the process conditions. The pollutant diffuses from the gas into the liquid when the liquid contains less than the equilibrium concentration of the gaseous component. The difference between the actual concentration and the equilibrium concentration provides the driving force for absorption.

A properly designed scrubber will provide thorough contact between the gas and the solvent in order to facilitate diffusion of the pollutant(s). [4] The rate of mass transfer between the two phases is largely dependent on the surface area exposed and the time of contact. Other factors governing the absorption rate, such as the solubility of the gas in the particular solvent and the degree of the chemical reaction, are characteristic of the constituents involved and are relatively independent of the equipment used.

1.1.2 Gas Absorber System Configurations

Scrubbers typically consist of a vertical and cylindrical column or tower in which the solvent is brought in contact with the exhaust gas that contains the pollutant to be removed. Several different designs of absorber towers are used. Commonly used designs include packed-bed scrubbers, spray tower scrubbers, and tray tower scrubbers. Venturi scrubbers may also function as gas absorbers; however, they are usually designed for control of particulates rather than acid gases or SO_2.

Gas and liquid flow through an absorber may be co-current flow, counter-flow, or cross-flow. The most commonly installed designs are countercurrent, in which the waste gas stream enters at the bottom of the absorber column and exits at the top. Conversely, the solvent stream enters at the top and exits at the bottom. Countercurrent designs provide the highest theoretical removal efficiency because gas with the lowest pollutant concentration contacts liquid with the lowest pollutant concentration. This serves to maximize the average driving force for absorption throughout the column.[2] Moreover, countercurrent designs usually require lower liquid-to-gas ratios than co-current and are more suitable when the pollutant loading is higher. [5, 6]

In a crosscurrent tower, the waste gas flows horizontally across the column while the solvent flows vertically down the column. As a rule, crosscurrent designs have lower pressure drops and require lower liquid-to-gas ratios than both co-current and countercurrent designs.

They are applicable when gases are highly soluble since they offer less contact time for absorption. [2, 6]

In co-current towers, both the waste gas and solvent enter the column at the top of the tower and exit at the bottom. Co-current designs have lower pressure drops, are not subject to flooding limitations and are more efficient for fine (i.e., submicron) mist removal. Co-current designs are only efficient where large absorption driving forces are available. Removal efficiency is limited since the gas-liquid system approaches equilibrium at the bottom of the tower.[2]

Gas absorbers can be classified as either “wet” or “dry” scrubbers depending on the physical state of the sorbent. In a wet scrubber, the sorbent is injected into the waste gas stream as an aqueous solution and the pollutants dissolve in the aqueous droplets and/or react with the sorbent. Dry scrubbers inject either dry, powdered sorbent or an aqueous slurry that contains a high concentration of the sorbent. In the latter case, the water evaporates in the high temperature of the flue gas, leaving solid sorbent particles that react with the sorbent. [3] Wet and dry scrubbers are used to control acid gases from combustion and industrial processes. Wet scrubbers usually achieve higher SO_2 removal efficiencies than dry scrubbers, however, dry scrubbers offer several advantages over wet scrubbers. They are generally less expensive, take up less space, are less prone to corrosion, and have lower operating costs than a comparable wet scrubber system. Table 1.1 compares the advantages and disadvantages of wet and dry scrubbers. [62]

Table 1.1: Comparison of Wet and Dry Scrubbers [16]

Scrubber Type	Advantages	Disadvantages
Wet Scrubbers	Higher pollutant removal efficiency than other SO_2 controls (SO_2 removal from flue gas is typically between 90 and 98%; new designs achieve 99% removal).	Higher water usage than comparable dry scrubber.
	Wide range of applications.	Typically requires some type of wastewater treatment.
	Reagent usage generally lower than for comparable dry scrubber.	Power usage generally higher than for comparable dry scrubbers.
	Can handle high temperature waste gas streams.	Requires protection from freezing temperatures resulting in higher capital costs and higher operating costs during winter months.
	High removal efficiencies for acid gases (e.g., HCl, HF, H_2SO_4) in industrial waste streams.	
	Can handle flammable and explosive waste streams.	High corrosion potential (Corrosion resistant alloys and coatings required for absorber and downstream equipment.)
	Gypsum ($CaSO_4$) can be recovered from limestone/lime wet FGD systems.	

Scrubber Type	Advantages	Disadvantages
	Capable of handling flue gas from combustion of high sulfur coals.	May have a visible white plume due to water vapor.
Dry Scrubbers	Lower capital and operating costs.	SO_2 removal efficiency typically lower than wet scrubbers (For spray dry absorbers (SDA), removal efficiencies are typically between 85 - 95%, with newer SDA designs capable of achieving 98%. For circulating dry scrubbers (CDS), removal efficiency is typically >95%, with newer designs capable of achieving up to 98%.)
	High removal efficiencies for acid gases (e.g., HCl, HF, H_2SO_4) in industrial waste streams.	
	Capable of controlling very acidic waste streams.	
	Handles high temperature waste streams.	Disposal of waste can incur higher costs.
	No visible condensation plume as flue gas temperature remains above the dew point.	Potential for clogging and erosion of the injection nozzles due to the abrasive characteristics of the slurry.
	Relatively compact system.	SDA systems are less effective at controlling SO_2 emissions from high-sulfur coal. (SDA systems are typically used to control flue gas emissions from coal with sulfur content <1.5%.)
	Wide range of applications.	
	Lower auxiliary power usage.	
	Less visible plume due to higher operating temperature.	Potential for slurry to buildup on walls of the SDA system and ductwork.
	No dewatering of collected solids.	
	Low water consumption.	
	No wastewater treatment.	Potential for blinded bags if slurry moisture does not evaporate completely.
	Lower corrosion potential for absorber and stack (absorber tower can be fabricated from carbon steel).	

Both wet and dry gas absorbers are commonly used to control SO_2, HCl, HF, HBr (hydrobromic acid), HCN (hydrogen cyanate), HNO_3 (nitric acid), H_2S (hydrogen sulfate), formic acid, chromic acid, and other acidic waste gases from large utility boilers, large industrial boilers, and a wide range of industrial processes. Gas absorbers have been used at refineries, fertilizer manufacturers, chemical plants (e.g., ethylene dichloride production), pulp and paper mills, cement and lime kilns, incinerators, glass furnaces, sulfuric acid plants, plating operations, steel pickling and metal smelters. Waste streams with flow rates ranging from 2,000 actual cubic feet per minute (acfm) to over 100,000 acfm can be treated with acid gas absorbers. Several vendors supply scrubbers of various sizes that are designed for specific industrial applications, such as sulfur recovery units (SRUs), fluidized catalytic cracking units (FCCUs), sulfuric acid production plants, aluminum production, and other non-ferrous metal smelters. These systems typically achieve control efficiencies greater than 98%; however, the removal efficiency achieved can be lower for systems where the waste gas characteristics are variable (e.g., varying acid gas concentrations, flow rates, or temperature). Some systems controlling SO_2 emissions include integrated sulfur recovery systems that produce commercial grade products, such as liquid SO_2, sulfuric acid, and sulfur, that can be used onsite or sold. [3, 60, 61, 68]

1.1.3 Structural Design of Wet Absorption Equipment

Wet gas absorbers can be packed towers, plate (or tray) columns, venturi scrubbers, recirculating fluidized beds, and spray towers.

Packed tower scrubbers are columns filled with packing materials that provide a large surface area to facilitate contact between the liquid and gas. Packed tower scrubbers can achieve higher removal efficiencies, handle higher liquid rates, and have relatively lower water consumption requirements than other types of gas absorbers.[2] Packed towers can be used to remove a wide range of pollutants, including halogens, ammonia, acidic gases, sulfur dioxide, and water soluble organic compounds (e.g., formaldehyde, methanol). However, packed towers may also have high system pressure drops, high clogging and fouling potential, and extensive maintenance costs due to the presence of packing materials. Installation, operation, and wastewater disposal costs may also be higher for packed bed scrubbers than for other absorbers.[2] In addition to pump and fan power requirements and solvent costs, packed towers have operating costs associated with replacing damaged packing.[2]

Plate, or tray, towers are vertical cylinders in which the liquid and gas are contacted in step-wise fashion on trays (plates). Liquid enters at the top of the column and flows across each plate and through a downspout (downcomer) to the plates below. Gas moves upwards through openings in the plates, bubbles into the liquid, and passes to the plate above. Plate towers are easier to clean and tend to handle large temperature fluctuations better than packed towers.[7] However, at high gas flow rates, plate towers exhibit larger pressure drops and have larger liquid holdups. Plate towers are generally made of materials that can withstand the force of the liquid on the plates and also provide corrosion protection such as stainless steel. Packed columns are preferred to plate towers when acids and other corrosive materials are involved because tower construction can then be of fiberglass, polyvinyl chloride, or other less costly, corrosive-resistant materials. Packed towers are also preferred for columns smaller than two feet in diameter and when pressure drop is an important consideration. [5, 8]

Venturi scrubbers employ a gradually converging and then diverging section, called the throat, to clean incoming gaseous streams. Liquid is either introduced to the venturi upstream of the throat or injected directly into the throat where it is atomized by the gaseous stream. Once the liquid is atomized, it collects particles from the gas and discharges from the venturi.[1] The high pressure drop through these systems results in high energy use. They are designed for applications requiring high removal efficiencies of submicron particles, between 0.5 and 5.0

micrometers in diameter.[7] Although they can be used to control other pollutants, the relatively short gas-liquid contact time restricts their application to highly soluble gases. They are infrequently used to control volatile organic compound (VOC) emissions in dilute concentrations.[2]

Spray towers operate by delivering liquid droplets through a spray distribution system. The droplets fall through a countercurrent gas stream under the influence of gravity and contact the pollutant(s) in the gas.[8] Spray towers are simple to operate and maintain and have relatively low energy requirements. However, they have the least effective mass transfer capability of the absorbers discussed and are usually restricted to particulate removal and control of highly soluble gases, such as SO_2 and ammonia. They also require higher water recirculation rates and are inefficient at removing very small particles. [2, 6]

In circulating fluidized bed systems, the waste gas stream is passed through a reactor vessel containing a circulating fluidized bed of absorbent. There are a variety of different designs available and they can be used to control acidic gases from production processes, as well as for flue gas desulfurization. For certain industrial applications, wet scrubbers may use water to absorb acids, such as HCl and H_2SO_4, resulting in wastewater comprising a weak acid solution that may be recovered for use elsewhere in the plant or sold as a by-product. However, scrubber efficiency is significantly improved if a strong alkali solution is used, such as sodium hydroxide (NaOH), sodium carbonate, calcium hydroxide, and magnesium hydroxide. For combustion sources, a lime or limestone slurry is typically used.

1.1.4 Factors Affecting the Performance

Wet scrubbers are used for a wide range of applications and typically achieve very high levels of pollutant removal. The scrubber design selected depends on the application. Spray towers are generally used in applications where the waste stream contains particulates, such as controlling SO_2 emissions in flue gas from coal-fired boilers and HF emissions from aluminum production. Packed bed and tray towers are used to control HF, HCl, HBr, fluorine (F_2), chlorine (Cl_2), and SO_2 from incinerators, chemical processes, plating, and steel pickling. Wet scrubbers typically achieve removal efficiencies of between 95 and 99% for most industrial applications. For some industrial applications, two or more absorber vessels arranged in series and using different scrubbing solutions can be used to achieve high removal efficiencies for waste gases that contain multiple pollutants. [66, 67, 69] The suitability of gas absorption as a pollution control method is generally dependent on the following factors:

- availability of suitable solvent;
- required removal efficiency;
- pollutant concentration in the inlet vapor;
- volumetric flow rate of the exhaust gas stream; and,
- recovery value of the pollutant(s) or the disposal cost of the spent solvent. [7]

Physical absorption depends on properties of the gas stream and solvent. Properties such as the density and viscosity of the solvent, as well as specific characteristics of the pollutant(s) in the gas and the liquid stream (e.g., diffusivity, equilibrium solubility) impact the performance of the absorber. These properties are temperature dependent. Lower temperatures generally favor absorption of gases by the solvent. [1]

The solvent chosen to remove the pollutant(s) should have a high solubility for the gas, low vapor pressure, low viscosity, and should be relatively inexpensive. [7] Water is the most common solvent used to remove inorganic contaminants; it is also used to absorb organic compounds having relatively high solubility in water. For organic compounds that have low water solubility, other solvents such as hydrocarbon oils are used, though only in industries where large volumes of these oils are available (i.e., petroleum refineries and petrochemical plants).[6]

Absorption is also enhanced by greater contacting surface, higher liquid-to-gas ratios, and higher concentrations of the pollutant in the gas stream. [1] Removal efficiency is also dependent on the recirculation rate and addition of fresh sorbent. Pollutant removal may be enhanced by manipulating the chemistry of the absorbing solution so that it reacts with the pollutant(s), e.g., caustic solution for acid-gas absorption vs. pure water as a solvent. Chemical absorption may be limited by the rate of reaction, although the rate-limiting step is typically the physical absorption rate, not the chemical reaction rate.

1.1.5 Structural Design of Dry Absorption Equipment

There are two types of dry scrubbers: spray dryer absorber (SDA) and circulating dry scrubber (CDS). Both types of dry scrubber systems consist of an absorber vessel, a bag house filter, an absorbent feeding tank, and an absorbent feeding system. Absorbents such as lime and sodium bicarbonate are often used. [68]

Like wet scrubbers, SDA systems have been used to control acid gases since the early 1970s and work by spraying a small amount of slurry into an absorber vessel. The SDA vessel is generally constructed from mild steel and uses a high-speed rotary or dual fluid atomizer to spray the absorbent slurry. At the high operating temperatures of the SDA absorber tower, the water is rapidly vaporized and exits the stack as a slightly visible plume. The absorbent reacts with the acidic gases in the waste stream to form a byproduct that is collected in a fabric filter located at the outlet of the absorber tower. [62, 63, 65]

The CDS system has been used since the mid-1990s and works by circulating dry solids consisting of dry absorbent (typically hydrated lime), air and reaction products through the absorber vessel and fabric filter. A small amount of water is used to promote the reaction of the solids with the waste gas constituents, but unlike the SDA, the water is injected separately from the dry absorbent powder in the reactor. The water is vaporized in the absorber vessel and the dry flue gas exits the absorber vessel and passes through a fabric filter that captures the solids produced. The water spray also controls the absorber temperature, while the separately controlled injection of hydrated lime controls the outlet emissions. One advantage of the CDS system is its ability to be easily adjusted to variable pollutant loads because of the separate control of the absorbent and water injection rates. Another advantage of the CDS system is that it provides an opportunity for the absorbent particles to react more than once. For example, in a CDS system used to control SO_2 emissions, the hydrated lime particle initially reacts to form a layer of calcium sulfite on its surface. On subsequent passes through the lime recirculation system, fresh hydrated lime is exposed as the surface crystals grow when they encounter water. [62, 63, 64]

Unlike wet scrubbers, the SDA and CDS systems do not produce a saleable byproduct (i.e., gypsum from wet limestone FGD systems) when used to control SO_2 emissions from combustion sources. However, the solid wastes collected do not have to have water removed (dewatered) before transfer and disposal in landfills. [63]

1.1.6 Equipment Life

Acid gas scrubbers are relatively reliable systems that have been demonstrated to be exceedingly durable. In the past, the EPA has generally used equipment life estimates of 20 to 30 years for analyses involving acid gas scrubbers, although these estimates are recognized to be low for many installations. Many FGD systems installed in the 1970s and 1980s have operated for more than 30 years (e.g., Coyote Station; H.L. Spurlock Unit 2 in Maysville, KY; East Bend Unit 2 in Union, KY; and Laramie River Unit 3 in Wheatland, WY) and some scrubbers may have lifetimes that are much longer. Manufacturers reportedly design scrubbers to be as durable as boilers, which are generally designed to operate for more than 60 years. [9, 10, 11, 12, 13]

1.2 Flue Gas Desulfurization Systems

FGD systems were first developed in Europe in the 1930s and were first installed on power plants in the United States in the 1970s. In 2019, approximately 250 coal-fired units at U.S. electric power plants were equipped with FGD systems, which accounted for about 52% of all coal-fired generators and 64% of all coal-fired electric generating capacity (approximately 144,000 megawatts (MW)). FGD systems control emissions using alkaline reagents that absorb and react with SO_2 in the waste gas stream to produce salts. Most FGD systems use limestone or lime, although systems using sodium-based alkaline reagents are also available. [3, 9, 14, 15, 16]

1.2.1 Types of FGD Systems

FGD systems are characterized as either “wet” or “dry” corresponding to the phase in which the flue gas reactions take place. Four types of FGD systems are currently available:

- Wet FGD systems use a liquid absorbent.
- Spray Dry Absorbers (SDA) are semi-dry systems in which a small amount of water is mixed with the sorbent.
- Circulating Dry Scrubbers (CDS) are either dry or semi-dry systems.
- Dry Sorbent Injection (DSI) injects dry sorbent directly into the furnace or into the ductwork following the furnace.

The number of wet and dry FGD systems installed in the U.S. are shown in Table 1.2.

Table 1.2: Number of Wet and Dry Flue Gas Desulfurization Systems Installed at U.S. Power Plants in 2019 [16]

Type of FGD Control System	**Number**
Wet FGD Systems	
Wet Limestone	180
Wet Lime	108
Sodium-Based Wet FGD	11
Dual Alkali	5
Dry FGD Systems	
Spray Dry Lime or Semi-Dry Lime (SDA)	105
Fluidized Bed Limestone Injection (FBLI)	39
Dry Sorbent Injection (DSI)	14
Other	2

About 170,000 MW of the U.S. electric generating capacity are controlled using wet scrubbers, while dry scrubbers account for only about 30,000 MW capacity. Most FGD systems were installed as retrofits to existing power plants. Limestone and lime are the most common sorbents used, although other sorbents can also be used. In recent years, the number of FGD systems operated at U.S. power plants has declined due to closure of coal-fired plants. Between 2018 and 2019, the number of operating wet FGD systems decreased by 37, while the number of dry FGD systems decreased by 4. [3, 9, 14, 56]

1.2.1.1 Wet Flue Gas Desulfurization Systems

Wet FGD systems control SO_2 emissions using solutions containing alkali reagents. Wet FGD systems may use limestone, lime, sodium-based alkaline, or dual alkali-based sorbents. Wet FGD systems can also be categorized as "once-through" or "regenerable" depending on how the waste solids generated are handled. In a once-through system the spent sorbent is disposed as waste. Regenerable systems recycle the sorbent back into the system and recover the salts for sale as byproduct (e.g., gypsum), and have higher capital costs than once-through systems due to the additional equipment required to separate and dry the recovered salts. However, regenerable systems may be the best option for plants where disposal options are limited or nearby markets for byproducts are available. [3, 14, 15, 17]

Most wet FGD systems use a limestone slurry sorbent which reacts with the SO_2 and falls to the bottom of the absorber tower where it is collected. Wet FGD systems generally have the highest control efficiencies. New wet FGD systems can achieve SO_2 removal of 99% and HCl removal of over 95%. Packed tower wet FGD systems may achieve efficiencies over 99% for some pollutant-solvent systems. However, packed tower wet FGD systems are not widely used due to the potential for deposits of calcium sulfate and calcium chloride on the packing materials. [1, 5, 15]

The wet lime FGD system uses hydrated lime, instead of limestone, in a countercurrent spray tower. The lime is shipped to the plant as quicklime and hydrated to form the lime slurry using a wet ball mill.

In the sodium-based wet scrubbing process (e.g., Wellman-Lord process), a regenerable process, SO_2 is absorbed in a sodium sulfite solution in water forming sodium bisulfite which precipitates. Upon heating, the chemical reactions are reversed, and sodium pyrosulfite is converted to a concentrated stream of sulfur dioxide and sodium sulfite. The sulfur dioxide can be used for further reactions (e.g., the production of sulfuric acid), and the sulfite is reintroduced into the process.

A dual alkali scrubber uses an indirect lime process for removing acid gas with a sodium-based absorbent. The sodium absorbent is regenerated through reaction with lime in a secondary water recycle unit. Calcium sulfite/sulfate is precipitated and discarded. Water and sodium ions are recycled back to the dual alkali scrubber. The system has zero liquid discharge. The precipitated calcium sulfite/sulfate is typically sent to a landfill.

For information on the typical level of SO_2 removal achieved by each type of FGD system and the annual average SO_2 emissions rates based on 2019 data, see Section 1.2.2, Performance of Wet and Dry FGD Systems, later on this chapter.

One benefit of wet FGD systems is their ability to also reduce mercury emissions from coal combustion by dissolving soluble mercury compounds (e.g., mercuric chloride). The level of mercury reduction depends on the mercury speciation, as flue gas from coal combustion contains varying percentages of three mercury species: particulate-bound, oxidized (Hg^{2+}), and elemental. The Hg^{2+} species is the only soluble form. Consequently, wet FGD systems are more effective at reducing mercury emissions where the fraction of Hg^{2+} in the waste gas stream is higher. The fraction of Hg^{2+} is generally higher in coal containing higher levels of chlorine, such as bituminous coal. Facilities may enhance mercury oxidation by directly injecting bromide or other halogens during combustion, mixing bromide with coal to produce refined coal; or using brominated activated carbon. Wet FGD systems that are used to control mercury as well as SO_2 generally have higher operating expenses due to costs for additives and additional monitoring of the oxidation/reduction potential necessary to optimize mercury removal. The control of mercury from coal combustion is complex due to mercury speciation and is generally achieved using a combination of air pollution control techniques that is beyond the scope of this chapter. [49, 50, 51]

1.2.1.2 Dry Flue Gas Desulfurization Systems

Dry Lime FGD systems are also called SDA (sometimes called Semi-Dry Absorbers) and are gas absorbers in which a small amount of water is mixed with the sorbent. Lime (CaO) is usually the sorbent used in the spray drying process, but hydrated lime ($Ca(OH)_2$) is also used and can provide greater SO_2 removal. Slurry consisting of lime and recycled solids is atomized/sprayed into the absorber. The SO_2 in the flue gas is absorbed into the slurry and reacts with the lime and fly ash alkali to form calcium salts. The scrubbed gas then passes through a particulate control downstream of the spray drier where additional reactions and SO_2 removal may occur, especially in the filter cake of a fabric filter (baghouse). Spray dryers can achieve SO_2 removal efficiencies up to 95%, depending on the type of coal burned.

A second type of dry scrubbing system is the CDS, which can achieve over 98% reduction in SO_2 and other acid gases. Similar to other dry flue gas desulfurization systems, the CDS system is located after the air preheater, and byproducts from the system are collected in an integrated fabric filter. Unlike the SDA systems, a CDS system is considered a circulating fluidized bed of hydrated lime reagent to remove SO_2 rather than an atomized lime slurry; however, similar chemical reaction kinetics are used in the SO_2 removal process. In a CDS system, flue gas is treated in a Dry Lime FGD system in which the waste gas stream passes through an absorber vessel where the flue gas stream flows through a fluidized bed of hydrated lime and recycled byproduct. Water is injected into the absorber through a venturi located at the base of the absorber for temperature control. Flue gas velocity through the vessel is maintained to keep the fluidized bed of particles suspended in the absorber. Water sprayed into the absorber cools the flue gas from approximately 300^{o}F at the inlet to the scrubber to approximately 160^{o}F at the outlet of the fabric filter. The hydrated lime absorbs SO_2 from the gas and forms calcium sulfite and calcium sulfate solids. The desulfurized flue gas passing out of the absorber contains solid sorbent mixed with the particulate matter, including reaction products, unreacted hydrated lime, calcium carbonate, and fly ash. The solid sorbent and particulate matter are collected by the fabric filter.

In general, dry scrubbers have lower capital and operating costs than wet scrubbers because dry scrubbers are generally simpler, consume less water and require less waste processing. Data reported to the U.S. Department of Energy, Energy Information Administration (EIA) by power plants in 2018 show that the total installed costs for a wet FGD system with a removal efficiency of 90% or greater range between $250,000 and $709 million, with an average installed cost of $114 million. For the SDA, the average installed cost achieving greater than 90% sulfur removal was reported to be $37 million, with the highest total installed costs reported to be $340 million. For the CDS system, the average installed cost for a CDS system capable of achieving greater than 90% sulfur removal was $81 million and the highest total installed costs reported to be $400 million. [18]

Although more expensive than dry scrubbers, wet scrubbers have higher pollutant removal efficiencies than dry scrubbers. [3] Wet scrubbers can also help remove other acid gases and inorganic HAP (e.g., mercury, sulfur trioxide, HCl and HF). [9] For information on the typical level of SO_2 removal achieved by each type of FGD system and the annual average SO_2 emissions rates based on 2019 data, see Section 1.2.2, Performance of Wet and Dry FGD Systems.

1.2.1.3 Other Designs

Unlike the three other FGD systems, dry sorbent injection (DSI) is not a typical stand-alone, add-on air pollution control system but a modification to the combustion unit or ductwork. DSI can typically achieve SO_2 control efficiencies ranging from 50 to 70% and has been used in power plants, biomass boilers, and industrial applications (e.g., metallurgical industries).

Fluidized Bed Limestone Injection (FBLI) is a boiler design in which fuel particles are suspended in a hot, bubbling fluid bed of ash and other particulate materials (sand, limestone etc.) through which jets of air are blown to provide the oxygen required for combustion or gasification. Limestone is used to precipitate out sulfate during combustion, which reduces SO_2 emissions. An FBLI system is installed on a coal and biomass-fired boiler at the University of Alaska in Fairbanks, AK.

DSI and FBLI are not covered in this chapter. This scrubbers chapter focuses on FGD systems that are stand-alone devices that can be installed on combustion units.

1.2.2 Performance of Wet and Dry FGD Systems

Table 1.3 summarizes the efficiency and SO_2 emission rates for FGD systems based on 2019 data for coal-fired units at power plants. The performance of FGD systems installed on power plants has improved over the last 20 years and many vendors have published SO_2 removal efficiencies of over 99% for new wet FGD systems, up to 95% for new SDA systems, and up to 98% for new dry FGD systems. [70, 71, 72, 73, 74 and 75] Figure 1.1 shows the 12-month average emission rate for the top performing 50% and top performing 20% of wet limestone, wet lime and dry lime gas absorbers in 2000, 2005, 2015 and 2019. The average SO_2 emission rate for the top performing 50% of wet limestone FGD systems dropped from 0.22 pounds SO_2 per million British Thermal Unit (lb/MMBtu) in 2000 to 0.04 lb/MMBtu in 2019. Similarly, the top performing 50% of wet lime FGD systems dropped from 0.23 lb/MMBtu in 2000 to 0.07 lb/MMBtu in 2019. Finally, the top performing 50% of dry lime FGD systems dropped from 0.14 lb/MMBtu in 2000 to 0.06 lb/MMBtu in 2019. The decrease in SO_2 emission rates is likely attributable to a variety of factors including improvements in the design and operation of FGD systems and operational changes at some utilities from switching to lower sulfur coal and operating at less than full capacity. Switching to lower operating capacity extends the residence time and results in a higher liquid-to-gas ratio, which increases SO_2 removal. [19]

Table 1.3: 12- Month Average SO_2 Emission Rates By FGD for 2019 Coal-Fired Power Plants [15, 16, 70, 71, 72, 73, 74 and 75]

FGD Type	Sorbent Type	SO_2 Removal Efficiency (%)	12-Month Average SO_2 Emissions Rates (lb/MMBtu)			
			Range	Average		
				For All Plants	Top 20 Percent	Top 50 Percent
Wet FGD	Limestone	92 – 99	0.001 – 1.92	0.13	0.02	0.04
	Lime	95 – 99	0.01 – 0.64	0.14	0.04	0.07
	Sodium-Based Alkali	90 – 95	0.07 – 0.38	0.16	0.07	0.11
	Dual Alkali	≤98	0.02 – 0.27	0.13	0.02	0.06
Spray Dry Absorber (SDA)	Lime	85 – 95	0.04 – 0.86	0.14	0.04	0.06
Circulating Dry Scrubber (CDS)	Lime	95 – 98	0.01 – 0.66	0.25	0.07	0.12

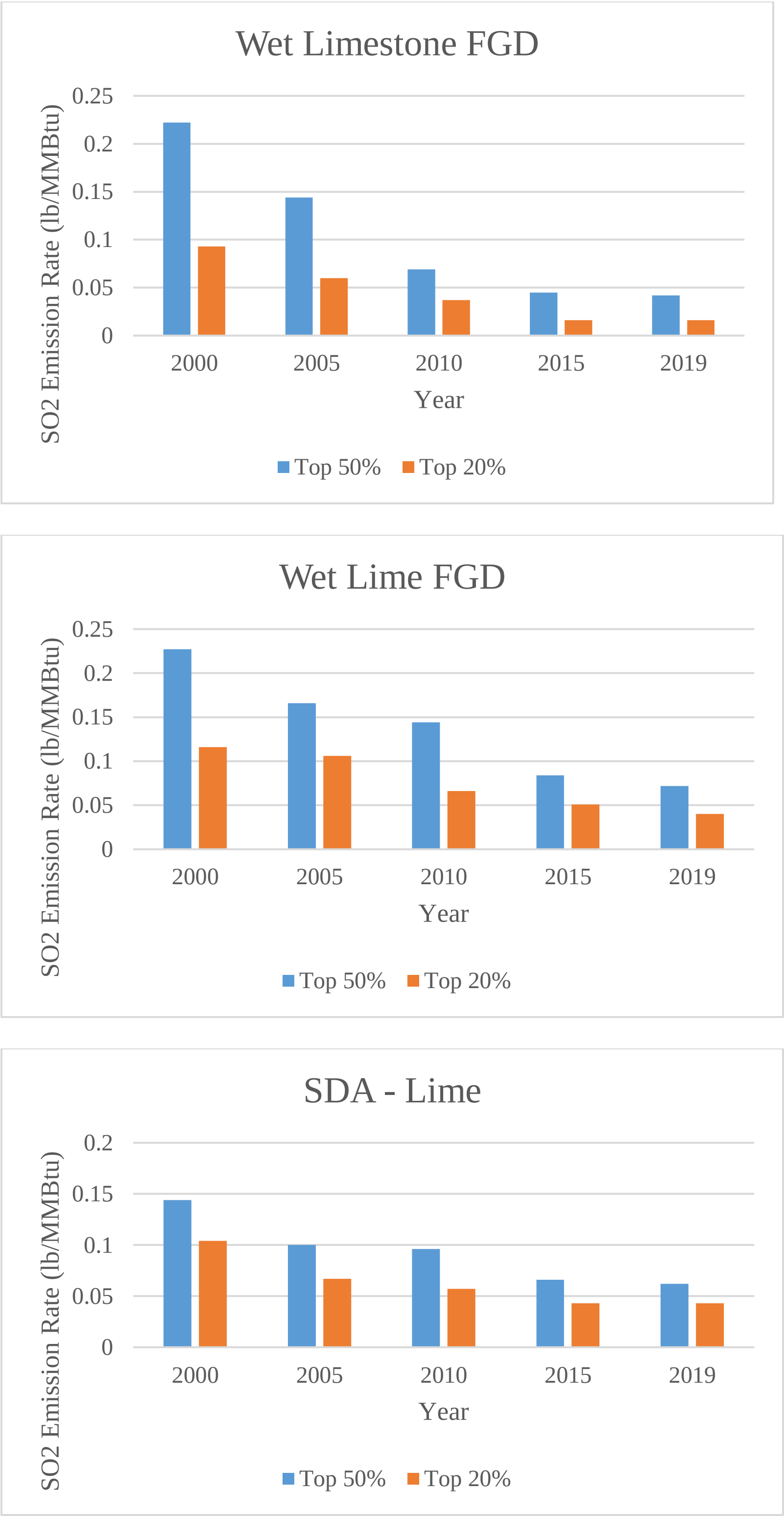

Figure 1.1: Changes in the 12-Month Average SO_2 Emission Rate for the Best Performing Wet Limestone, Wet Lime and Dry Lime Absorbers Installed on Coal-Fired Units at U.S. Power Plants: 2000 through 2019 [19]

Wet and dry FGD systems are described in Sections 1.2.4 and 1.2.5, respectively. Each section provides a general description of the technology, methods for determining the design parameters and equations for estimating the capital and operating costs. In Section 1.2.3, we describe some general concepts that are used later in Sections 1.2.4 and 1.2.5 for determining design parameters and developing cost estimates for wet FGD and SDA systems.

1.2.3 General Parameters used to Develop Scrubber Design Parameters and Cost Estimates for FGD Systems

1.2.3.1 Boiler Heat Input

The maximum potential heat released by the boiler or heat input rate, Q_B, is calculated from the higher heating value, *HHV*, multiplied by the maximum fuel consumption rate in pounds of fuel burned per hour:

$$Q_B = HHV \times q_{fuel} \times \frac{1}{10^6} \tag{1.1}$$

Where:

Q_B = maximum heat rate input to the boiler, MMBtu/hr
HHV = higher heating value of the fuel, Btu/lb
q_{fuel} = maximum fuel consumption rate of the boiler, lb/hr
$1/10^6$ = conversion factor of 1 MMBtu/10^6 Btu.

Table 1.4 provides typical HHVs for various coals.

Table 1.4: Higher Heating Values for Various Coals

Type of Coal	Energy Content (Btu/lb)
Lignite	5,000–7,500
Subbituminous	8,000–10,000
Bituminous	11,000–15,000
Anthracite	14,000

1.2.3.2 Capacity Factor

The total system capacity factor, CF_{total}, is a measure of the average annual use of the combustion unit in conjunction with its control system and is given by:

$$CF_{\text{total}} = \text{CF}_{\text{plant}} \times \text{CF}_{APCD} \tag{1.3}$$

Where:

CF_{total} = total system capacity factor

CF_{plant} = combustion unit capacity factor, which is the ratio of the actual quantity of fuel burned annually to the potential maximum quantity of fuel burned

CF_{APCD} = air pollution control device capacity factor, which is the ratio of the actual days the control device is operated to the total number of days the plant (or boiler) operated during the year (t_{APCD} and t_{plant}).

$$CF_{APCD} = \frac{t_{APCD}}{t_{plant}} \tag{1.4}$$

Where:

t_{APCD} = actual days the air pollution control device is operated annually, days/year

t_{plant} = number of days the plant (or boiler) operated in a year, days.

The effective operating time per year, t_{op}, is estimated in hours using the system capacity factor, CF_{total}:

$$t_{op} = CF_{total} \times 8{,}760 \ \frac{hours}{year} \tag{1.5}$$

1.2.3.3 Heat Rate Factor

The heat rate factor (*HRF*) is the ratio of gross heat rate of the boiler in MMBtu/MWh divided by 10 MMBtu/MWh. The cost methodology for the wet FGD presented in Section 1.2.3 and the SDA FGD presented in Section 1.2.4 use the *HRF* in the equations for the capital cost estimates to account for observed differences in actual costs for different coal-fired combustion units.

$$HRF = \frac{HR}{10} \tag{1.6}$$

Where:

HRF = Heat rate factor

HR = Gross plant heat rate of the system, MMBtu/MWh

10 = Heat rate basis of the SDA FGD and wet FGD base module capital costs, in MMBtu/MWh.

1.2.3.4 Site Elevation Factor

The original cost calculations presented in Sections 1.2.3 and 1.2.4 for wet and dry FGD systems were developed based on data for plants located within 500 feet of sea level. However, for these two systems, the capital costs for the absorber (ABS_{Cost}) and the balance of plant costs (BOP_{Cost}) are impacted by the plant site's elevation with respect to sea level. Therefore, the base cost equations for ABS_{Cost} and BOP_{Cost} are adjusted by elevation factor, *ELEVF*, for the effects of elevation on the flue gas volumes encountered at elevations above 500 feet above sea level. The elevation factor, *ELEVF,* is calculated using Equations 1.7 and 1.8 for plants located above 500 feet. For plants located at elevations below 500 feet elevation, an elevation factor of 1 should be used to calculate the ABS_{Cost} and BOP_{Cost} [20, 21]

$$ELEVF = \frac{P_0}{P_{ELEV}} \tag{1.7}$$

Where:

$ELEVF$ = elevation factor
P_0 = atmospheric pressure at sea level, 14.7 pounds per square inch absolute (psia)
P_{ELEV} = atmospheric pressure at elevation of the unit, psia.

The P_{ELEV} can be calculated using Equation 1.8 [22]:

$$P_{ELEV} = 2116 \times \left[\frac{59-(0.00356 \times h)+459.7}{518.6}\right]^{5.256} \times \frac{1}{144} \tag{1.8}$$

Where:

P_{ELEV} = atmospheric pressure at elevation of the unit, psia
h = altitude, feet.

1.2.3.5 Retrofit Factor

Equipment and installation costs for FGD systems can vary significantly from site-to-site depending on site characteristics. For this reason, the capital cost equations in Sections 1.2.4 for wet FGD systems and 1.2.5 for SDA systems include a retrofit factor (RF) that allows users to adjust the costs estimates, depending on the site-specific conditions and level of difficulty. An RF of 1 should be used to estimate costs for a project of average difficulty. For retrofits that are more complicated than average, a retrofit factor of greater than 1 can be used to estimate capital costs provided the reasons for using a higher retrofit factor are appropriate and fully documented. Similarly, new construction and retrofits of existing plants that are less complicated should use an RF less than 1. Each project should be evaluated to determine the appropriate value for RF. The capital costs for the control systems are calculated for multiple modules and then totaled. The cost for each module is calculated using a separate equation with its own RF. Thus, depending on the site-specific circumstances, different RFs may be used for different modules.

Factors that should be considered when evaluating the RF for retrofits include site congestion, site access, and capacity of existing infrastructure. The amount of space available near the utility boiler can significantly impact the costs. Site congestion can be caused by existing generators, conveyors, and environmental control equipment. Costs will be higher if portions of the FGD system must be elevated or existing generators, control equipment, buildings, and other infrastructure must be relocated to accommodate the FGD system. In some cases, site congestion results in the FGD system being installed further away from the boiler, resulting in extra costs for additional duct work and fans. Costs can also be impacted if ancillary equipment, such as wastewater treatment and absorbent storage and preparation areas, must be located further from the absorber. A congested site can also increase construction costs by requiring greater reliance on cranes to locate equipment and limiting access by construction workers.

Site access can impact installation costs if it is difficult for cranes and other heavy construction equipment to access the construction site. A retrofit for a plant that is bound on two or more sides by adjacent generating units, roadways, rivers, wetlands, or other barriers will likely be more challenging than sites that are more open for equipment access.

The capacity, condition, and location of existing infrastructure can also impact costs. Costs will be lower if existing equipment can be used. For example, new fans may not be needed where existing fans have adequate design margins for handling gas flow. If the site has an existing FGD system, the existing ancillary units for absorbent storage and preparation equipment may be sufficient to support the new system. Similarly, if two FGD systems are planned to be installed at the same plant, costs may be reduced by installing a single wastewater treatment system capable of treating FGD wastewater for both absorbers.

Based on the information available at the time this chapter was prepared, the RF value should be between 0.7 and 1.3 for wet FGD systems and between 0.8 and 1.5 for dry FGD systems, depending on the level of difficulty. Costs for new construction are typically, though not for every instance, 20 to 30 percent less than for average retrofits for units of the same size and design. An RF of 0.77 is recommended for estimating capital costs for new construction. [3, 52, 53]

1.2.4 Wet Flue Gas Desulfurization

1.2.4.1 Wet FGD Systems Process Description

The typical wet FGD system controls waste gases containing SO_2 concentrations up to 2000 ppm using relatively inexpensive and readily available alkali reagents. A typical wet FGD system consists of sorbent storage and preparation equipment, an absorber vessel, a mist eliminator, and waste collection and treatment vessels. An additional or upgraded induced draft fan may also be needed to compensate for the pressure drop across the absorber. Figure 1.2 shows a schematic of a wet FGD system using limestone as the sorbent. Most wet FGD systems use a limestone sorbent that is prepared by first crushing the limestone into a fine powder using a ball mill and then mixing the powder with water in the slurry preparation tank. Particle size of the limestone impacts the efficiency of SO_2 removal. In the typical wet limestone FGD system, the limestone is ground to an average size of 5 to 20 µm. [15]

The sorbent slurry is pumped from the slurry preparation tank to the absorber. The flue gas is ducted to the absorber where the aqueous slurry of sorbent is injected into the flue gas stream through injection nozzles.

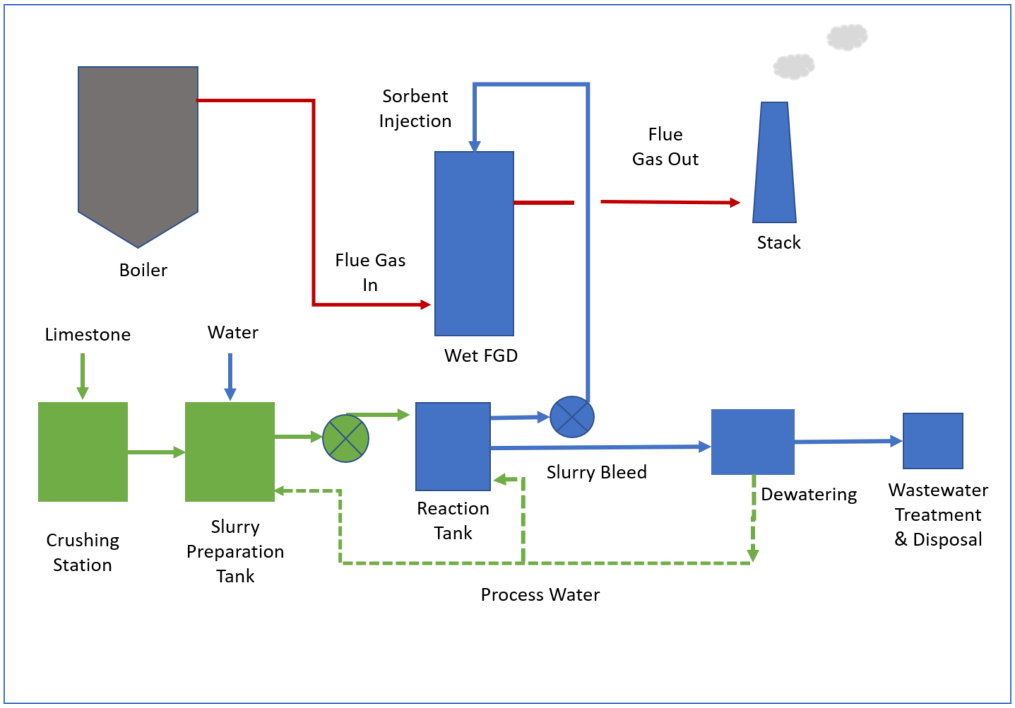

Figure 1.2: Flow Diagram for a Wet FGD System [14]

Although there are several different designs of absorbers, the most common is a counter-flow, vertically oriented spray tower, where the flue gas flows upwards and the sorbent slurry is sprayed downwards. The flue gas inlet temperature is typically between 300 and 700°F. The design and location of the injection nozzles in the spray tower are selected to ensure the sorbent is evenly dispersed in the flue gas stream and that contact between the flue gas and sorbent is optimized. The flue gas becomes saturated with water vapor as a portion of the water in the sorbent slurry is evaporated and the sorbent becomes fine droplets. SO_2, and other acid gases dissolve in the slurry and react with the sorbent to produce salts. The sorbent slurry falls to the bottom of the absorber tower where it is collected and transferred to a waste handling system. [3, 14] A mist eliminator located immediately after the absorber removes any remaining slurry droplets from the flue gas stream. The mist eliminator may be followed by a heat exchanger to raise the temperature of the flue gas. The combined effect of the mist eliminator and heat exchanger reduces opacity in the stack plume and reduces the potential for corrosion of downstream equipment and ducts. [14] Due to the acidic properties of flue gas, corrosion is a significant issue for FGD systems. Methods for mitigating corrosion are discussed in more detail later in this section (see the section titled *Potential Operating Issues for Wet FGD Systems-Corrosion*).

The waste from the absorber (called the slurry bleed) is collected, dewatered, and transferred to an onsite or offsite disposal. Alternatively, the calcium sulfate may be recovered and sold to wallboard manufacturers. Where calcium sulfate is recovered and sold to wallboard manufacturers, the solids are typically dewatered with a vacuum belt and the liquid returned to the FGD system as reclaimed water. [3, 14]

The wet FGD system may be a “regenerable” system where the sorbent is recovered from the waste slurry and recycled back to the absorber, or “once-through” system where the waste slurry is sent to a landfill for disposal or sold as byproduct. The majority of wet FGD systems are once-through systems. [3, 23]

For systems using limestone, the overall reactions are summarized by the following equations:

$$SO_2 + CaCO_3 + 1/2H_2O \rightarrow CaSO_3 \cdot 1/2H_2O + CO_2$$

$$SO_2 + 1/2O_2 + CaCO_3 + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O + CO_2$$

The calcium carbonate ($CaCO_3$) in limestone is only slightly soluble in water, but in the presence of acid, it reacts much more vigorously. It is the acid generated by absorption of the SO_2 into the liquid that drives the limestone dissolution process. [24]

The degree to which the calcium sulfite is converted to calcium sulfate depends on the amount of oxygen present in the absorber and reaction vessel. High levels of oxygen result in higher yields of calcium sulfate (gypsum), which can be sold as a byproduct. [14] For example, gypsum can be sold to cement plants or wallboard manufacturers. [9] The slurry is filtered to recover the gypsum. The gypsum is then dewatered before shipping offsite. The water from filtration is typically sent to a wastewater treatment plant or recirculated back into the FGD system. The typical wastewater treatment consists of increasing the pH to precipitate metals as metal hydroxides. Additives may be added to promote coagulation and flocculation. [15]

In addition to the basic system described above, there are four design variations on the wet FGD system:

1. Limestone Slurry Forced Oxidation (LSFO):

Conventional wet limestone FGDs operate under natural oxidation where $CaSO_3$ is only partly oxidized by the oxygen contained in the flue gas. The product produced is a 50% to 60% mixture of $CaSO_3 \cdot 1/2H_2O$ and $CaSO_4 \cdot 2H_2O$, which is a soft, difficult-to-dewater material that has little practical value.

Many existing scrubbers are equipped with forced-air oxidation systems in which air is injected in the reaction tank and the oxygen converts $Ca(HSO_3)_2$ to $CaSO_4$. The result is a product that contains 90% $CaSO_4 \cdot 2H_2O$. Dewatering of the filtered product is also easier because the gypsum crystals are larger (<100 micrometers or microns, µm) than those produced by natural oxidation in a conventional wet limestone FGD system (<5 µm). [14, 15] Seed crystals of calcium sulfate are used to promote rapid precipitation of the calcium sulfate, which is then dewatered. The air can be blown into either the reaction tank or into an additional holding tank. [14] Sulfate precipitates with calcium as gypsum, which typically forms a cake-like material when subjected to vacuum filtration.

Capital costs are higher than those for the basic wet FGD system as additional compressors, blowers, and piping are required. [14] However, modern scrubber systems can produce gypsum “cake” with as little as 10% moisture (water content by mass), which makes it very easy to handle or sell to wallboard manufacturers.[24] For example, in 2018, the Centralia power plant reported 0.04 – 0.05 lb/mmBtu annual SO_2 emissions from its tangentially-fired boilers and all gypsum produced by the scrubber is recycled in the production of wallboard.

2. Limestone Inhibited Oxidation (LSIO):

Sulfur or sodium thiosulfate is added to the limestone slurry to inhibit the oxidation of calcium sulfite to calcium sulfate (gypsum). LSIO systems can be used to control waste gas with higher SO_2 concentrations than systems using only limestone. The waste $CaSO_3$ produced by the LSIO system must be landfilled. [3, 14]

3. Lime Process:

This process uses calcium hydroxide (lime slurry) instead of the limestone slurry in a countercurrent spray tower. The hydrated lime reacts more readily with the dissolved SO_2 than does the limestone slurry. Hence, less lime is needed to achieve the same level of SO_2 removal. However, the lime process is generally more expensive to operate than a limestone-based system due to the higher cost of the lime. The lime process can also use fly ash to increase the alkalinity of the sorbent slurry. [3, 14]

For systems using lime, the overall reactions are summarized by the following equations:

$$CaO + H_2O \rightarrow Ca(OH)_2 SO_2 + H_2O \leftrightarrows H_2SO_3$$

$$H_2SO_3 + Ca(OH)_2 \rightarrow CaSO_3 + 2H_2O$$

$$CaSO_3 + 1/2O_2 \rightarrow CaSO_4$$

4. Magnesium Enhanced-Lime Process (MEL):

MEL uses either magnesium-enhanced lime that contains approximately 5% to 8% magnesium oxide (MgO) or dolomite lime that contains about 20% magnesium oxide. MEL achieves higher SO_2 removal efficiencies and uses less sorbent. The MgO in the lime reacts with SO_2 to form soluble magnesium sulfite ($MgSO_3$). The $MgSO_3$ accumulates in the lime slurry and acts as a pH buffer, which increases the mass transfer of SO_2 resulting in SO_2 removal of greater than 95% at low liquid-to-gas ratio. For example, one system achieved 96% SO_2 removal with an L/G ratio of 21 gpm per 1000 acfm. [25]

For systems using MEL, the overall reactions for magnesium oxide are summarized by the following equations:

$$MgO + SO_3 \rightarrow MgSO_4$$

$$MgO + SO_2 \rightarrow MgSO_3$$

$$MgSO_3 + 1/2O_2 \rightarrow MgSO_4$$

The magnesium sulfite formed in the absorber is water soluble and therefore less likely to form deposits inside the absorber and on downstream ductwork. Forced oxidation downstream from the absorber can be used to convert $MgSO_3$ and $CaSO_3$ to $MgSO_4$ and $CaSO_4 \cdot 2H_2O$ (gypsum). The gypsum is recovered by filtration and sold to either cement or wallboard manufacturers. Most of the filtrate is returned to the absorber. However, some of the filtrate is removed to purge accumulated salts, including $MgSO_4$ and $MgCl_2$, which can be used to produce magnesium hydroxide ($Mg(OH)_2$). [14, 25]

Factors Affecting Wet FGD System Operation

The principal factors affecting the operation of a wet FGD system are:

1. Flue gas flow rate;
2. Liquid-to-gas (L/G) ratio;
3. Flue gas SO_2 concentration;
4. Distribution of the flue gas and sorbent in the absorber;
5. Oxygen level in the absorber and reaction tank;
6. Slurry pH;
7. Slurry solids concentration; and
8. Solids retention time.

Flue Gas Flow Rate. The optimal waste gas flow rate for a wet FGD system is dependent on the design of the absorber unit. For some absorbers, such as those using the counterflow design, the maximum waste gas flow rate is dependent on the capacity of the mist eliminator. For absorbers that use trays, the waste gas flow rate is limited to a specific range to prevent flooding of the scrubber. [26]

Liquid-to-gas (L/G) ratio. The liquid-to-gas (L/G) ratio (in gallons of slurry per 1000 ft^3 of flue gas) in the wet FGD system is optimized to achieve the highest absorption and reaction rate, and hence the highest SO_2 removal. Systems that use higher amounts of sorbent (i.e., have higher L/G ratios) can achieve higher removal efficiencies. However, increasing the L/G ratio results in higher sorbent consumption and higher operating costs. The L/G ratio for wet FGD systems using limestone as sorbent is typically 40-100 gal to 1000 cfm. [14]

Flue Gas SO_2 Concentration. In general, the SO_2 removal efficiency decreases as the SO_2 concentration in the inlet waste gas stream increases. The high SO_2 concentrations result in rapid reductions in the alkalinity of the sorbent liquid and reduce the amount of SO_2 that will be dissolved in the liquid slurry droplets, which reduces the amount of SO_2 reacting with the sorbent. [14] SO_2 removal efficiency is also lower when the SO_2 concentration in the inlet waste gas stream is very low.

Distribution of the flue gas and sorbent in the absorber. Good distribution of the flue gas and sorbent within the absorber is necessary to ensure maximum SO_2 removal. Poor distribution of sorbent and flue gas can be compensated for by increasing the L/G ratio, however, this results in higher operating costs due to the increased sorbent usage. Wet FGDs use spray nozzles that atomize the sorbent to maximize the contact between the sorbent and the flue gas. Other techniques, such as installing guide vanes along the perimeter of the tower, may be used to improve mixing.[14]

Oxygen Level. The level of oxygen available in the absorber and reaction tank affects the chemical balance between $CaSO_3$ and $CaSO_4$. Oxidizing conditions in the absorber and reaction vessel favor the production of $CaSO_4$ since $CaSO_3$ is converted to $CaSO_4$. Since $CaSO_4$ is easier to dewater, many wet FGD systems use limestone slurry forced oxidation (LSFO) systems. In these systems, air is injected into the reaction tank, which oxidizes the calcium sulfite to calcium sulfate (gypsum). The gypsum crystals formed by the LSFO are larger and settle out from the wastewater more readily and makes the processing of the FGD system waste easier and less expensive. However, the LSFO system requires air blowers that increase the capital and operating costs for LSFO equipped wet FGD systems. [14]

Slurry pH. In the reaction tank the dissolution and crystallization reactions are controlled by the pH of the liquid, which is a function of limestone stoichiometry. The pH and limestone stoichiometry are preset parameters that are monitored and used to control the operation of the absorber. The stoichiometry for a wet FGD system using limestone as the sorbent typically varies from 1.01 to 1.1 moles of $CaCO_3$ per mole of SO_2. The pH is typically maintained between 5.0 to 6.0. The pH is used to determine the amount of limestone added to the absorber. A drop in the pH of the reaction vessel, for example, indicates increased SO_2 in the flue gas, increased consumption of the sorbent, and triggers an increase in the feed rate of the sorbent. [14]

Slurry Solids Concentration. In addition to the pH, the solids concentration in the slurry is also important for the correct operation of the wet FGD system because high solids content can cause scaling to occur in the absorber (discussed below). The solids concentration of the slurry is monitored. The typical system maintains the slurry solids concentration between 10% and 15% by weight. [14]

Solids Retention Time. The retention time for solids in the reaction tank is important for achieving high utilization of the sorbent. It also ensures the dewatering properties of solids is managed correctly. For a typical wet FGD system, the solids retention time is 12 to 14 hours. [14]

Types of Sorbent

Wet FGD systems use limestone ($CaCO_3$) or lime ($Ca(OH)_2$) as the sorbent. Limestone is the most commonly used sorbent because of its low cost. Typical wet FGD systems using lime or limestone as the sorbent can achieve SO_2 removal efficiencies of between 95% and 99%. However, the removal efficiency for limestone-based systems can be lower than those using lime. Lime is more reactive and easier to handle than limestone. Some researchers have found that aluminum and silicon impurities in limestone are retained in recovered gypsum product. Limestone with purity above 95% is recommended for wet limestone FGDs. [3, 14, 15] Typical costs for limestone and lime are reported to be \$28/ton and \$75/ton, respectively, based on 2016 data. [27]

Some manufacturers sell proprietary sorbents that contain additives (e.g., dibasic acid (DBA), formic acid) designed to promote the absorption and reaction of SO_2. These additives buffer the scrubber slurry, which controls the SO_2 vapor pressure in the scrubber and thereby maximizes the SO_2 adsorption rate. Wet FGD systems using these proprietary sorbents can achieve SO_2 removal efficiencies of 97% to 99%. [3, 18, 27]

Wet FGD systems that use seawater as the sorbent solution are also available. In these systems, the limestone dissolves completely in the seawater. The SO_2 reacts with the sorbent liquid in a counterflow absorber. The sulfate produced remains in solution in the seawater and is discharged into the ocean. Generally, wastewater treatment is not required as sulfate salts are already present in seawater and the increase in concentration is minimal. However, the high chloride concentration of seawater can result in corrosion problems. Hence, wet FGD systems using seawater must be constructed of corrosion resistant materials that increase the capital costs. [14]

Removal Efficiency

Wet FGD systems typically achieve between 90% and 99% removal efficiency. The typical LSFO system can achieve 96% SO_2 removal efficiency and SO_2 emission rates as low as 0.06 lbs/MMBtu. [3, 10] In 2018, the Centralia power plant reported 0.04 – 0.05 lb/mmBtu annual SO_2 emissions from its tangentially-fired boilers with LSFO.

Factors impacting the removal efficiency include the type and quantity of sorbent used, effectiveness of the sorbent injection system in dispersing the sorbent slurry throughout the flue gas stream and the concentration of SO_2 in the flue gas. Removal efficiencies are also impacted by the amount of sorbent used. The liquid-to-gas (L/G) ratio (in gallons of slurry per 1000 ft^3 of flue gas) is typically used to provide a guide to optimizing the design of a wet scrubber system. Systems that use higher amounts of sorbent (i.e., have higher L/G ratios) can achieve higher removal efficiencies. [14]

Wastewater Treatment

Wet FGD scrubber systems may be either recirculating or once-through wet FGD systems. In a recirculating system, most of the slurry at the bottom of the scrubber is recirculated back within the scrubber and occasionally a blowdown or bleed stream is transferred to a solids separator, where the wastewater is either recycled back to the scrubber or transferred to a wastewater treatment system. In a once-through system, all of the FGD slurry at the bottom of the scrubber is transferred to the solids separator without being recirculated. In addition to the wastewater generated by the solids separator, wastewater may also be generated from the solids drying process and from gypsum washing. [57, 58, 59]

Wastewater generated by wet FGD systems often contain metals, such as arsenic, beryllium, cadmium, chromium, mercury, lead, selenium, and copper, as well as other pollutants, including cyanide, ammonia, phosphorus, nitrogen, and total suspended solids (TSS). Pollutant concentrations vary depending on numerous factors, including the type of fuel burned, type of sorbent used, the level of recirculation of the slurry before discharge, and the materials used in the construction of the wet FGD system. The types of air pollution controls operated upstream of the wet FGD system can also affect the wastewater characteristics. For example, plants that operate a particulate collection system (*e.g.*, electrostatic precipitator) upstream from the wet FGD system will generate wastewater with lower levels of arsenic, mercury, and TSS due to the removal of fly ash. [57, 58]

The EPA published technology-based effluent limitations guidelines and standards (ELGs) for the steam electric power generating point source category in 2015, which included ELGs for wastewater generated by wet FGD systems. The rule set standards for both the direct discharge of FGD wastewater to surface water and the discharge to a publicly owned treatment works (POTW). The rule included ELGs for both new sources (greenfield) and existing sources. The EPA revised the 2015 ELGs for FGD wastewater at existing plants in 2020. As a result, wastewater generated by wet FGD systems are subject to numeric effluent limitations on mercury, arsenic, selenium, and nitrate/nitrite such as nitrogen. [57]

For existing facilities, the ELGs can be met using chemical precipitation followed by biological treatment and filtration as a final polishing step. For the chemical precipitation pretreatment step, chemicals are added to help remove suspended and dissolved solids and metals. The precipitated solids are then removed from solution by coagulation/flocculation, followed by clarification and/or filtration. The chemical precipitation uses hydroxide precipitation, sulfide precipitation (organosulfide), and iron coprecipitation to remove suspended solids and to convert soluble metal ions to insoluble metal hydroxides or sulfides. [57]

In the biological treatment step, microorganisms are used to treat the FGD wastewater. The 2015 rule focused on a high residence time biological reduction system that uses anoxic/anaerobic fixed-film bioreactors with a residence time of approximately 10 to 16 hours. For the 2020 amendments, EPA based in the revised ELGs for existing plants on low hydraulic residence time biological reduction (LRTR), which is a biological treatment system that targets removal of selenium and nitrate/nitrite using fixed-film bioreactors in smaller, more compact reaction vessels than those used in the biological treatment system evaluated in the 2015 rule (i.e., high residence time biological reduction). The LRTR system is designed to operate with a shorter residence time (approximately 1 to 4 hours), while still removing significant volumes of selenium and nitrate/nitrite. [57]

For new plants, the ELGs were based on a thermal evaporation process that uses a falling-film evaporator or brine concentrator to produce a concentrated wastewater stream and a distillate stream. The thermal evaporation process reduces the volume of wastewater by 80 to 90% thereby reducing the discharge of pollutants. The concentrated wastewater is usually further processed in a crystallizer, which produces a solid residue for landfill disposal and additional distillate that can be reused within the plant or discharged. The principal benefit of using a thermal evaporation process is that it removes a broad range of pollutants.

The ELG includes a Voluntary Incentives Program that allows existing plants until December 31, 2028 to comply if they adopt additional process changes and controls that achieve more stringent limitations on mercury, arsenic, selenium, nitrate/nitrite, bromide, and TDS. [57]

Potential Operating Issues for Wet FGD Systems

Corrosion. Both the acids in the flue gas and the sorbent slurry can cause significant corrosion in wet FGD systems. Acids are formed when the hot, dry flue gases initially enter the moist environment of the absorber vessel. The sorbent slurry is abrasive and can cause pitting of metal surfaces that increases the risk of corrosion. Flue gas exiting the absorber is saturated with water that contains some acidic gases (e.g., SO_3) that are corrosive to downstream equipment such as fans, ducts, and stacks. The high temperatures and presence of halides in the flue gas (primarily chlorides and fluorides from impurities in limestone and coal) also contribute to the corrosive conditions.[14, 28]

For constructing the absorbers and ancillary equipment (e.g., slurry tanks, pumps, downstream control system components and stacks), manufacturers select materials that can withstand the corrosive conditions that prevail in each component of the FGD system. The level of acidity within the various components can vary significantly. For example, the pH of gas entering the absorber can be as low as 1, while the acidity in the absorber typically ranges from to 4.0 to 5.5. The flue gas leaving the absorber contains some acid, which will condense if the gas contacts cold duct walls as it exits through the outlet ducting and chimney stack. The condensate is highly acidic and can cause significant corrosion. The flue gas leaving the stack can mix with moisture in the atmosphere, forming acidic droplets that fall into the stack and cause corrosion to occur on the stack walls. [28] The cost for retrofitting an existing stack with corrosion-resistant liners can be sufficiently high that, for some retrofits, the cost of retrofitting an existing stack can be higher than the cost of replacing the existing stack with a new stack. [52]

Nickel alloys and nickel-alloy clad steel can be used to construct components that experience the most corrosive conditions. Nickel alloys are often used to fabricate the absorber outlet duct, any downstream auxiliary equipment (e.g., wet electrostatic precipitator, heat exchanger) and chimney stack. Nickel-chromium and iron-nickel chromium alloys that contain molybdenum have been found to be effective in FGD systems, with alloys containing more molybdenum found to be the most resistant to corrosion. Super-austenitic stainless steels containing 6% to 7% molybdenum can be used for fabricating the absorber vessel, while austenitic stainless steels and duplex alloys can be used to fabricate components of the FGD system that experience less corrosive environments (e.g., the inlet and bypass ducts). [28] Using high alloy containing steels or rubber-lined absorber vessels or pumps increases the reliability of the wet FGD system but can increase the capital cost by 10-20%. [52]

Fiberglass reinforced plastics (FRP) have also been used in FGD systems as a reliable and less expensive alternative to nickel alloys. Chemical resistance of FRP made from epoxy vinyl ester resin is said to be equivalent to nickel alloys. FRP are not chemically affected by the majority of chloride and sulfite trace elements found in FGD systems that increase the susceptibility of stainless steel to pitting and stress-corrosion cracking. Although some FRP are reported to withstand temperatures of up to 600 ^{o}F, FRP are generally used in applications of between 300 to 400^{o}F. However, FRP may not be suitable for systems where temperatures exceed 600°F for more than a few minutes, where wide swings in temperature occur, or where extreme abrasive conditions exist. [54, 55]

To reduce corrosion inside the absorber, some absorbers use spray systems that continuously wash the inside walls with slurry. The absorber vessel floor and ducts are designed so that condensate is drained from the system. Chimney stacks are typically constructed of nickel alloy clad steel and have condensate collection systems made of alloy plate to collect condensate that forms on the interior walls of the stack. [28]

To minimize corrosion problems in equipment downstream from the absorber, many manufacturers also use mist eliminators followed by heat exchangers. The mist eliminator removes any remaining droplets of slurry from the flue gas stream. The heat exchangers keep the temperature of the exhaust gases above their dew point and reduce condensation inside the ducts, fans and stack. This approach also reduces plume opacity. [14, 28] Electrochemical protection has been demonstrated to reduce corrosion in reaction tanks. [29]

Scale Formation. Calcium sulfate (gypsum) formed from the oxidation of calcium sulfite can form scale in the absorber, mist collector and downstream ducts. Scale formation generally occurs when the fraction of calcium sulfate in the slurry exceeds 15 percent. Scale formation can be reduced by increasing the L/G ratio, using LSFO technology described above, or by using the limestone inhibition oxidation process (LSIO). In the LSIO process, sulfur or sodium thiosulfate ($Na_2S_2O_3$) is added to the limestone slurry to prevent oxidation of the sulfite to sulfate. Where sulfur is used in place of $Na_2S_2O_3$, between 50% and 75% of the sulfur reacts with sulfite to produce thiosulfate. The amount of $Na_2S_2O_3$ added depends on the SO_2 emissions and the amount of oxygen in the waste gas stream. LSIO is useful for applications where coal with high sulfur content is burned. [14]

High levels of chlorine in the flue gas can also contribute to scale formation as the chlorine reacts with the sorbent to produce calcium chloride. [14, 26]

1.2.4.2 Wet FGD System Design Procedures

Methods for estimating the limestone feed rate, make-up water consumption, waste generation rate and auxiliary power consumption for a wet FGD system are presented in this section. These parameters are used in the cost methodology presented in Section 1.2.3.4 for estimating the annual costs for operating a wet FGD system. The methods presented in this section are those developed by the EPA's Clean Air Markets Division (CAMD) for the Integrated Planning Model (IPM) and its cost methodology. [21]

Limestone Feed Rate

Reagent usage is a function of the combustion unit size, SO_2 emission rate and the removal efficiency of the wet FGD system. In the IPM, the reagent feed rate equation is based on a 98% SO_2 removal. The equation below includes an adjustment factor that allows the reagent usage to be adjusted to reflect the actual or target removal efficiency. The IPM cost methodology also assumes a calcium-to-sulfur stoichiometric ratio of 1.03 and a limestone purity of 90% $CaCO_3$, with the balance being inert materials. The reagent usage rate is calculated as follows:

$$Q_{Limestone} = \frac{17.52 \times A \times S \times HRF}{2{,}000} \times \frac{EF}{0.98} \qquad (1.9)$$

Where

$Q_{Limestone}$ = lime consumption rate, tons/hour
S = uncontrolled SO_2 emission rate, lb/MMBtu
A = Unit size (Gross), MW
HRF = heat rate factor (calculated using Equation 1.6)

EF = removal efficiency, fraction
0.98 = SO_2 removal efficiency basis (listed as a fraction) for the IPM cost model.

Make-up Water

The makeup water rate is a function of combustion unit size and SO_2 emissions. The make-up water consumption rate for a wet FGD system is calculated using the following equation:

$$q_{water} = \frac{[(1.674 \times S) + 74.68] \times A \times CoalF \times HRF}{1{,}000} \tag{1.10}$$

Where

q_{water} = make-up water consumption rate, 1,000 gallons (kgallons)/hour
S = uncontrolled SO_2 emission rate, lb/MMBtu
A = Unit size (Gross), MW
$CoalF$ = coal factor (CoalF=1 if bituminous; CoalF=1.05 if Powder River Basin (PRB)[2]; CoalF=1.07 if Lignite)
HRF = heat rate factor (calculated using Equation 1.6).

Waste Generation Rate

The waste generation rate is directly proportional to the reagent usage. The wastewater generation rate estimation method in the IPM assumes a 10% moisture content in the by-product and is based on an SO_2 removal of 98% of the inlet SO_2. The ratio of the actual removal efficiency to the base model removal efficiency of 98% adjusts the waste generation rate for actual or target removal efficiency of the wet FGD system.

$$q_{waste} = 1.811 \times Q_{Limestone} \times \frac{EF}{0.98} \tag{1.11}$$

Where:

q_{waste} = waste generation rate, tons/hour
$Q_{Limestone}$ = limestone input rate, tons/hour
EF = SO_2 removal efficiency, fraction
0.98 = SO_2 removal efficiency basis for the IPM cost methodology.

Auxiliary Power Costs

The auxiliary power (also commonly called parasitic energy loss) is the additional power required to operate the wet FGD system. The additional auxiliary power required for the wet FGD system in kilowatts (kW) is estimated in the IPM as a percentage of the total unit gross production and is dependent on the sulfur emissions and size of the combustion unit. The

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[2] PRB was the most common type of subbituminous coal in use in the U.S. as of 2016. Also, PRB was approximately 95% of all subbituminous coal produced in the U.S., based on coal production data from the Energy Information Administration's (EIA) December 2019 Quarterly Coal Report (Table 2) and the 2018 Annual Coal Report (Table ES1).

additional power calculated using the IPM model includes the increased fan power required to account for the additional pressure drop caused by the installation of the wet FGD system.

$$P = 0.0112e^{0.155 \times S} \times CoalF \times HRF \times A \times 1{,}000 \tag{1.12}$$

Where:

P = electrical power consumption of the FGD system, kW
S = SO_2 emission rate, lb/MMBtu
0.0112 = constant in the equation
0.155 = constant in the equation
A = Unit size (Gross*)*, MW
$CoalF$ = coal factor (CoalF = 1 if bituminous; CoalF = 1.05 if PRB; CoalF = 1.07 if Lignite)
HRF = heat rate factor (calculated using Equation 1.6)
1,000 = conversion factor, kW/MW.

1.2.4.3 Estimating Total Capital Investment

The methodology presented here provides an approach to estimate study-level capital and annual costs for wet FGD systems. The cost equations are based on the methodology developed by Sargent & Lundy, LLC (S&L) for EPA's CAMD and used with the IPM.[3] [21] The IPM version 6 includes algorithms for estimating the capital and operating costs. The IPM algorithms were compared with recent wet FGD system installation projects and found to be consistent with current costs.[21] Also presented in our methodology here, as found in the ELG technical documentation, are capital and operating cost algorithms for a wastewater treatment system consisting of chemical pretreatment, low hydraulic residence time biological reduction and ultrafiltration process to treat wastewater generated by the wet FGD system. The cost equations are based on data collected by the EPA during the development of the ELG (see discussion of wastewater treatment in section 1.2.4.1) and can be used to estimate costs of compliance with the ELG standards for existing sources. The equations are based on the FGD wastewater flow rate. If the actual FGD wastewater flow rate is unknown, the rate can be estimated using an average FGD wastewater flow rate of 0.4 pgm/MW. [57]

The IPM equations estimate the purchased equipment cost and the direct and indirect installation costs together based on cost data for multiple lump-sum contracts. Turnkey contracts, where the price is fixed at the time the contract is signed and the contractor undertakes

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[3] The IPM is based on a statistical evaluation of cost data from various sources, including estimates for wet FGD systems from the "Analysis of MOG and LADCO's FGD and SCR Capacity and Cost Assumptions in the Evaluation of Proposed EGU 1 and EGU 2 Emission Controls" prepared for Midwest Ozone Group (MOG), data from the "Current Capital Cost and Cost-effectiveness of Power Plant Emissions Control Technologies" prepared by J. E. Cichanowicz for the Utility Air Regulatory Group (UARG) in 2010, 2007 to 2008 data published by G. W. Sharp in "Update: What's That Scrubber Going to Cost?" published in Power Magazine, March 2009, and unpublished from the S&L in-house database of wet FGD projects. Industry data from "Current Capital Cost and Cost-effectiveness of Power Plant Emissions Control Technologies" prepared by J.E. Cichanowicz for the Utility Air Regulatory Group (UARG) in 2012 to 2014 were used to update the algorithms to 2016.

responsibility for the completion of the project, are generally 10 to 15% higher than the multiple lump-sum contracts. The capital costs equations estimate costs in 2016 dollars.

The capital costs estimated using the IPM methodology are consistent with the installed costs reported to the U.S. Department of Energy, Energy Information Administration (EIA). The EIA data includes various designs of wet FGD systems. For those systems with removal efficiency of 90% or greater, the installed costs reported to EIA range between $250,000 to $709 million, with an average cost of $114 million. For a 500 MW unit calculated using the IPM methodology, the TCI is estimated at $244 million (see the example in Section 1.2.3.6).

The capital cost equations developed by S&L are for typical wet FGD system installations retrofitted on existing coal-fired utility units of capacity of over 100 MW (or 1,000 MMBtu/hour), achieving a removal efficiency of 98% with SO_2 emission level of 0.06lb/MMBtu Over 100 MW (or 1,000 MMBtu/hour). For units under 100 MW, capital costs for a wet FGD system are approximately $900/kW. [21]

The capital costs for units over 100 MW are impacted primarily by the unit size and the type and sulfur content of the coal burned. The larger the size of the combustion unit, the larger the amount of flue gas generated. The size of the absorber, reagent preparation equipment, waste handling, and the balance of plant costs are all impacted by the quantity of flue gas generated. The size and cost of the reagent and waste handling equipment are also dependent on the type of coal burned. The capital cost estimated using this methodology includes the capital costs for a wastewater treatment facility consisting of only minor physical and chemical wastewater treatment. The capital costs should be adjusted if a more extensive wastewater treatment system is required at a specific plant site.

Total capital investment (TCI) includes direct and indirect costs associated with purchasing and installing the wet FGD system equipment. TCI includes the equipment cost for the wet FGD system, the cost of auxiliary equipment, direct and indirect installation costs, additional costs due to installation, costs for buildings and site preparation, cost of land and working capital. A more detailed discussion of what is included in capital costs can be found in Section 1, Chapter 2 of this Manual.

Cost calculations in the IPM are based primarily on data for retrofits. However, retrofits are typically 30% more than for new units of the same size and design. [3] To adjust for the additional costs associated with retrofits, we have included a retrofit factor in the TCI equations. A retrofit factor of 0.77 should be used to estimate capital costs for new construction using these equations and a retrofit factor of 1 should be used for wet FGD system retrofits to existing units, where the retrofit is of an average level of difficulty. Retrofit costs vary significantly from site to site and depend on the amount of space available and whether significant modifications to existing equipment (e.g., ductwork, stack) are needed. A higher retrofit factor should be used for congested sites.

The IPM capital cost equation for a wet FGD system is presented in Equation 1.13:

$$TCI = 1.3 \times \ (ABS_{Cost} + RPE_{Cost} + WHE_{cos\,t} + BOP_{Cost}) + WWT_{Cost} \tag{1.13}$$

Where:

TCI = total capital investment for a wet FGD system, $

ABS_{Cost} = cost of the absorber island, $

RPE_{Cost} = reagent preparation equipment cost, $

WHE_{Cost} = waste handling equipment cost, $

BOP_{Cost} = balance of plant costs, $

WWT_{Cost} = cost of wastewater treatment facilities, $.

The TCI calculation shown in Equation 1.13 includes a factor of 1.3 to estimate engineering and construction management costs, installation, labor adjustment for the wet FGD system, and contractor profit and fees. The owner's costs (for owner activities related to engineering, management, and procurement) and costs such as allowance for funds used during construction (AFUDC) are capital cost items that are not included in the EPA Control Cost Manual methodology and are inconsistent with the overnight cost method[4] that is a key basis for the Control Cost Manual methodology, and thus are not included in the TCI estimates in this section or in other Control Cost Manual chapters. The 1.3 factor does not apply to the wastewater treatment system as the direct and indirect costs for the wastewater system are included in the cost equations.

The costs for the absorber island (ABS_{Cost}), reagent preparation (RPE_{Cost}), waste handling equipment (WHE_{cost}), balance of plant costs (BOP_{Cost}) and wastewater treatment facility ($\mathrm{WWT}_{\mathrm{Cost}}$) are discussed in the following sections. The estimated cost varies depending on the size (or capacity) of the unit. The gross heat rate impacts the amount of flue gas generated and ultimately impacts the size of the absorber, reagent preparation, waste handling, and the balance of plant costs. The SO_2 emissions rate primarily impacts the capital costs of the reagent and waste handling facilities, while the type of fuel burned impacts the quantity of flue gas and thereby impacts the BOP_{Cost}.

The ABS_{Cost} and BOP_{Cost} are also impacted by the unit's elevation with respect to sea level. The original cost calculations were developed for wet FGD systems located within 500 feet of sea level. Therefore, for plants located at higher elevations, the base costs should be adjusted for the effects of elevation on the flue gas volume using an elevation factor, *ELEVF*. The elevation factor, *ELEVF*, is 1 for plants located below 500 feet above sea level. If the plant is located more than 500 feet above sea level, *ELEVF* is calculated using Equations 1.7 and 1.8 in Section 1.2.2.4.

The capital costs for the absorber island are estimated as follows:

$$ABS_{Cost} = 584{,}000 \times RF \times (CoalF \times HRF)^{0.6} \times (S/2)^{0.02} \times A^{0.716} \times ELEVF \qquad (1.14)$$

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[4] The overnight cost estimation method presumes costs are incurred as if the project in question incurred no interest during construction, or was built "overnight." Another description of this method is the present value cost that would have to be paid as a lump sum up front to completely pay for a construction project. For more information, see "Conducting Technical and Economic Evaluations – As Applied for the Process and Utility Industries," Recommended Practice 16R-90, American Association of Cost Engineering International. April 1991, and Section 1, Chapter 2 of this Control Cost Manual.

Where

ABS_{Cost} = Wet FGD system absorber island costs, \$
584,000 = constant
HRF = heat rate factor
$CoalF$ = coal factor ($CoalF$ = 1 if bituminous; $CoalF$ = 1.05 if PRB; $CoalF$ = 1.07 if Lignite)
S = SO_2 emission rate, lb/MMBtu
RF = retrofit factor (RF = 0.77 for new construction; RF = 1 for retrofits with average level of difficulty)
A = Unit size (Gross), MW
$ELEVF$ = elevation factor (calculated using Equations 1.7 and 1.8, if plant is located above 500 feet above sea level; ELEVF = 1 for plants located at 500 ft above sea level).

The reagent preparation and handling equipment costs (RPE_{Cost}) for wet FGD systems are calculated as follows:

$$RPE_{Cost} = 202{,}000 \times RF \times (S \times HRF)^{0.3} \times A^{0.716} \tag{1.15}$$

Where:

RPE_{Cost} = reagent preparation and handling equipment cost, \$
202,000 = constant in the equation
S = SO_2 emission rate, lb/MMBtu
RF = retrofit factor (RF = 0.77 for new construction; RF = 1 for retrofits with average level of difficulty)
HRF = heat rate factor (calculated using Equation 1.6)
A = Unit size (Gross), MW.

The waste handling equipment costs (WHE_{cost}) are calculated as follows:

$$WHE_{Cost} = 106{,}000 \times RF \times (S \times HRF)^{0.45} \times A^{0.716} \tag{1.16}$$

Where:

WHE_{Cost} = Waste handling equipment costs, \$
106,000 = constant in the equation
A = Unit size (Gross), MW
HRF = heat rate factor
S = SO_2 emission rate, lb/MMBtu
RF = retrofit factor (RF = 0.77 for new construction; RF = 1 for retrofits with average level of difficulty).

The BOP_{Cost} are calculated as follows:

$$BOP_{Cost} = 1{,}070{,}000 \times RF \times (CoalF \times HRF)^{0.4} \times A^{0.716} \times ELEVF \tag{1.17}$$

Where:

BOP_{Cost} = Balance of plant costs, $

1,070,000 = constant in the equation

A = Unit size (Gross), MW

HRF = heat rate factor

$CoalF$ = coal factor ($CoalF$ = 1 if bituminous; $CoalF$ = 1.05 if PRB; $CoalF$ = 1.07 if Lignite)

RF = retrofit factor (RF = 0.77 for new construction; RF = 1 for retrofits with average level of difficulty)

$ELEVF$ = elevation factor (calculated using Equations 1.7 and 1.8, if plant is located above 500 feet above sea level; $ELEVF$ = 1 for plants located at 500 ft above sea level).

The WWT_{Cost} are capital costs for chemical and biological wastewater treatment facilities. The chemical pre-treatment system includes pumps, tanks and mixers, reactors, chemical feed systems, clarifiers, filter presses, and sand filters. The biological treatment system includes anoxic/anaerobic bioreactors, control skids, backwash skids, tanks, pumps, a heat exchanger, an ultrafilter, and chemical feed skids. Equations 1.18a and 1.18b calculate the total capital costs for the wastewater treatment system, including the purchased equipment costs for the system components, capital costs for pollutant monitoring and analysis equipment (including a mercury analyzer), and direct and indirect costs for the installing the wastewater treatment system. Direct costs include costs for freight, installing piping and electric supply, site preparation, constructing buildings, and installing instrumentation/controls. Indirect costs include engineering and other costs associated with the wastewater treatment system. The capital cost equations assume the nitrate/nitrite concentrations of the FGD wastewater are less 50 mg/L. An additional pretreatment system is required for FGD wastewater with higher concentrations of nitrate/nitrite.

For plants with onsite landfills, the capital costs for chemical pretreatment, LRTR biological treatment and ultrafiltration can be estimated using the following equation:

$$WWT_{cost} = (41.36\ F + 11{,}157{,}588) \times RF \times 0.898 \tag{1.18a}$$

For plants without onsite landfills, the capital costs for chemical pretreatment, LRTR biological treatment and ultrafiltration can be estimated using the following equation:

$$WWT_{cost} = (41.16\ F + 11{,}557{,}843) \times RF \times 0.898$$

(1.18b)

The FGD wastewater flow rate, F, is estimated as:

$$F = A\ \times 0.4 gpm/MW \tag{1.18c}$$

Where:

WWT_{Cost} = capital costs for wastewater treatment facility, \$
A = Unit size (Gross), MW
F = FGD wastewater flow rate, gallons per minute
RF = retrofit factor (RF = 0.77 for new construction; RF = 1 for retrofits with average level of difficulty).

1.2.4.4 Estimating Total Annual Cost for a Wet FGD System

In general, the total annual costs (TAC) consist of direct costs, indirect costs, and recovery credits. Direct annual costs are associated with and proportional to the quantity of waste gas processed by the control system. Indirect (fixed) annual costs are independent of the operation of the control system and would be incurred even if it were shut down. Recovery credits are the credits from the sale of any salvageable byproducts. Each of these costs is discussed in the sections below. A more detailed discussion of annual costs can be found in Section 1, Chapter 2 of this Cost Manual.

Direct Annual Costs

Direct annual costs (*DAC*) include variable and semi-variable costs. Variable direct annual costs account for purchase of reagent, utilities (electrical power and water), and any additional waste disposal costs resulting from the operation of the wet FGD system. Semi-variable direct annual costs include operating and supervisory labor and maintenance (labor and materials). These costs are discussed individually below. The direct annual costs for a wet FGD system are calculated as shown in Equation 1.19.

$$DAC = \begin{pmatrix} Annual \\ Maintenance \\ Cost \end{pmatrix} + \begin{pmatrix} Annual \\ Operator \\ Cost \end{pmatrix} + \begin{pmatrix} Annual \\ Reagent \\ Cost \end{pmatrix} + \begin{pmatrix} Annual \\ Make-up \\ Water \\ Cost \end{pmatrix} + \begin{pmatrix} Annual \\ Waste \\ Disposal \\ Cost \end{pmatrix} + \begin{pmatrix} Annual \\ Auxiliary \\ Power \\ Cost \end{pmatrix} + \begin{pmatrix} Annual \\ Wastewater \\ Treatment \\ Cost \end{pmatrix} \tag{1.19}$$

Annual Maintenance Cost. The annual maintenance cost includes the costs for labor and material and is estimated to be 1.5% of the *TCI*. This is a fairly standard percentage used to estimate the maintenance costs for many types of emissions control devices. The equation for annual maintenance cost, *AMC,* is given by:

$$AMC = 0.015 \times TCI \tag{1.20}$$

Annual Operator Labor Cost. The cost for operator labor is calculated using Equation 1.21, where *FT* is the number of additional full-time personnel required to operate the wet FGD system:

$$Annual\ Operator\ Labor\ Cost = FT \times 2080 \times Hourly\ Labor\ Rate \qquad (1.21)$$

In general, 16 additional full-time personnel are required to operate a wet FGD system for combustion unit greater than 500 MW (5,000 MMBtu/hour) and 12 additional full-time personnel for units equal to or less than 500 MW.

Annual Reagent Cost. The annual cost of reagent purchased, AR_{cost}, in \$/yr is estimated using the limestone input rate, $Q_{Limestone}$, (calculated using Equation 1.9, the effective operating time, t_{op} (calculated using Equation 1.5 in Section 1.2.2.2)) and the cost of limestone, $Cost_{limestone}$, in dollars per ton (\$/ton).

$$AR_{Cost} = Q_{Limestone} \times Cost_{Limestone} \times t_{op} \qquad (1.22)$$

Where

AR_{cost} = annual lime cost, \$/year
$Q_{Limestone}$ = limestone input rate, tons/hour
$Cost_{limestone}$ = purchase price of lime, \$/ton
t_{op} = effective number of operating hours, hour/year.

Annual Make-up Water Cost. Using the estimated make-up water consumption rate calculated in Equation 1.10 and the effective number of operating hours (t_{op}) calculated using Equation 1.5 in Section 1.2.2.2, the annual cost of make-up water in \$/yr is given by:

$$Annual\ water\ cost = q_{water} \times Cost_{water} \times t_{op} \qquad (1.23)$$

Where:

q_{water} = make-up water consumption rate, 1,000 gallons/hour
$Cost_{water}$ = unit cost of water, \$/gallon
t_{op} = effective number of operating hours, hours/year.

Annual Waste Disposal Cost. Using the estimated waste generation rate calculated in Equation 1.11, and the effective number of operating hours (t_{op}) calculated using Equation 1.5 in Section 1.2.2.2, the annual cost of waste disposal is given by:

$$Annualwastedisposal\ cos\,t = q_{waste} \times Cost_{disposal} \times t_{op} \qquad (1.24)$$

Where

q_{waste} = waste generation rate, tons/hour
$Cost_{disposal}$ = unit cost of waste disposal, \$/ton
t_{op} = effective number of operating hours, hours/year.

Annual Auxiliary Power Cost. Using the estimated auxiliary power consumption from Equation 1.12 and the effective number of operating hours (t_{op}) calculated using Equation 1.5 in Section 1.2.2.2, the annual cost of electricity is calculated as follows:

$$Annual\ electricity\ cost = P \times Cost_{elec} \times t_{op} \tag{1.25}$$

Where:

P = electrical power consumption of the control device, kW
$Cost_{elec}$ = unit price of electricity, \$/kWh
t_{op} = effective number of operating hours, hours/year.

Annual Wastewater Treatment Costs. The annual operating costs for the wastewater treatment system include operating labor, maintenance materials and labor, chemicals, energy, sludge transportation and disposal. Operating costs vary depending on whether the solids are disposed onsite or offsite. For plants with onsite landfills, the operating costs for chemical pretreatment, biological treatment, and ultrafiltration can be estimated using the following equation:

$$Annual\ Wastewater\ Treatment\ Cost\ = (4.847\mathrm{F} + 479{,}023) \times 0.958 \times CF \tag{1.26a}$$

For plants without onsite landfills, the operating costs for chemical pretreatment, biological treatment and ultrafiltration can be estimated using the following equation:

$$Annual\ Wastewater\ Treatment\ Cost\ = (6.3225\mathrm{F} + 472{,}080) \times 0.958 \times CF \tag{1.26b}$$

Where:

F = FGD wastewater flow rate, gallons per minute (calculated using Equation 1.18(c))
CF = System capacity factor.

As stated in the ELG, EPA recommends FGD wastewater treatment systems be equipped with a continuous water quality monitor for measuring mercury concentrations in the treatment system effluent. These monitors enable plant operators to adjust the chemical precipitation process as needed to optimize pollutant removal. The expected life of a mercury analyzer is 6 years and costs approximately \$100,000 to replace.

Indirect Annual Costs

In general, indirect annual costs (fixed costs) include the capital recovery cost, property taxes, insurance, administrative charges, and overhead. Capital recovery cost is based on the anticipated equipment life and the annual interest rate employed. The equipment life is the expected service life of the control device. As noted in Section 1.1.2, we expect an equipment life of 20 to 30 years for wet FGD systems. One study of coal-fired U.S. power plants found that 50% of the scrubbers at power plants were over 20 years old, with the oldest still operating after 34 years.[27]. The wastewater treatment system can reasonably be expected to operate for over 20 years based on the reported performance characteristics of the wastewater system components. However, the remaining life of the controlled combustion unit may also be a determining factor when deciding on the correct equipment life for calculating the total annual costs. Given these considerations, we estimate an equipment life of 30 years as appropriate for wet FGD systems. An equipment life of 30 years is assumed for the example problem presented in Section 1.2.3.6.

For this analysis, taxes are assumed to be zero since property taxes generally do not apply to air pollution control equipment. The cost of overhead for a wet FGD system is also considered to be zero as costs for insurance on an absorber system and additional payroll overhead (i.e., expenses related to labor employed in operation and maintenance of hardware) and plant overhead (i.e., costs for plant protection, control laboratories, and parking areas) are expected to be minimal.

Using these assumptions, indirect annual costs in \$/yr, *IDAC*, consist of both administrative charges and capital recovery, which can be expressed as:

$$IDAC = AC + CR \tag{1.27}$$

where *AC* represents the administrative charges and *CR* represents the capital recovery cost. Administrative charges may be calculated as:

$$AC = 0.03 \times (\text{Annual Operator Cost} + 0.4 \text{ x Annual maintenance cost}) \tag{1.28}$$

The annual maintenance and operator costs are calculated using Equations 1.19 and 1.20, respectively.

Capital recovery is estimated as:

$$CR = CRF \times TCI \tag{1.29}$$

Where TCI is the total capital investment in dollars and CRF is the capital recovery factor. Capital recovery factor was defined in Section 1, Chapter 2 of the Manual as:

$$CRF = \frac{i(1+i)^n}{(1+i)^n - 1} \tag{1.30}$$

Where *i* is the assumed interest rate and *n* is the equipment life of the wet FGD system.

Total Annual Cost

The total annual cost, *TAC,* for owning and operating the SDA FGD system is the sum of direct and indirect annual costs as given in the following equation:

$$TAC = DAC + IDAC \tag{1.31}$$

1.2.4.5 Cost Effectiveness

The cost in dollars per ton of SO_2 removed per year, is calculated as follows:

$$Cost\ Effectiveness\ = \frac{TAC}{SO_2\ Removed/yr} \tag{1.32}$$

Where:

Cost Effectiveness = the cost effectiveness, \$/ton
SO_2 *Removed/yr* = annual mass of SO_2 removed by the absorber, ton/yr
TAC = total annual cost, \$/year.

1.2.4.6 Example Problem for a Wet FGD System

This section provides an example wet FGD system illustrating how the design parameters and capital and annual costs are calculated using the methodology outlined in Sections 1.2.3.3 through 1.2.3.5. This example is purely illustrative, and is not meant to serve in place of the actual data on a wet FGD system that is included in an actual cost analysis. The design basis is a wet FGD system installed on an existing 500 MW utility boiler firing bituminous coal with onsite landfill for disposal of residue from the wastewater treatment system. For the purposes of this example, the boiler is located at an altitude between sea level and 500 feet. The retrofit is assumed to be of typical (average) complexity with the site having sufficient space available to install the wet FGD system and requiring minimal changes to the existing plant. The following assumptions are made to perform the calculations:

Boiler MW rating full load capacity, *A*	500 MW
Gross heat input rate, *GHR*	9,500 Btu/kWh
Uncontrolled SO_2 emission rate, S_{in}	3.0 lb/MMBtu
Required controlled SO_2 emission rate, S_{out}	0.15 lb/MMBtu
Effective annual operating hours, t_{op}	8,000 hours/year
Retrofit factor, *RF*	1
Elevation factor, *ELEVF*	1

In addition to these assumptions regarding the physical parameters of the boiler, the economic factors for the example cost calculations are as follows:

Cost Year	2016
Equipment Life	30 years
Annual Interest Rate[5]	3.25%
Limestone cost[6]	30 \$/ton

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[5] As stated in Section 1, Chapter 2 of the Cost Manual, the interest rate that is appropriate for annualizing capital cost is either the bank prime rate (currently 3.25%), which is an interest rate set by the Federal Reserve Board that fluctuates with the market for financial credit, or a firm-specific rate that reflects the rates of debt and equity for the firm owning the unit at issue if that firm can justify this alternative rate. The bank prime rate serves as a default if the firm can, or decides not to, provide an interest rate reflecting its rates for debt borrowing and/or equity..

[6] U.S. Environmental Protection Agency (EPA). Documentation for EPA's Power Sector Modeling Platform v6 Using the Integrated Planning Model. Office of Air and Radiation. January 2017. Available at: https://www.epa.gov/airmarkets/documentation-epas-power-sector-modeling-platform-v6.

Make-up water cost[7]	0.0042 $/gal
Waste disposal cost[11]	30 $/ton
Electricity Cost[8]	0.0361 $/kWh
Labor rate[11]	60 $/hour

<u>Design Parameters</u>

The SO_2 removal efficiency, *EF*, is calculated from the inlet SO_2 emission level and the required stack SO_2 emission level:

$$EF = \left[\frac{SO_{2In} - SO_{2Out}}{SO_{2In}}\right] \times 100$$

$$EF = \left[\frac{3.0\frac{lb}{MMBtu} - 0.15\frac{lb}{MMBtu}}{3.0\frac{lb}{MMBtu}}\right] \times 100 = 95\%$$

The SO_2 removed per hour (lb/hr) is calculated from the inlet SO_2 emission level, the SO_2 removal efficiency, and the maximum heat rate to the boiler:

$$SO_2\frac{Removed}{hr} = 3.0\frac{lb}{MMBtu} \times 0.95 \times 500MW \times \frac{9{,}500Btu}{kWh} \times \frac{MMBtu}{10^6Btu} \times \frac{1{,}000kW}{MW} \quad = 13{,}538\frac{lb}{hr}$$

The heat rate factor (*HRF*) is the ratio of actual heat rate of the boiler, in terms of the boiler MMBtu/MWh, compared to a typical heat rate of 10 MMBtu/MWh. In this example, the gross plant heat input was given as 9,500 Btu/kWh. By substituting this value in Equation 1.3, we can calculate the heat rate factor for the 500 MW utility unit as follows:

$$HRF = \frac{9{,}500\ Btu/kWh}{10{,}000\ Btu/kWh} = 0.95$$

<u>Reagent Feed Rate</u>. The limestone consumption rate is calculated using Equation 1.9:

$$Q_{Limestone} = \frac{17.52 \times 500 \times 3.0 \times 0.95}{2{,}000} \times \frac{0.95}{0.98} = 12.1\ tons/hour$$

<u>Make-up Water Consumption</u>. The make-up water consumption rate is calculated using Equation 1.10:

$$q_{water} = \frac{[(1.674 \times 3.0) + 74.68] \times 500 \times 1 \times 0.95}{1{,}000} = 37.9\ kgallons/hour$$

<u>Waste Generation</u>. Using Equation 1.11, the waste generation rate for the wet FGD system is calculated as follows:

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[7] Average water rates for industrial facilities in 2016 compiled by Black & Veatch. See "50 Largest Cities Water/Wastewater Rate Survey – 2018-2019." Available at: https://www.bv.com/sites/default/files/2019-10/50_Largest_Cities_Rate_Survey_2018_2019_Report.pdf.

[8] U.S. Energy Information Administration. Electric Power Annual 2016. Table 8.4. Published December 2017. Available at: https://www.eia.gov/electricity/annual/pdf/epa.pdf.

$$q_{waste} \;=\; 1.811 \times Q_{Limestone} \times \frac{EF}{0.98} = 1.811 \times 12.1\frac{tons}{hour} \times \frac{0.95}{0.98} = 21.2\; tons/hour$$

Auxiliary Power. The auxiliary power required for the wet FGD system is estimated using Equation 1.12:

$$P \;=\; 0.0112e^{0.155 \times S} \times CoalF \times HRF \times A \times 1{,}000$$

$$P = 0.0112e^{(0.155 \times 3.0)} \times 1 \times 0.95 \times 500 \times 1{,}000$$

$$P \;=\; 8{,}470\; kW$$

FGD Wastewater Flow Rate. The FGD wastewater flow rate, F, is estimated as:

$$F = 500\; \times \left(\frac{0.4\; gpm}{MW}\right) = \; 200\; gpm$$

Total Capital Investment

Once the wet FGD system is sized, the capital and annual costs can be estimated. The total capital investment costs (TCI) are estimated using Equation 1.13:

$$TCI = 1.3 \times \; (ABS_{Cost} + RPE_{Cost} + WHE_{cost} + BOP_{Cost}) + WWT_{Cost}$$

Where:

TCI = total capital investment for a wet FGD system, \$
ABS_{Cost} = cost of the absorber island, \$
RPE_{Cost} = reagent preparation equipment cost, \$
WHE_{Cost} = waste handling equipment cost, \$
BOP_{Cost} = balance of plant costs, \$
WWT_{Cost} = cost of wastewater treatment facility, \$.

The capital costs for the absorber island are estimated using Equation 1.14:

$$ABS_{Cost} = \; 584{,}000 \times RF \times (CoalF \times HRF)^{0.6} \times (S/2)^{0.02} \times A^{0.716} \times ELEVF$$

$$ABS_{Cost} = \; 584{,}000 \times 1 \times (1.0 \times 0.95)^{0.6} \times (3.0/2)^{0.02} \times (500)^{0.716} \times 1 = \$48{,}868{,}764$$

The reagent preparation and handling equipment costs (RPE_{Cost}) for wet FGD system are calculated using Equation 1.15 as follows:

$$RPE_{Cost} = \; 202{,}000 \times RF \times (S \times HRF)^{0.3} \times A^{0.716}$$

$$RPE_{Cost} = \; 202{,}000 \times 1 \times (3 \times 0.95)^{0.3} \times 500^{0.716} = \$23{,}673{,}766$$

The waste handling equipment costs (WHE_{cost}) include costs are calculated using Equation 1.16:

$$WHE_{Cost} = \ 106{,}000 \times RF \times (S \times HRF)^{0.45} \times A^{0.716}$$

$$WHE_{Cost} = \ 106{,}000 \times 1 \times (3 \times 0.95)^{0.45} \times (500)^{0.716} = \$14{,}536{,}123$$

The BOP_{Cost} can be calculated using Equation 1.17:

$$BOP_{Cost} = \ 1{,}070{,}000 \times RF \times (CoalF \times HRF)^{0.4} \times A^{0.716} \times ELEVF$$

$$BOP_{Cost} = \ 1{,}070{,}000 \times 1 \times (1 \times .95)^{0.4} \times 500^{0.716} \times 1 = \$89{,}729{,}602$$

For plants with onsite landfills, the WWT_{Cost} can be calculated using Equation 1.18(a):

WWT_{cost} = (41.36 F + 11,157,588) x RF x 0.898 = (41.36(200) + 11,157,588) x 1 = $10,026,942

Finally, the total capital investment (*TCI*) is calculated as follows:

$$TCI = \ 1.3 \times (\$48{,}868{,}764 + \$23{,}673{,}766 + \$14{,}536{,}123 + \$89{,}729{,}602 + \$10{,}026{,}942) = \$237{,}685{,}756$$

Total Annual Costs

Assuming the effective operating hours for the control is 8,000 hours per year, the annual variable costs for maintenance, operator labor, reagent, make-up water, waste disposal, auxiliary power and wastewater treatment are calculated using Equations 1.20, 1.21, 1.22, 1.23, 1.24, 1.25 and 1.26, respectively.

$$Annual\ Maintenance\ Cost\ (AMC) = 0.015\ \times \$237{,}685{,}756 = \$3{,}565{,}286$$

Annual Operator Labor Cost = 12 × 2,080 hours/year × $60/hour = $1,497,600

$$Annual\ reagent\ cost = 12.1\frac{tons}{hour} \times 30\frac{\$}{ton} \times \left[8{,}000\frac{hrs}{yr}\right] = \$2{,}904{,}208$$

$$Annual\ water\ cost = 37.9\frac{kgallons}{hour} \times \frac{1{,}000gallons}{kgallons} \times 0.0042\frac{\$}{gal} \times \left[8{,}000\frac{hrs}{yr}\right] = \$1{,}272{,}044$$

$$Annual\ waste\ disposal\ cost = 21.2\frac{tons}{hour} \times 30\frac{\$}{ton} \times \left[8{,}000\frac{hrs}{yr}\right] = \$5{,}098{,}515$$

$$Annual\ electricity\ cost = 8{,}470kW \times 0.0361\frac{\$}{kWh} \times 8{,}000\frac{hrs}{yr} = \$2{,}445{,}996$$

For plants with onsite landfills, the operating costs for chemical pretreatment, biological treatment, and ultrafiltration can be estimated as follows:

$Annual\ Wastewater\ Treatment\ Cost\ = (4.847(200) + 479{,}023) \times 0.958\ \times (8{,}000hrs/8760hrs) = \$419{,}939$

In addition to the annual costs, the facility will also have additional costs for replacing the mercury monitor once every six years. The estimated cost for the monitor is $100,000, which when annualized over a period of 6 years at 3.25% interest rate is:

$$Annual\ Costs\ for\ Mercury\ Monitor\ = 0.1861 \times \$100{,}000 = \$18{,}610$$

The total direct annual cost (DAC), the sum of the cost of the maintenance, labor, reagent, make-up water, waste disposal, auxiliary power, and wastewater treatment is given by the sum of the annual costs:

$$DAC = \$3{,}565{,}286 + \$1{,}497{,}600 + \$2{,}904{,}208 + \$1{,}272{,}044 + \$5{,}098{,}515 + \$2{,}445{,}996 + \$419{,}939 + \$18{,}610 = \$17{,}222{,}198/year$$

As discussed in Section 1.2.3.4, property taxes and overhead are both assumed to be zero, and insurance costs are assumed to be negligible. Thus, administrative charges and capital recovery are the only components of indirect annual costs estimated in this analysis. Administrative charges are calculated using Equation 1.28 as:

Annual Administrative Cost = 0.03 × ($1,497,600 + 0.4 × $3,565,286) = $87,711

Assuming an equipment life of 30 years and an interest rate of 3.25 percent, the capital recovery factor, CRF, is defined in Equation 1.30 as:

$$CRF = \frac{0.0325\ (1+0.0325)^{30}}{(1+0.03255)^{30}-1} = 0.0527$$

The capital recovery is calculated using Equation 1.29 as follows:

CR = CRF × TCI = 0.0527 × $237,685,756 = $12,526,039/year

The total indirect annual costs (IDAC) are calculated in Equation 1.27:

$$IDAC\ =\ AC + CR = \$87{,}711 +\ \$12{,}526{,}039 = \$12{,}613{,}750$$

The total annual cost is the sum of the direct and indirect annual costs:

$$TAC = DAC + IDAC = \$17{,}222{,}198 + \$12{,}613{,}750 = \$29{,}835{,}948$$

Cost Effectiveness

The total amount of SO_2 removed per year is calculated as follows:

$$SO_2\frac{removed}{yr} = \frac{13{,}538\frac{lb}{hour}\times\left[8{,}000\frac{hr}{yr}\right]}{2{,}000\frac{lb}{ton}} = 54{,}150\ tons/year$$

The annual cost in terms of SO_2 removed, or cost effectiveness, is calculated using Equation 1.32:

$$Annual\ Cost\ Effectiveness = \frac{TAC}{SO_2 removed} = \frac{\$29{,}835{,}948/year}{54{,}150\ tons/year} = \$551/ton$$

1.2.5 Dry Flue Gas Desulfurization Systems

This section focuses on dry FGD systems used to control emissions of acid gases, such as HCl, SO_2 and sulfur trioxide (SO_3). Dry FGD systems can be used to control gas streams with typical SO_2 pollutant concentrations ranging from 250 to 1,000 ppmv (approximately 2 lb/MMBtu). These systems provide removal efficiencies between 50% and 95%.

In the past, dry FGD systems were typically installed on smaller boilers, furnaces, and incinerators. However, in recent years, some newer dry FGD systems have been installed on combustion units larger than 500 MW (5,000 MMBtu/hour) burning bituminous and subbituminous coal. Dry FGD systems typically have lower capital and operating costs and require less space than wet FGD systems, but generally use more expensive types of sorbent. [3, 16]

1.2.5.1 Process Description

Three basic types of dry FGD systems are currently available: dry sorbent injection (DSI), spray dryer absorber (SDA), and a post combustion circulating dry scrubber (CDS). In this section, we describe the SDA and CDS systems. DSI systems are not covered in this chapter as mentioned earlier in Section 1.1.

Spray Drying Absorber Systems

The SDA system is a semi-dry system that injects an aqueous slurry with a high concentration of sorbent into the flue gas stream. A diagram of a typical SDA system is shown in Figure 1.3. The SDA system requires a separate absorber vessel. The high temperature of the flue gas entering the absorber chamber causes the water to evaporate and the pollutants in the flue gas are absorbed to and react with the surface of the sorbent. Water remaining on the solid sorbent surface promotes the reaction with the acid gases. However, the amount of water entering the absorber must be carefully controlled to avoid saturating the gas stream. Saturation can cause the solids to adhere to components inside the absorber and particulate collection device. To avoid the buildup of solids, the water fed into the spray dryer is controlled to avoid saturation of the flue gas but maintain high SO_2 removal. The flue gas in the absorber chamber is typically maintained so that it approaches but remains above adiabatic saturation to achieve optimal pollutant removal. Temperatures between 20 and 30°F above the saturation temperature are typically considered optimal. [14, 30]

SDA systems use either calcium or sodium-based sorbents. However, most SDA systems use lime as the sorbent since it is more reactive than limestone, easier to manage than limestone, and less expensive than sodium-based reagents. The typical SDA system operates at a lime to sulfur stoichiometry of 0.9 for combustion units burning low sulfur coals and 1.3 to 1.5 for units burning coal with high sulfur content. [3, 14]

An SDA system can achieve up to 95% SO_2 removal and reduce SO_2 emissions to as low as 0.03 lbs/MMBtu have been reported for coal-fired combustion. [31, 32] SDA systems are generally used for applications where the sulfur content of the coal burned is less than 3 lbs SO_2/MMBtu because the removal efficiency is poor for emission levels higher than 3 lbs SO_2/MMBtu. Combustion units that are retrofit with SDA systems lose the option of burning high sulfur content coal. [10]

Most of the waste is collected in a particulate control device (e.g., a fabric filter (FF) or an electrostatic precipitator (ESP)), while a small amount of waste also collects at the bottom of the absorber chamber. The waste sorbent mixture collected may be disposed at a landfill, sold as a byproduct, or recycled back to the sorbent feed tank. An ESP or baghouse (fabric filter) system is used to collect the waste generated. [14] Although ESP units are used in combination with SDA systems, calcium-based reagents can reduce the resistivity of particles and thereby reduce the efficiency of the ESP.

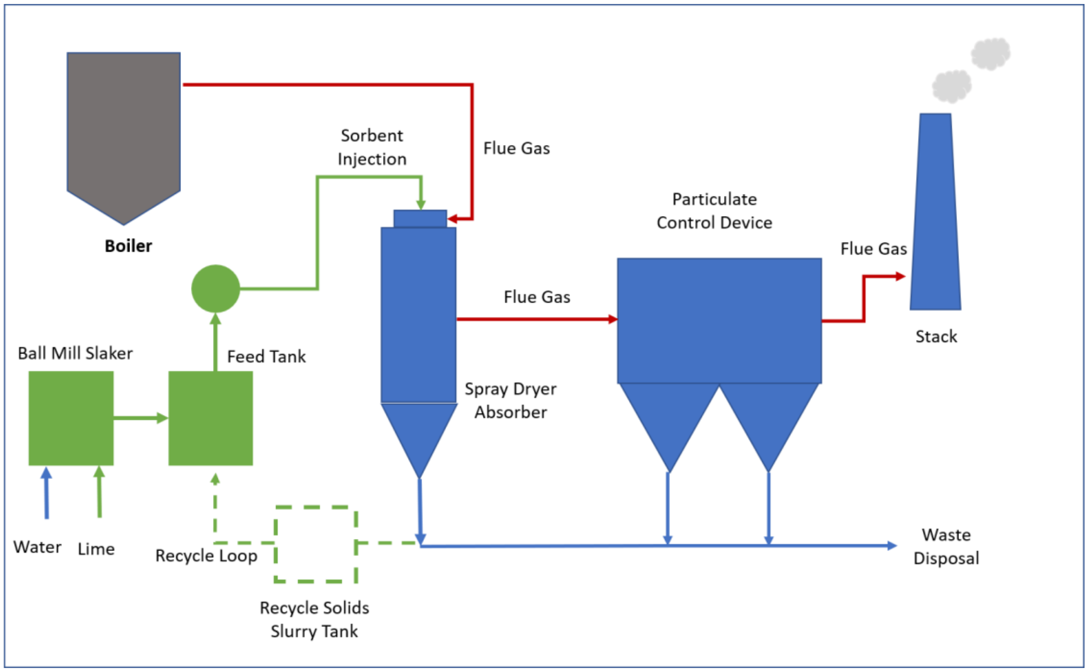

Figure 1.3: Diagram of a Typical SDA System [14]

SDA systems have lower energy consumption than wet FGD systems. One source reported energy consumption for an SDA system as 30% to 50% less than a comparable wet limestone FGD system. However, operating costs are higher due to the higher sorbent costs. [15]

Circulating Dry Scrubber

Higher removal efficiencies of greater than 98% for SO_2 are possible using a circulating dry scrubber (CDS). In the CDS system, the waste gas stream passes through an absorber vessel in which the solid sorbent is mixed with the waste gas stream. In one example of a CDS, the waste gas enters at the bottom of a reaction vessel through a venturi section, where the entering gas agitates the solid sorbent and creates a fluid bed of dry sorbent. In the fluid bed, contact between pollutants and sorbent is optimized, thereby promoting absorption of pollutant molecules onto the surface of the solid sorbent. The pollutant molecules then react with the sorbent to form salts. The waste solids are then collected in a particulate control device and the sorbent is returned to the absorber vessel. The absorption of acid gases is enhanced by the injection of a small amount of water into the fluid bed. The water allows the fluid bed to closely approach the adiabatic saturation temperature of the gas. The water rapidly evaporates allowing dry capture of the particulate downstream. Diagrams of a typical CDS system are shown in Figures 1.4 and 1.5. [14, 33, 34, 35]

CDS systems have seen only limited use in the United States but have been used in Europe to control SO_2 emissions from smaller combustion units ranging in size from 50 to 250 MW. [14] CDS systems can be used in applications where higher sulfur fuels are combusted and have low sorbent and water consumption. CDS systems have been applied to power plants, steel mills, refineries, waste-to-energy plants, combined heat and power plants. SO_2 removal efficiencies at 98% are reported by several vendors, while one vendor reported SO_2 removal as high as 99%. [33, 34, 35]

In 2015, CDS systems were installed at two power plants in the US: Lansing Generating Station in Lansing, Iowa and the Big Stone Power Plant in South Dakota. The CDS system installed at the Lansing Generating Station in Lansing, Iowa in 2015 is designed to achieve 98% SO_2 removal efficiency when the plant is burning Powder River Basin (PRB) coal.[33]

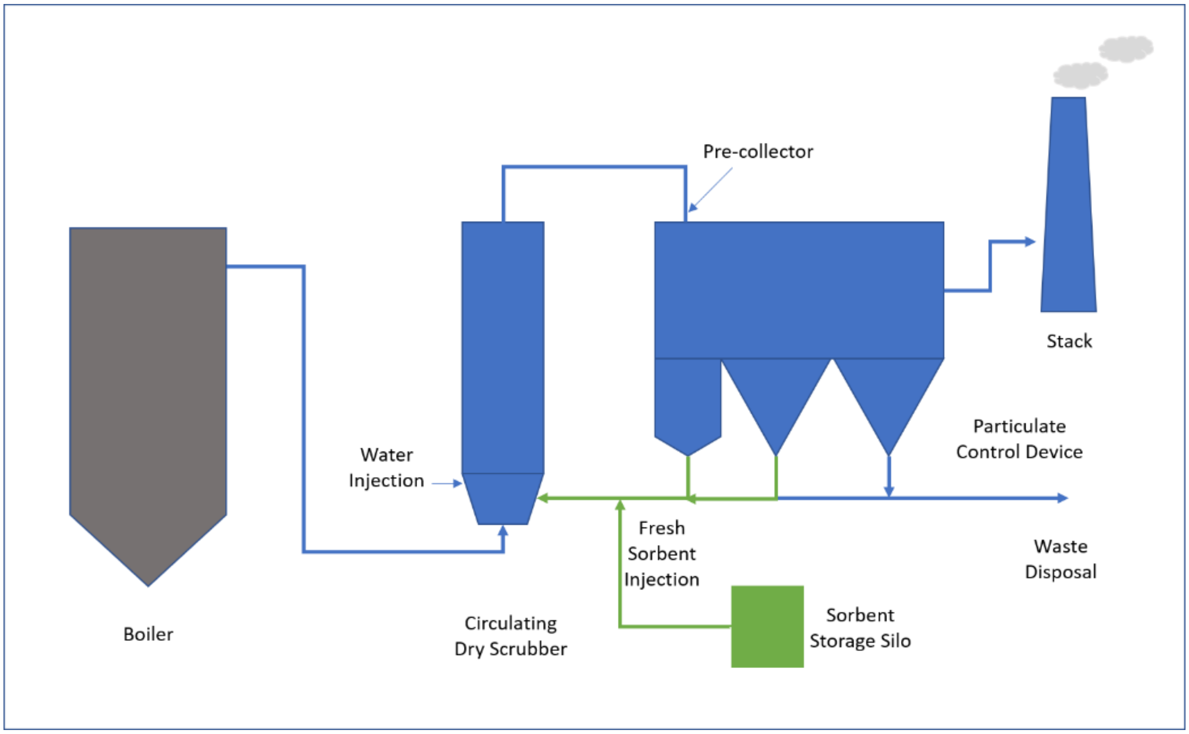

Figure 1.4: Diagram of a Typical CDS System [14]

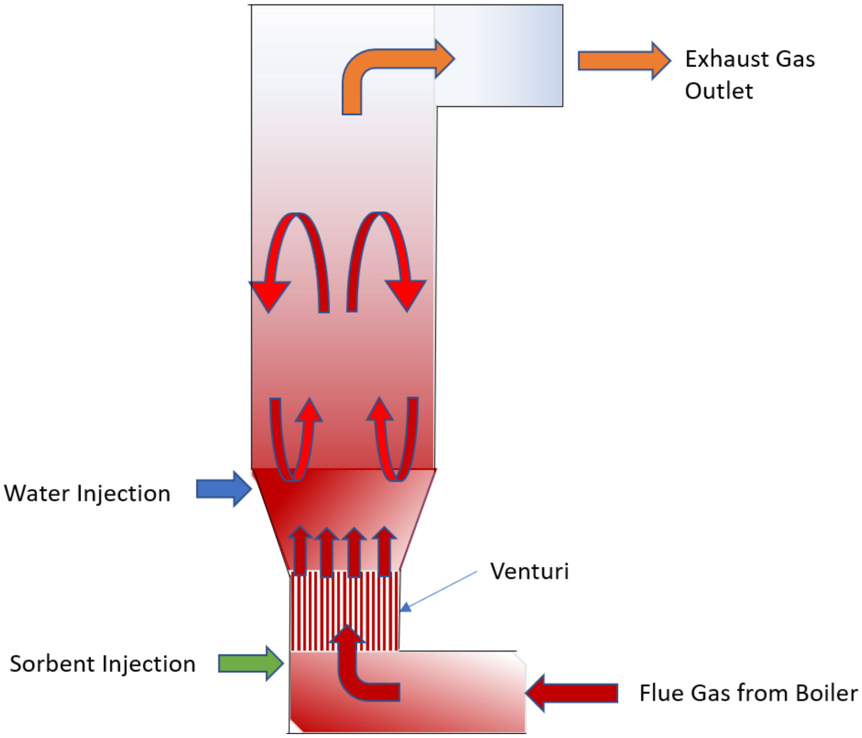

Figure 1.5: Cross-Section of a Typical CDS Absorber [33, 34]

Types of Sorbent

Dry FGD systems may use either calcium or sodium-based carbonates as sorbent reagents. However, lime is the most commonly used sorbent in SDA and CDS systems. FGD systems using lime typically add a small amount of water to improve removal efficiency. Hydrated lime increases the ash resistivity and the injected sorbent adds to the particulate loading of the precipitator or filter bag system.

System Efficiencies and Performance

The suitability and costs of dry FGD systems as pollution control methods are generally dependent on the following factors:

1) unit size;
2) required removal efficiency;
3) inlet pollutant (e.g. chlorides, sulfur and mercury) concentrations;
4) anticipated capacity factor; and
5) available space.

1.2.5.2 Design Criteria

The capital and operating cost for a dry FGD system varies depending on the following factors:

1) Original equipment installation or retrofit of existing boiler/furnace;
2) permitted emission rates for HCl, SO_2, SO_3, NO_X, and Hg;
3) fuel types and future fuel mix;
4) required removal efficiency; and,
5) selected sorbent and waste disposal cost.

These factors require a design concept that determines not just the current fuel and operational characteristics but how these might change in future operation.

1.2.5.3 Design Parameters for Study-Level Estimates for SDA FGD Control Systems

Methods for estimating the reagent feed rate, make-up water consumption, waste generation rate and auxiliary power consumption for an SDA system are presented in this section. These parameters are used in the cost methodology presented in Section 1.2.4.5 for estimating the annual operating costs for an SDA system. The methods presented in this section are those developed by the EPA’s CAMD for the cost methodology used with the IPM. [20]

Reagent Consumption

The reagent usage is a function of the boiler size, SO_2 emission rate and the removal efficiency of the SDA system. The estimated reagent usage was based on a flue gas temperature into the SDA system of 300°F and an adiabatic approach to saturation of 30°F. The calcium-to-sulfur stoichiometric ratio varies based on the inlet sulfur content; however, the variation in stoichiometric ratio was accounted for in the calculation method. The method is valid only up to 3 lb SO_2/MMBtu inlet emission rate. The lime is assumed to be 90% pure CaO, with the balance being inert material. Since the equation for lime consumption is based on data for systems with a 95% SO_2 removal efficiency, the IPM equation has been adjusted for the actual SO_2 removal achieved by the SDA system by the factor EF (removal efficiency)/0.95. The reagent usage rate is calculated as follows:

$$Q_{Lime} = \frac{[(0.6702 \times S^2) + (13.42 \times S)] \times A \times HRF}{2{,}000} \times \frac{EF}{0.95} \qquad (1.33)$$

Where

Q_{Lime} = lime consumption rate, tons/hour
S = SO_2 emission rate, lb/MMBtu
A = Unit size (Gross), MW
HRF = heat rate factor (calculated using Equation 1.6)
EF = removal efficiency, fraction.

Make-up Water

The makeup water rate is a function of gross unit size (actual gas flow rate) and SO_2 emission rate. The make-up water consumption rate for a utility boiler is calculated using the following equation:

$$q_{water} = \frac{[(0.04898 \times S^2) + (0.5925 \times S) + 55.11)] \times A \times CoalF \times HRF}{1{,}000} \qquad (1.34)$$

Where

q_{water} = make-up water consumption rate, 1,000 gallons/hour
S = SO_2 emission rate, lb/MMBtu
A = Unit size (Gross), MW
$CoalF$ = coal factor ($CoalF$ = 1 if bituminous; $CoalF$ = 1.05 if PRB; $CoalF$ = 1.07 if Lignite)
HRF = heat rate factor (calculated using Equation 1.6).

Waste Generation Rate

The design waste generation rate for SDA systems can be calculated as follows:

$$q_w = \frac{(0.8016 \times S^2 + 31.1917 \times S) \times A \times HRF}{2{,}000} \times \frac{EF}{0.95} \qquad (1.35)$$

Where:

q_w = waste generation rate, tons/hour
S = SO_2 emission rate, lb/MMBtu
A = Unit size (Gross), MW
HRF = heat rate factor (calculated using Equation 1.6)
EF = SO_2 removal efficiency, fraction
0.95 = SO_2 removal efficiency basis for model.

Auxiliary Power Costs

The additional auxiliary power required for the SDA system in kilowatts (kW) is estimated as a percent of the total unit gross production.

$$P = ([(0.000547 \times S^2) + (0.00649 \times S) + 1.3] \times CoalF \times HRF)/100 \times A \times 1{,}000 \quad (1.36)$$

Where:

P = electrical power consumption of the SDA system, kW
S = SO_2 emission rate, lb/MMBtu
0.000547 = constant
0.00649 = constant
A = Unit size (Gross), MW
$CoalF$ = coal factor (*CoalF* = 1 if bituminous*; CoalF* = 1.05 if PRB; *CoalF* = 1.07 if lignite)
HRF = heat rate factor (calculated using Equation 1.6)
100 = conversion factor, percent to decimal fraction
1,000 = conversion factor, kW/MW.

1.2.5.4 Capital Costs for SDA Control Systems

This section presents a methodology for estimating the capital and annual costs for SDA systems. The cost equations are based on those developed by the EPA's CAMD as part of the IPM version 6 (IPM v6). [20] The methodology provides an approach for estimating study-level capital and annual costs. The capital costs are in 2016 dollars. The IPM algorithms are based on cost data from S&L's in-house database of SDA projects and various industry publications. [20] The capital cost algorithms are valid for units of 50 MW or more. Capital costs for units smaller than 50 MW are approximately $1,000/kW.

The IPM does not provide cost equations for CDS systems; however, the capital costs for a CDS system are similar to the SDA system for a combustion unit of the same size and sulfur emission rate.

The IPM is based primarily on data for retrofits, which are typically 30% more than for new units of the same size and design. To correct for the additional costs associated with retrofits, we have included a retrofit factor in the TCI equations. A retrofit factor (*RF*) of 0.77 should be used for new construction and a retrofit factor of 1 should be used for retrofits to existing units, where the retrofit is of an average level of difficulty. Retrofit costs vary significantly from site to site and depend on the amount of space available and whether modifications to existing equipment (e.g., ductwork, stack) are necessary. A higher retrofit factor should be used for sites that are congested or where extensive modifications are required for existing systems.

The capital cost equations are presented in this Section 1.2.4.4 and the methods for estimating the annual costs are provided in Section 1.2.4.5. The total capital investment (TCI) includes direct and indirect costs associated with purchasing and installing the control equipment. TCI includes the equipment cost for the SDA, the cost of auxiliary equipment, the cost of a fabric filter/baghouse, direct and indirect installation costs, costs for buildings and site preparation, cost of land and working capital. The capital costs assume the installation is completed using multiple lump sum contracts. Costs for turnkey (where the project price is fixed at the time a construction contract is signed) engineering procurement construction contracts may be 10% to 15% higher than those calculated using the IPM equations. [20]

The TCI does not include the owner's costs (for owner activities related to engineering, management, and procurement) and costs such as allowance for funds used during construction (AFUDC). Although these are capital cost items included in the IPM cost estimates, they are excluded from the cost methodologies presented in the EPA Control Cost Manual because they are inconsistent with the overnight cost method[9] that is a key basis for the Control Cost Manual methodology. A more detailed discussion of capital costs can be found in Section 1, Chapter 2 of this Manual.

Table 1.5 shows the installed costs reported to the U.S. Department of Energy, Energy Information Administration (EIA). The EIA data includes dry scrubbers with a removal efficiency of 80% or greater. Table 1.5 also shows the capital costs estimated for an SDA and CDS system installed on a 500 MW unit calculated using the IPM methodology. For the SDA and CDS systems, the IPM estimates are higher than the average costs but are well below the highest costs reported to EIA.

Table 1.5: Comparison of the IPM Capital Cost Methodology Against Installed Costs Reported to the Energy Information Administration [18]

		EIA Cost Data				**IPM TCI**
Fuel Type	**Scrubber Type**	**Median $/KWh**	**Average $/KWh**	**Minimum $/KWh**	**Maximum $/KWh**	**Estimated $/KWh**
Subbituminous	CDS	215	281	65	665	499
Bituminous	CDS	248	288	182	434	No data

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[9] The overnight cost estimation method presumes costs are incurred as if the project in question incurred no interest during construction, or was built "overnight." Another description of this method is the present value cost that would have to be paid as a lump sum up front to completely pay for a construction project. For more information, see "Conducting Technical and Economic Evaluations – As Applied for the Process and Utility Industries," Recommended Practice 16R-90, American Association of Cost Engineering International. April 1991, or Section 1, Chapter 2 of this Control Cost Manual.

Subbituminous	SDA	178	220	19	606	499
Bituminous	SDA	188	377	135	923	No data

Total Capital Investment for SDA FGD Control Systems

The equipment required for an SDA system using lime as sorbent varies by the pollutant and the emission requirements of the source. The algorithms presented in this section are based on the typical SDA system and are appropriate for facilities that are using coal types with less than 3 lb SO_2/MMBtu achieving a removal efficiency of 95%.

The typical SDA system using pebble lime consists of:

1) Receiving and storage silos;
2) Conveying equipment for lime unloading;
3) Lime feeder;
4) Lime slaker;
5) Lime slurry tank;
6) Slurry pump;
7) Slurry metering pump;
8) Dust control filtering systems;
9) Spray dryer absorber vessel;
10) Atomizers;
11) Waste extraction system;
12) Compressed air;
13) Valves; and
14) Data Control System.

The total capital investment costs for an SDA system include the costs for the absorber island, reagent preparation equipment, baghouse, waste recycling/handling facilities, and additional costs associated with installation. These additional costs of installation are called "balance of plant (BOP) costs" and include the costs for additional duct work, piping, buildings, foundations, electrical wiring and booster fans. The TCI calculation for the SDA system includes a factor of 1.3 to estimate (1) engineering and construction management costs, (2) labor adjustment for overtime pay, and (3) contractor profit and fees. Each of these 3 items is estimated to be 10% of the TCI.

The total capital investment cost for an SDA system is given by the Equation 1.37:

$$TCI = 1.3 \times \ (ABS_{Cost} + BMF_{Cost} + BOP_{Cost}) \tag{1.37}$$

Where:

TCI = total capital investment for an SDA FGD system, \$
ABS_{Cost} = cost of the SDA FGD system absorber island, \$
BMF_{Cost} = reagent preparation and waste recycle/handling equipment cost, \$
BOP_{Cost} = balance of plant costs, \$.

The costs for the absorber island and baghouse (ABS_{Cost}), reagent preparation and waste recycle/handling equipment (BMF_{cost}) and balance of plant costs (BOP_{Cost}) are discussed below. The capital cost for an SDA control system varies depending on the size of the unit. The gross heat rate impacts the amount of flue gas generated and ultimately impacts the size of the absorber, reagent preparation, waste handling, and the BOP costs. The SO_2 emissions rate primarily impacts the capital costs of the reagent and waste handling facilities, while the type of fuel burned impacts the quantity of flue gas and thereby impacts the BOP_{Cost}.

The ABS_{Cost} and BOP_{Cost} are also impacted by the unit's elevation with respect to sea level. The original cost calculations were developed for SDA FGD systems located within 500 feet of sea level. Therefore, for SDA FGD systems located at higher elevations, the ABS_{Cost} and BOP_{Cost} should be increased by multiplying by the elevation factor, *ELEVF*, which is calculated using Equations 1.7 and 1.8 in Section 1.2.2.4. [20]

Capital Costs for Coal-fired Combustion Units >600 MW (6,000 MMBtu/hour)

The capital costs for the SDA FGD system absorber island and baghouse are calculated as follows:

$$ABS_{Cost} = \ A \times 98{,}000 \times ELEVF \tag{1.38}$$

Where:

ABS_{Cost} = SDA FGD system absorber island and baghouse costs, \$
98,000 = constant
A = Unit size (Gross), MW
$ELEVF$ = elevation factor (calculated using Equations 1.7 and 1.8, if plant is located above 500 feet above sea level; ELEVF = 1 for plants located at 500 ft above sea level).

The costs for the reagent preparation and waste recycle/handling costs for SDA FGD systems greater than 600 MW (6,000 MMBtu/hour) are calculated as follows:

$$BMF_{Cost} = \ A \times 52{,}000 \tag{1.39}$$

Where:

BMF_{Cost} = reagent preparation and waste recycle/handling cost, $
52,000 = constant in the equation
A = Unit size (Gross), MW.

The balance of plant costs (BOP_{cost}) for coal-fired units > 600 MW (6,000 MMBtu/hour) include costs for piping, water treatment for dilution water, ductwork, auxiliary power modifications, and other electrical and site upgrades that are typically necessary as part of the installation of the SDA FGD system. The BOP costs are calculated as follows:

$$BOP_{Cost} = 138{,}000 \times A \times ELEVF \tag{1.40}$$

Where:
BOP_{Cost} = Balance of plant costs, $
138,000 = constant in the equation
A = Unit size (Gross), MW
$ELEVF$ = elevation factor (calculated using Equations 1.7 and 1.8, if plant is located above 500 feet above sea level; *ELEVF* = 1 for plants located at 500 ft above sea level).

Capital Costs for Coal-fired Combustion Units Between 50 and 600 MW (500 and 6,000 MMBtu/hour)

For coal-fired utility units between 50 and 600 MW (500 to 6,000 MMBtu/hour), the capital costs for the SDA FGD absorber island vary based on the type of coal burned and the sulfur dioxide emission rate. The capital costs of the absorber island, including the costs for a baghouse, are calculated as follows:

$$ABS_{Cost} = 637{,}000 \times (A)^{0.716} \times (CoalF \times HRF)^{0.6} \times (S/4)^{0.01} \times ELEVF \times RF \tag{1.41}$$

Where:

ABS_{Cost} = SDA FGD absorber island costs (includes cost of an absorber and a baghouse), $
637,000 = constant in the equation
A = Unit size (Gross), MW
HRF = heat rate factor
$CoalF$ = coal factor (*CoalF* = 1 if bituminous; *CoalF* = 1.05 if PRB; *CoalF* = 1.07 if Lignite)
$ELEVF$ = elevation factor (calculated using Equations 1.7 and 1.8, if the plant is located above 500 feet above sea level; ELEVF = 1 for plants located at as below 500 ft above sea level)
S = SO_2 emission rate, lb/MMBtu
RF = retrofit factor (RF = 0.77 for new construction; RF = 1 for retrofits with average level of difficulty).

The coal factor, *CoalF*, depends on the type of coal burned. A factor of 1 is used for bituminous coal, 1.05 for Powder River Basin (PRB) coal, and 1.07 for lignite coal. The heat rate factor (HRF) is calculated using Equation 1.6.

The reagent preparation and waste recycle/handling costs (BMF_{Cost}) are for SDA FGD systems between 50 and 600 MW (500 and 6,000 MMBtu/hour) are calculated as follows:

$$BMF_{Cost} = 338{,}000 \times (A)^{0.716} \times (S \times HRF)^{0.2} \times RF \tag{1.42}$$

Where:

BMF_{Cost} = reagent preparation and waste recycle/handling cost, \$
338,000 = constant in the equation
HRF = heat rate factor (calculated using Equation 1.6)
S = SO_2 emission rate, lb/MMBtu
RF = retrofit factor (RF = 0.77 for new construction; RF = 1 for retrofits with average level of difficulty)
A = Unit size (Gross), MW.

The balance of plant costs (BOP_{cost}) vary based on the size of the unit, type of coal burned, and the elevation of the unit. The BOP_{cost} are calculated as follows:

$$BOP_{Cost} = 899{,}000 \times (A)^{0.716} \times (CoalF \times HRF)^{0.4} \times ELEVF \times RF \tag{1.43}$$

Where:

BOP_{Cost} = Balance of plant costs, \$
899,000 = constant in the equation
A = Unit size (Gross), MW
$CoalF$ = coal factor (*CoalF* = 1 if bituminous; *CoalF* = 1.05 if PRB; *CoalF* = 1.07 if Lignite)
HRF = heat rate factor (calculated using Equation 1.6)
$ELEVF$ = elevation factor (calculated using Equations 1.7 and 1.8 if plant is located above 500 feet above sea level; *ELEVF* = 1 for plants located at 500 ft above sea level)
RF = retrofit factor (RF = 0.77 for new construction; RF = 1 for retrofits with average level of difficulty).

1.2.5.5 Total Annual Costs for SDA Systems

The annual costs for controls comprise of the direct and indirect annual costs. The methods for estimating the direct annual costs of operating the control system, such as the cost of reagent and operating labor, are presented in this section. The indirect annual costs are the fixed costs associated with owning and operating the control system and include the annualized capital recovery cost and the costs for insurance and overhead. Methods for calculating the indirect costs also are presented in this section. The methods presented in this section are those developed specifically for SDA control systems by the EPA's CAMD for the IPM cost methodology. [20]

Direct Annual Costs

Direct annual costs (*DAC*) include variable and semi-variable costs. Variable direct annual costs account for purchase of reagent, utilities (electrical power and water), and any additional waste disposal costs resulting from the operation of the dry scrubber. Semi-variable direct annual costs include operating and supervisory labor and maintenance (labor and materials). For the SDA FGD system, DAC is calculated as follows:

$$DAC = \begin{pmatrix} Annual \\ Maintenance \\ Cost \end{pmatrix} + \begin{pmatrix} Annual \\ Operator \\ Cost \end{pmatrix} + \begin{pmatrix} Annual \\ Reagent \\ Cost \end{pmatrix} + \begin{pmatrix} Annual \\ Make-up \\ Water \\ Cost \end{pmatrix} + \begin{pmatrix} Annual \\ Waste \\ Disposal \\ Cost \end{pmatrix} + \begin{pmatrix} Annual \\ Auxiliary \\ Power \\ Cost \end{pmatrix} \tag{1.44}$$

Annual Maintenance Cost. The annual maintenance cost (*AMC*) includes the costs for labor and material. For an SDA FGD system, the *AMC* is estimated to be 1.5%. This is a standard percentage for maintenance on emissions control devices. The equation for annual maintenance cost for an SDA FGD system is given by:

$$AMC = 0.015 \times TCI \tag{1.45}$$

Annual Operator Cost. Therefore, the cost of operator labor is calculated as follows:

$$Annual\ Operator\ Labor\ Cost = FT \times 2{,}080 \times Hourly\ Labor\ Rate \tag{1.46}$$

where FT is the number of additional full-time personnel required to operate the control system and 2,080 is the annual number of work hours per full time employee. Generally, eight additional full-time personnel are required to operate an SDA FGD system.

Annual Reagent Cost. The annual cost of reagent (AR_{cost}) is estimated using the reagent consumption rate, the total effective operating time (t_{op}) and the cost of the reagent ($Cost_{Lime}$). The lime consumption rate (Q_{Lime}) is calculated using Equation 1.31. The total effective operating time (t_{op}) is calculated using Equations 1.3 through 1.5 in Section 1.2.2.2.

$$AR_{Cost} = Q_{Lime} \times Cost_{Lime} \times t_{op} \tag{1.47}$$

Where:

AR_{cost} = annual lime cost, \$/year
Q_{Lime} = reagent consumption rate, tons/hour
$Cost_{Lime}$ = purchase price of reagent, \$/ton.
t_{op} = effective number of operating hours, hour/year.

Annual Waste Disposal Cost. Using the estimated waste generation rate calculated in Equation 1.35, the annual cost of wastewater disposal is given by:

$$Annual\ waste\ disposal\ cost = q_{waste} \times Cost_{disposal} \times t_{op} \tag{1.48}$$

Where:

q_{waste} = waste generation rate, tons/hour
$Cost_{disposal}$ = unit cost of waste disposal, \$/ton
t_{op} = effective number of operating hours, hours/year.

Annual Auxiliary Power Cost. Using the estimated power consumption from Equation 1.36, the annual cost of electricity is calculated as follows:

$$Annual\ electricity\ cost = P \times Cost_{elec} \times t_{op} \tag{1.49}$$

Where:

P = electrical power consumption for the dry scrubber, kW
$Cost_{elec}$ = unit price of electricity, \$/kWh
t_{op} = effective number of operating hours, hours/year.

Annual Make-up Water Cost for SDA Systems. Using the estimated make-up water consumption rate calculated in Equation 1.34, the annual cost of make-up water in \$/yr is given by:

$$Annual\ water\ cost = q_{water} \times Cost_{water} \times t_{op} \tag{1.50}$$

Where:

q_{water} = make-up water consumption rate, kgallons (1,000 gallons)/hour
$Cost_{water}$ = unit cost of water, \$/gallon
t_{op} = number of operating hours, hours/year.

Indirect Annual Costs for SDA FGD Control Systems

In general, indirect annual costs are fixed costs that include the capital recovery cost, property taxes, insurance, administrative charges, and overhead. The capital recovery cost is based on the anticipated equipment lifetime and the annual interest rate employed. As mentioned earlier in this chapter, acid gas scrubber systems are expected to have an equipment lifetime of 30 years. However, for retrofits on older combustion units, the remaining life of the controlled combustion unit may be an important factor for determining the expected lifetime for a dry scrubber.

For this analysis, taxes are assumed to be zero since property taxes generally do not apply to air pollution control equipment given that purchase of additional land is unusual for installation of an add-on control device. Insurance for an absorber system is expected to be minimal. The cost of overhead for an SDA FGD system is also assumed to be zero since control systems add only minimal overhead costs to those already incurred by the plant. Overhead includes payroll overhead (i.e., expenses related to labor employed in operation and maintenance of hardware) and plant overhead (i.e., costs for plant protection, control laboratories, and parking areas).

Using these assumptions, indirect annual costs in \$/yr, *IDAC*, consist of both administrative charges and capital recovery, which can be expressed as:

$$IDAC = AC + CR \tag{1.51}$$

where *AC* represents the administrative charges and *CR* represents the capital recovery cost. Administrative charges may be calculated as:

$$AC = 0.03 \times (Annual\ Operator\ Cost + 0.4\ x\ Annual\ maintenance\ cost) \tag{1.52}$$

Capital recovery is estimated as:

$$CR = CRF \times TCI \tag{1.53}$$

where TCI is the total capital investment in dollars and CRF is the capital recovery factor. Capital recovery factor is a function of the equipment life and the opportunity cost of the capital (i.e., interest rate). The CRF is defined as:

$$CRF = \frac{i(1+i)^n}{(1+i)^n - 1} \tag{1.54}$$

where *i* is the assumed interest rate and *n* is the equipment life of the dry scrubber system.

Total Annual Cost

The total annual cost, *TAC*, for owning and operating a dry scrubber is the sum of direct and indirect annual costs as given in the following equation:

$$TAC = DAC + IDAC \tag{1.55}$$

1.2.5.6 Cost Effectiveness

The cost in dollars per ton of SO_2 removed per year, is calculated as follows:

$$Cost\ Effectiveness\ = \frac{TAC}{SO_2\ Removed/year} \tag{1.56}$$

Where:

Cost Effectiveness = the cost effectiveness, \$/ton
SO_2 Removed/year = annual mass of SO_2 removed by the absorber, tons/yr
TAC = total annual cost, \$/year.

1.2.5.7 Example Problem

This section illustrates how to calculate the design parameters and capital and annual costs for the retrofit of an existing 500 MW utility boiler burning bituminous coal with an SDA FGD system. This example is purely illustrative, and is not meant to serve in place of the actual data on a SDA FGD system that is included in an actual cost analysis.

Design Basis for Example Project

For the purposes of this example, the 500 MW boiler is located at approximately 500 feet above sea level and the retrofit is assumed to be of typical complexity with the site having sufficient space available to install the absorber, reagent preparation and storage, and wastewater handling facilities. The following assumptions are made to perform the calculations:

Boiler MW rating full load capac	500 MW
Coal consumption rate, q_{fuel}	326,600 lb/hour
High heating value of bituminous coal, *HHV*	15,000 Btu/lb
Gross heat input rate, *GHR*	9.8 MMBtu/MWh
Uncontrolled SO_2 emission rate, S_{in}	2.0 lb/MMBtu
Required controlled SO_2 emission rate, S_{out}	0.1 lb/MMBtu
Annual effective operating hours, t_{op}	8,000 hours/year
Retrofit factor, *RF*	1
Elevation factor, *ELEVF*	1

In addition to these assumptions regarding the physical parameters of the boiler, the economic factors for the example cost calculations are as follows:

Cost Year	2016
Equipment Life	30 years
Annual Interest Rate[10]	3.25%
Lime cost[11]	125 $/ton
Water Consumption Cost[12]	0.0042 $/gal
Electricity Cost[13]	0.0361 $/kWh
Waste Disposal Cost[15]	30 $/ton
Labor rate[15]	60 $/hour

---

[10] As stated in Section 1, Chapter 2 of the Cost Manual, the interest rate that is appropriate for annualizing capital cost is either the bank prime rate (currently 3.25%), which is an interest rate set by the Federal Reserve Board that fluctuates with the market for financial credit, or a firm-specific rate that reflects the rates of debt and equity for the firm owning the unit at issue if that firm can justify this alternative rate. The bank prime rate serves as a default if the firm can, or decides not to, provide an interest rate reflecting its rates for debt borrowing and/or equity.

[11] U.S. Environmental Protection Agency (EPA). Documentation for EPA's Power Sector Modeling Platform v6 Using the Integrated Planning Model. Office of Air and Radiation. January 2017. Available at: https://www.epa.gov/airmarkets/documentation-epas-power-sector-modeling-platform-v6.

[12] Average water rates for industrial facilities in 2016 compiled by Black & Veatch. See "50 Largest Cities Water/Wastewater Rate Survey – 2018-2019." Available at https://www.bv.com/sites/default/files/2019-10/50_Largest_Cities_Rate_Survey_2018_2019_Report.pdf.

[13] U.S. Energy Information Administration. Electric Power Annual 2016. Table 8.4. Published December 2017. Available at: https://www.eia.gov/electricity/annual/pdf/epa.pdf.

<u>Design Parameters for Example Retrofit Project</u>

<u>Boiler Heat Rate</u>. The boiler annual heat input rate, Q_B, is calculated from the *HHV* for bituminous coal given in Table 1.4 and the maximum fuel consumption rate, q_{fuel}, using Equation 1.1:

$$Q_B = \frac{15{,}000\,\frac{Btu}{lb} \times 326{,}600\,\frac{lb}{hr}}{10^6\,\frac{Btu}{MMBtu}} = 4{,}899\,\frac{MMBtu}{hr}$$

<u>SO_2 Removal Efficiency</u>. The SO_2 removal efficiency, *EF*, is calculated from the inlet SO_2 emission level and the required stack SO_2 emission level:

$$EF = \left[\frac{SO_{2In} - SO_{2Out}}{SO_{2In}}\right] \times 100$$

$$EF = \left[\frac{2.0\,\frac{lb}{MMBtu} - 0.10\,\frac{lb}{MMBtu}}{2.0\,\frac{lb}{MMBtu}}\right] \times 100 = 95\%$$

The SO_2 removed per hour (lb/hr) is calculated from the inlet SO_2 emission level, the SO_2 removal efficiency, and the maximum heat rate to the boiler:

$$SO_2\ Removed/hr = SO_{2in} \times EF \times Q_B$$

$$SO_x\ Re\,m\,oved/hr = 2.0\,\frac{lb}{MMBtu} \times 0.95 \times \frac{4{,}899 MMBtu}{hour} = 9{,}308\,\frac{lb}{hr}$$

<u>Reagent Consumption Rate</u>. The lime consumption rate is calculated using Equation 1.33:

$$Q_{Lime} = \frac{[(0.6702 \times S^2) + (13.42 \times S)] \times A \times HRF}{2{,}000} \times \frac{EF}{0.95}$$

$$Q_{Lime} = \frac{[(0.6702 \times (2.0)^2) + (13.42 \times 2.0)] \times 500 \times 0.98}{2{,}000} \times \frac{0.95}{0.95} = 7.23\ tons/hour$$

<u>Make-up Water Consumption</u>. The make-up water consumption rate is calculated using Equation 1.34:

$$q_{water} = \frac{[(0.04898 \times S^2) + (0.5925 \times S) + 55.11)] \times A \times CoalF \times HRF}{1{,}000}$$

$$q_{water} = \frac{[(0.04898 \times (2.0)^2) + (0.5925 \times 2.0) + 55.11)] \times 500 \times 1 \times 0.98}{1{,}000}$$

$$= 27.7\ kgallons/hour$$

Waste Generation. Using Equation 1.35, design waste generation rate for SDA systems can be calculated as follows:

$$\mathrm{q_{waste}} = \frac{(0.8016 \times \mathrm{S}^2 + 31.1917 \times \mathrm{S}) \times \mathrm{A} \times \mathrm{HRF}}{2{,}000} \times \frac{\mathrm{EF}}{0.95}$$

$$q_{waste} = \frac{[0.8016 \times (2.0)^2 + (31.1917 \times 2.0)] \times 500 \times 0.98}{2{,}000} \times \frac{0.95}{0.95} = 16.1\ tons/hour$$

Electricity Consumption. The auxiliary power required for the SDA FGD is estimated using Equation 1.36:

$$\mathrm{P} = ([(0.000547 \times \mathrm{S}^2) + (0.00649 \times \mathrm{S}) + 1.3] \times \mathrm{CoalF} \times \mathrm{HRF})/100 \times \mathrm{A} \times 1{,}000$$

$$P = ([(0.000547 \times (2.0)^2) + (0.00649 \times 2.0) + 1.3] \times 1 \times 0.98)/100 \times 500 \times 1{,}000$$

$$P = 6{,}444\ kW$$

Cost Estimation for Example Retrofit Project

Once the SDA FGD system is sized, the capital and annual costs can be estimated. The total capital investment costs (TCI) are estimated using Equation 1.37:

$$TCI = 1.3 \times \ (ABS_{Cost} + BMF_{Cost} + BOP_{Cost})$$

Where:

TCI = total capital investment for an SDA FGD system, $
ABS_{Cost} = cost of the SDA FGD system absorber island and baghouse, $
BMF_{Cost} = reagent preparation and waste recycle/handling equipment cost, $
BOP_{Cost} = balance of plant costs, $.

The capital costs for the absorber island and baghouse are estimated using Equation 1.38:

$$ABS_{Cost} = \ 637{,}000 \times (A)^{0.716} \times (CoalF \times HRF)^{0.6} \times (S/4)^{0.01} \times ELEVF \times RF$$

$$ABS_{Cost} = \ 637{,}000 \times (500)^{0.716} \times (1.0 \times 0.98)^{0.6} \times (2/4)^{0.01} \times 1 \times 1 = \$53{,}496{,}738$$

The costs for the reagent preparation and waste recycle/handling costs can be calculated using Equation 1.39:

$$BMF_{Cost} = \ 338{,}000 \times (A)^{0.716} \times (S \times HRF)^{0.2} \times RF$$

$$BMF_{Cost} = \ 338{,}000 \times (500)^{0.716} \times (2.0 \times 0.98)^{0.2} \times 1 = \$33{,}100{,}186$$

The BOP_{Cost} can be calculated using Equation 1.40:

$$BOP_{Cost} = 899{,}000 \times (A)^{0.716} \times (CoalF \times HRF)^{0.4} \times ELEVF \times RF$$

$$BOP_{Cost} = 899{,}000 \times (500)^{0.716} \times (1 \times 0.98)^{0.4} \times 1 \times 1 = \$76{,}333{,}055$$

The total capital investment can be calculated using the values above:

$$TCI = 1.3 \times (\$53{,}496{,}738 + \$33{,}100{,}186 + \$76{,}333{,}055) = \$211{,}808{,}973$$

Assuming the effective operating hours for the SDA FGD system are 8,000 hours per year, then the annual variable costs for maintenance, operator labor, reagent, waste, auxiliary power, and make-up water can be calculated using Equations 1.45, 1.46, 1.47, 1.48, 1.49 and 1.50, respectively.

$$AnnualMaintenanceCost(AMC) = 0.015 \times \$211{,}808{,}973 = \$3{,}177{,}135$$

Annual Operator Labor Cost = 8 × 2080 hours/year× $60/hour = $998,400

$$Annual\ reagent\ cost = 7.23\frac{tons}{hour} \times 125\frac{\$}{ton} \times \left[8{,}000\frac{hrs}{yr}\right] = \$7{,}232{,}596$$

$$Annual\ waste\ disposal\ cost = 16.1\frac{tons}{hour} \times 30\frac{\$}{ton} \times \left[8{,}000\frac{hrs}{yr}\right] = \$3{,}856{,}680$$

$$Annual\ electricity\ cost = 6{,}444kW \times 0.0361\frac{\$}{kWh}\left[8{,}000\frac{hrs}{yr}\right] = \$1{,}861{,}121$$

$$Annual\ water\ cost = 27.7kgph \times \frac{1{,}000gal}{kgal} \times 0.0042\frac{\$}{gal} \times \left[8{,}000\frac{hrs}{yr}\right] = \$930{,}067$$

The total direct annual cost (DAC), the sum of the cost of the maintenance, labor, reagent, waste disposal, auxiliary power, and make-up water is given by the sum of the annual costs:

$$\begin{aligned} DAC &= \$3{,}177{,}135 + \$998{,}400 + \$7{,}232{,}596 + \$3{,}856{,}680 + \$1{,}861{,}121 + \$930{,}067 \\ &= \$18{,}055{,}998 \end{aligned}$$

Property taxes and overhead are both assumed to be zero, and insurance costs are assumed to be negligible. Thus, administrative charges and capital recovery are the only components of indirect annual costs estimated in this analysis. Administrative charges are calculated using Equation 1.52 as:

Annual Administrative Cost = 0.03 × ($998,400 + 0.4 × $3,177,135) = $68,078

The capital recovery factor, CRF, is a function of the equipment life (30 years) and the opportunity cost of the capital (i.e., interest rate), defined in Equation 1.54:

$$CRF = \frac{0.0325\ (1 + 0.0325)^{30}}{(1 + 0.0325)^{30} - 1} = 0.0527$$

The capital recovery is calculated using Equation 1.53 as follows:

$$CR = 0.0527 \times \$211{,}808{,}973 = \$11{,}162{,}333/year$$

The total indirect annual costs (IDAC) are calculated in Equation 1.51:

$$IDAC\ \ =\ \ AC + CR = \$68{,}078 +\ \ \$11{,}162{,}333\ \ =\ \ \$11{,}230{,}411$$

The total annual cost is the sum of the direct and indirect annual costs:

$$TAC = DAC + IDAC = \$18{,}055{,}998 + \$11{,}230{,}333 = \$29{,}286{,}408$$

The total amount of SO_2 removed is calculated as follows:

$$Annual\ SO_2\ Removed = \frac{2.0\frac{lb}{MMBtu} \times 0.95 \times 4{,}900\frac{MMBtu}{hr} \times\ 8{,}000\frac{hrs}{yr}}{2{,}000\frac{lb}{ton}} = 37{,}232\ tons/year$$

The annual cost in terms of SO_2 removed, or cost effectiveness, for the SDA retrofit is calculated using Equation 1.56:

$$Annual\ Cost\ Effectiveness = \frac{\$29{,}286{,}408}{37{,}232\ tons} = \$787/ton$$

1.3 Wet Packed Tower Gas Absorbers

Packed tower absorbers are the most common approach to removing acidic gases from flue gas. The large wetted area provided by the packing makes it possible to achieve very high removal efficiencies at a relatively low capital and operating costs. In this section, we describe the process, outline how to determine the critical design characteristics, and provide a method for estimating the capital and operating costs. The methods outlined in this section can be used for estimating costs for packed tower absorbers used to control flue gas containing any acidic pollutants, such as HCl and HF. The method can be used as an alternative approach to the method provided in Section 1.2.3 for estimating capital and operating costs for packed tower FGD systems.

1.3.1 Process Description

1.3.1.1 Packed Tower Designs

A basic packed tower absorber is comprised of a column shell, mist eliminator, liquid distributors, packing materials, packing support, and may include a packing restrainer. Corrosion resistant alloys or plastic materials such as polypropylene are required for column internals when highly corrosive solvents or gases are used. A schematic drawing of a countercurrent packed tower is shown in Figure 1.6. In this figure, the packing is separated into two sections. This configuration is more expensive than designs where the packing is not so divided.[6]

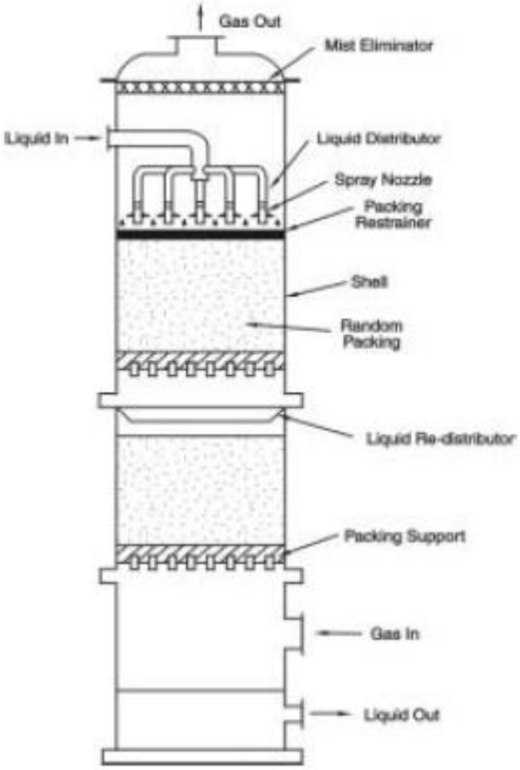

Figure 1.6: Packed Tower for Gas Absorption

The tower shell may be made of steel or plastic, or a combination of these materials depending on the corrosiveness of the gas and liquid streams, and the process operating conditions. One alloy that is chemical and temperature resistant or multiple layers of different, less expensive materials may be used. The shell is sometimes lined with a protective membrane, often made from a corrosion resistant polymer. For absorption involving acid gases, an interior layer of acid resistant brick provides additional chemical and temperature resistance.[36]

At high gas velocities, the gas exiting the top of the column may carry droplets of liquid as a mist. To prevent this, a mist eliminator in the form of corrugated sheets or a layer of mesh can be installed at the top of the column to collect the liquid droplets, which coalesce and fall back into the column.

A liquid distributor is designed to wet the packing bed evenly and initiate uniform contact between the liquid and vapor. The liquid distributor must spread the liquid uniformly, resist plugging and fouling, provide free space for gas flow, and allow operating flexibility.[37] Large towers frequently have a liquid redistributor to collect liquid off the column wall and direct it toward the center of the column for redistribution and enhanced contact in the lower section of packing.[7] Liquid redistributors are generally required for every 8 to 20 feet of random packing depth.[6, 38]

Distributors fall into two categories: gravitational types, such as orifice and weir types, and pressure-drop types, such as spray nozzles and perforated pipes. Spray nozzles are the most common distributors, but they may produce a fine mist that is easily entrained in the gas flow. They also may plug, and usually require high feed rates to compensate for poor distribution. Orifice-type distributors typically consist of flat trays with a number of risers for vapor flow and perforations in the tray floor for liquid flow. The trays themselves may present a resistance to gas flow.[37] However, better contact is generally achieved when orifice distributors are used.[5]

Packing materials provide a large, wetted surface for the gas stream maximizing the area available for mass transfer. Packing materials are available in a variety of forms, each having specific characteristics with respect to surface area, pressure drop, weight, corrosion resistance, and cost. Packing life varies depending on the application. In ideal circumstances, packing will last as long as the tower itself. In adverse environments packing life may be as short as 1 to 5 years due to corrosion, fouling, and breakage.[39]

Packing materials are categorized as random or structured. Random packings are usually dumped into an absorption column and allowed to settle. Modern random packings consist of engineered shapes intended to maximize surface-to-volume ratio and minimize pressure drop.[2] Examples of different random packings are presented in Figure 1.7. The first random packings specifically designed for absorption towers were made of ceramic. The use of ceramic has declined because of their brittleness, and the current markets are dominated by metal and plastic. Metal packings cannot be used for highly corrosive pollutants, such as acid gas, and plastic packings are not suitable for high temperature applications. Both plastic and metal packings are generally limited to an unsupported depth of 20 to 25. At higher depths the weight may deform the packing.[38]

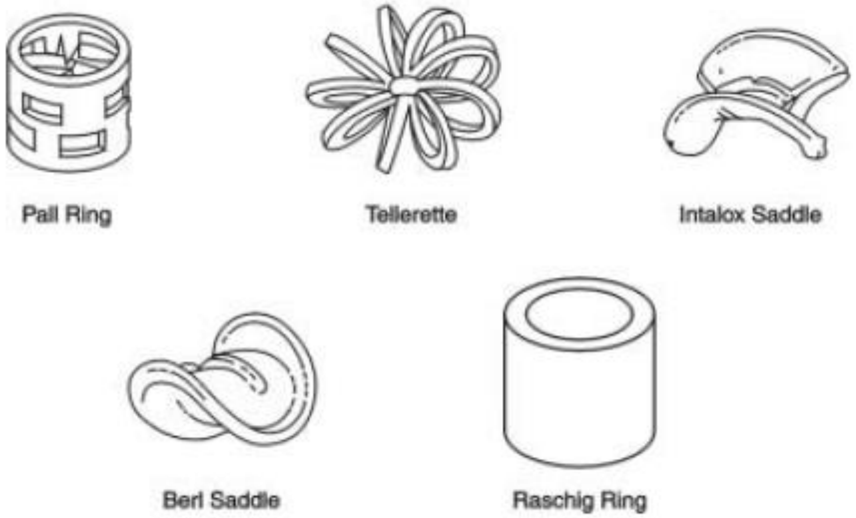

Figure 1.7: Random Packing Material

Structured packing may be random packings connected in an orderly arrangement, interlocking grids, or knitted or woven wire screen shaped into cylinders or gauze like arrangements. They usually have smaller pressure drops and can handle greater solvent flow rates than random packings.[7] However, structured packings are more costly to install and may not be practical for smaller columns. Most structured packings are made from metal or plastic.

To ensure the waste gas is well distributed, an open space between the bottom of the tower and the packing is necessary. Support plates hold the packing above the open space. The support plates must have enough strength to carry the weight of the packing, and enough free area to allow solvent and gas to flow with minimum restrictions.[7]

High gas velocities can fluidize packing on top of a bed. The packing could then be carried into the distributor, become unleveled, or be damaged.[37] A packing restrainer may be installed at the top of the packed bed to contain the packing. The packing restrainer may be secured to the wall so that column upsets will not dislocate it, or a “floating” unattached weighted plate may be placed on top of the packing so that it can settle with the bed. The latter is often used for fragile ceramic packing.

1.3.1.2 Packed Tower Operation

As discussed in Section 1.3.1.1, the most common packed tower designs are countercurrent. As the waste gas flows up the packed column it will experience a drop in its pressure as it meets resistance from the packing materials and the solvent flowing down. Pressure drop in a column is a function of the gas and liquid flow rates and properties of the packing elements, such as surface area and free volume in the tower. A high pressure drop results in high fan power to drive the gas through the packed tower and consequently higher costs. The pressure drop in a packed tower generally ranges from 0.5 to 1.0 in. H_2O/ft of packing.[8]

For each column, there are upper and lower limits to solvent and vapor flow rates that ensure satisfactory performance. The gas flow rate may become so high that the drag on the solvent is sufficient to keep the solvent from flowing freely down the column. Solvent begins to accumulate and blocks gas flow in the entire cross-section, which increases the pressure drop and prevents the packing from mixing the gas and solvent effectively.

When all the free volume in the packing is filled with liquid and the liquid is carried back up the column, the absorber is considered to be flooded.[7] Most packed towers operate at 60 to 70% of the gas flooding velocity, as it is not practical to operate a tower in a flooded condition.[8] A minimum liquid flow rate is also required to wet the packing material sufficiently for effective mass transfer to occur between the gas and liquid.[8]

The waste gas inlet temperature is another important parameter. In general, the higher the gas temperature, the lower the absorption rate, and vice-versa. Excessively high gas temperatures also can lead to significant solvent loss through evaporation. Consequently, precoolers (e.g., spray chambers) may be needed to reduce the air temperature to acceptable levels.[4]

For operations that are based on chemical reaction with absorption, an additional concern is the rate of reaction between the solvent and pollutant(s). Most gas absorption chemical reactions are relatively fast and the rate limiting step is the physical absorption of the pollutants into the solvent. However, for solvent-pollutant systems where the chemical reaction is the limiting step, the rates of reaction would need to be analyzed kinetically.

Heat may be generated as a result of exothermal chemical reactions. Heat may also be generated when large amounts of solute are absorbed into the liquid phase, due to the heat of solution. The resulting change in temperature along the height of the absorber column may damage equipment and reduce absorption efficiency. This problem can be avoided by adding cooling coils to the column.[8] However, in those systems where water is the solvent, adiabatic saturation of the gas occurs during absorption due to solvent evaporation. This causes a substantial cooling of the absorber that offsets the heat generated by chemical reactions. Thus, cooling coils are rarely required with those systems.[6] In any event, packed towers may be designed assuming that isothermal conditions exist throughout the column.[8]

The effluent from the column may be recycled into the system and used again. This is usually the case if the solvent is costly, e.g., hydrocarbon oils or caustic solution. Initially, the recycle stream may go to a waste treatment system to remove the pollutants or the reaction product. Make-up solvent may then be added before the liquid stream re-enters the column. Recirculation of the solvent requires a pump, solvent recovery system, solvent holding and mixing tanks, and any associated piping and instrumentation.

1.3.2 Design Procedures for Packed Tower Absorbers

The design of packed tower absorbers for controlling gas streams containing a mixture of pollutants and air depends on knowledge of the following parameters:

- Waste gas flow rate
- Waste gas composition and concentration of the pollutants in the gas stream
- Required removal efficiency
- Equilibrium relationship between the pollutants and solvent
- Properties of the pollutant(s), waste gas, and solvent: diffusivity, viscosity, density, and molecular weight.

The primary objectives of the design procedures are to determine column surface area and pressure drop through the column. The following steps are used to determine the surface area and pressure drop:

- Determine the gas and liquid stream conditions entering and exiting the column.
- Determine the absorption factor (AF).
- Determine the diameter of the column (D).
- Determine the tower height (H_{tower}) and surface area (S).
- Determine the packed column pressure drop (P).

To simplify the sizing procedures, a number of assumptions have been made. For example, the waste gas is assumed to comprise a two-component waste gas mixture (pollutant/air), where the pollutant consists of a single compound present in dilute quantities. The waste gas is assumed to behave as an ideal gas and the solvent is assumed to behave as an ideal solution. Heat effects associated with absorption are considered to be minimal for the pollutant concentrations encountered. The procedures also assume that, in chemical absorption, the process is not reaction rate limited, i.e., the reaction of the pollutant with the solvent is considered fast compared to the rate of absorption of the pollutant into the solvent.

The design procedures presented here are complicated, and careful attention to units is required. A list of all design variables referred to in in the following sections and their appropriate measurement units is provided in Appendix A, Table A-3.

1.3.2.1 Determining Gas and Liquid Stream Conditions

Gas absorbers are designed based on the ratio of liquid to gas entering the column (L_i/G_i), slope of the equilibrium curve (m), and the desired removal efficiency (η). These factors are calculated from the inlet and outlet gas and liquid stream variables:

- Waste gas flow rate, in actual cubic feet per minute (acfm), entering and exiting column (G_i and G_o, respectively)
- Pollutant concentration (lb-moles pollutant per lb-mole of pollutant free gas) entering and exiting the column in the waste gas (Y_i and Y_o, respectively)
- Solvent flow rate, in gallons per minute (gpm), entering and exiting the column (L_i and L_o, respectively)
- Pollutant concentration (lb-moles pollutant per lb-mole of pollutant free solvent) entering and exiting the column in the solvent (X_i and X_o, respectively)

For this design approach, we assume the inlet gas stream variables are known, and that a specific pollutant removal efficiency has been chosen as the design basis (i.e., the variables G_i, Y_i, and η are known). For the dilute concentrations typically encountered in pollution control applications and negligible changes in moisture content, G_i is assumed equal to G_o. If a once-through process is used, or if the spent solvent is regenerated by an air stripping process before it is recycled, the value of X_i will approach zero. The remaining stream variables (Y_o, L_i L_o, and X_o), the following procedures must be followed. A schematic diagram of a packed tower with inlet and outlet flow and concentration variables labeled is presented in Figure 1.7.

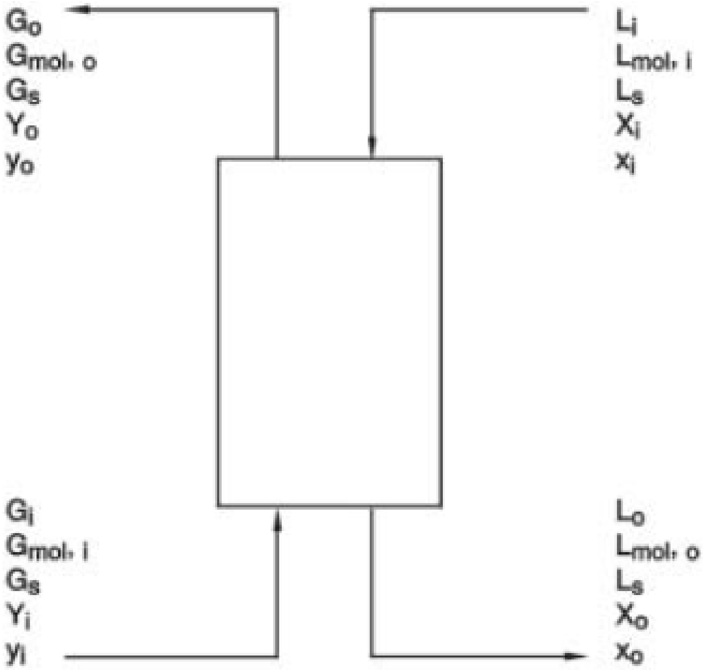

Figure 1.8: Schematic Diagram of Countercurrent Packed Bed Operation

The exit pollutant concentration may be calculated using the following equation:

$$Y_o = Y_i\left(1 - \frac{\eta}{100}\right) \tag{1.57}$$

The liquid flow rate entering the absorber, L_i (gpm), is then calculated using a graphical method. Figure 1.8 presents an example of an equilibrium curve and operating line. The equilibrium curve indicates the relationship between the concentration of pollutant in the waste gas and the concentration of pollutant in the solvent at a specified temperature. The operating line indicates the relationship between the concentration of the pollutant in the gas and solvent at any location in the gas absorber column. The vertical distance between the operating line and equilibrium curve indicates the driving force for diffusion of the pollutant between the gas and liquid phases. The minimum amount of liquid which can be used to absorb the pollutant in the gas stream corresponds to an operating line drawn from the outlet concentration in the gas stream (Y_o) and the inlet concentration in the solvent stream (X_i) to the point on the equilibrium curve corresponding to the entering pollutant concentration in the gas stream (Y_i). At the intersection point on the equilibrium curve, the diffusional driving forces are zero, the required time of contact for the concentration change is infinite, and an infinitely tall tower results.

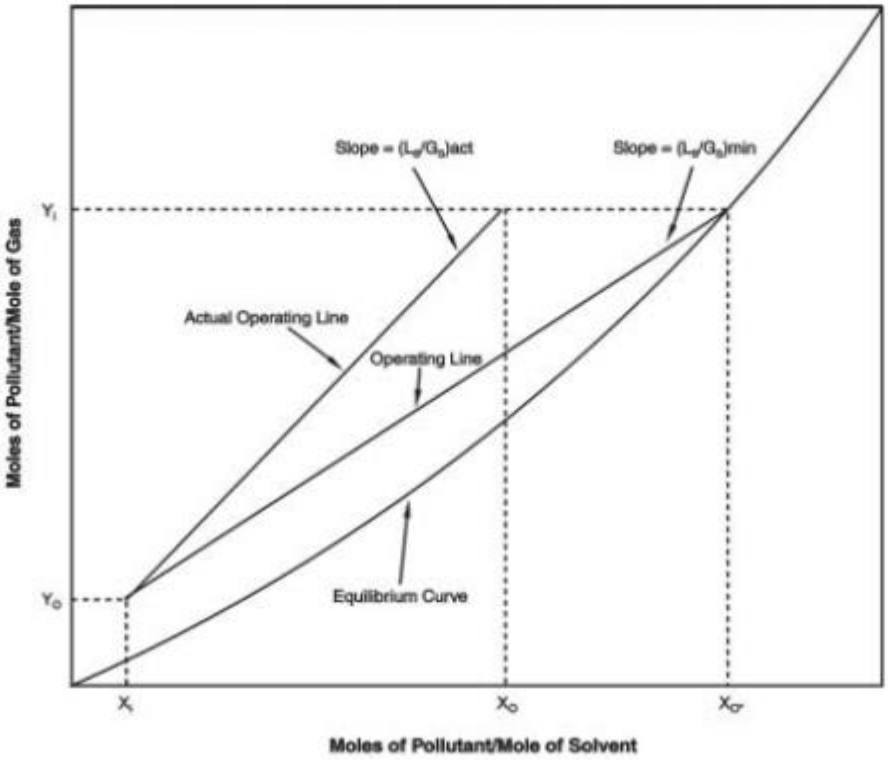

Figure 1.9: Minimum and Actual Liquid-to-Gas Ratio.

The slope of the operating line intersecting the equilibrium curve is equal to the minimum L/G ratio on a moles of pollutant-free solvent (L_s) per moles of pollutant-free gas basis G_s. In other words, the values L_s and G_s do not include the moles of pollutant in the liquid and gas streams. The values of L_s and G_s are constant through the column if a negligible amount of moisture is transferred from the liquid to the gas phase. The slope may be calculated from the following equation:

$$\left(\frac{L_S}{G_S}\right)_{min} = \frac{Y_i - Y_o}{X_o^* - X_i} \tag{1.58}$$

where X^*_o would be the maximum concentration of the pollutant in the liquid phase if it were allowed to come to equilibrium with the pollutant entering the column in the gas phase, Y_i. The value of X^*_o is taken from the equilibrium curve. Because the minimum L_s/G_s, ratio is an unrealistic value, it must be multiplied by an adjustment factor, commonly between 1.2 and 1.5, to calculate the actual L/G ratio:[8]

$$\left(\frac{L_S}{G_S}\right)_{act} = \left(\frac{L_S}{G_S}\right)_{min} \times (adjustment\ factor) \tag{1.59}$$

The variable G_s may be calculated using the equation:

$$G_s = \frac{60\rho_G G_i}{MW_G(1+Y_i)} \tag{1.60}$$

where 60 is the conversion factor from minutes to hours, MW_G, is the molecular weight of the gas stream (lb/lb-mole), and ρ_G is the density of the gas stream (lb/ft^3). For pollutant concentrations typically encountered, the molecular weight and density of the waste gas stream are assumed to be equal to that of ambient air.

The variable L_s may then be calculated by:

$$L_s = \left(\frac{L_s}{G_s}\right)_{act} \times G_s \tag{1.61}$$

The total molar flow rates of the gas and liquid entering the absorber ($G_{mol,i}$ and $L_{mol,i}$) are calculated using the following equations:

$$G_{mol,i} = G_s(1 + Y_i) \tag{1.62}$$

$$L_{mol,i} = L_s(1 + X_i) \tag{1.63}$$

The volume flow rate of the solvent, L_i, may then be calculated by using the following relationship:

$$L_i = \frac{7.48 L_{mol,i} MW_L}{60 r_L} \qquad (1.63)$$

where 60 is the conversion factor from minutes to hours, MW_L, is the molecular weight of the liquid stream (lb/lb-mole), ρ_L is the density of the liquid stream (lb/ft^3), and 7.48 is the factor used to convert cubic feet to gallons. If the volume change in the liquid stream entering and exiting the absorber is assumed to be negligible, then $L_i = L_o$.

Gas absorber vendors have provided a range for the L_i/G_i ratio for acid gas control from 2 to 20 gpm of solvent per 1000 cfm of waste gas.[40] Even for pollutants that are highly soluble in a solvent (e.g., HCl in water), the adjusted L_i/G_i ratio calculated using Equations 1.59 to 1.64 would be much lower than this range, because these equations do not consider the flow rate of the solvent required to wet the packing.

Finally, the actual operating line may be represented by a material balance equation over the gas absorber:[7]

$$X_i L_s + Y_i G_s = X_o L_s + Y_o G_s \tag{1.65}$$

Equation 1.65 may then be solved for X_o:

$$X_o = \frac{Y_i - Y_o}{\left(\frac{L_s}{G_s}\right)} + X_i \tag{1.66}$$

1.3.2.2 Determining Absorption Factor

The absorption factor (AF) value is frequently used to describe the relationship between the equilibrium line and the liquid-to-gas ratio. For many pollutant-solvent systems, the most economical value for AF ranges around 1.5 to 2.0.[25] The following equation may be used to calculate AF: [7, 8]

$$AF = \frac{L_{mol,i}}{mG_{mol,i}} \tag{1.67}$$

where *m* is the slope of the equilibrium line on a mole fraction basis. The value of *m* may be obtained from available literature on vapor/liquid equilibrium data for specific systems. Since the equilibrium curve is typically linear in the concentration ranges usually encountered in air pollution control, the slope, *m* would be constant (or nearly so) for all applicable inlet and outlet liquid and gas streams. The slope may be calculated from mole fraction values using the following equation:[7]

$$m = \frac{y_o^* - y_i^*}{x_o - x_i} \tag{1.68}$$

where y^*_i and y^*_o are the mole fractions of the pollutant in the vapor phase in equilibrium with the mole fractions of the pollutant entering and exiting the absorber in the liquid, x_i and x_o, respectively. The slope of the equilibrium line in Figure 1.8 is expressed in terms of concentration values X_i, X_o, Y^*_i, and Y^*_o. These values may be converted to x_i, x_o, y^*_i, and y^*_o using the equations:

$$x_i = \frac{X_i}{1+X_i} \tag{1.69}$$

$$x_o = \frac{X_o}{1+X_o} \tag{1.70}$$

$$y_i^* = \frac{Y_i^*}{1+Y_i^*} \tag{1.71}$$

$$y_o^* = \frac{Y_o^*}{1+Y_o^*} \tag{1.72}$$

where the units for each of these variables are listed in Appendix A, Table A-3.

The absorption factor (AF) is used to calculate the theoretical number of transfer units and the theoretical height of a transfer unit. First, however, the column diameter must be determined.

1.3.2.3 Determining Column Diameter

Once stream conditions have been determined, the diameter of the column may be estimated. The design presented in this section is based on selecting a fraction of the gas flow rate at flooding conditions. Alternatively, the column may be designed for a specific pressure drop (see Section 1.3.2.6). Eckert's modification to the generalized correlation for randomly packed towers based on flooding considerations is used to obtain the superficial gas flow rate entering the absorber, $G_{sfr,i}$ (lb/sec-ft^2), or the gas flow rate per cross-sectional area based on the $L_{mol,i}/G_{mol,i}$ ratio calculated in Section 1.3.2.2.[38] The cross-sectional area (A) of the column and the column diameter (D) can then be determined from $G_{sfr,i}$. Figure 1.9 presents the relationship between $G_{sfr,i}$ and the $L_{mol,i}/G_{mol,i}$ ratio at the tower flood point. The *Abscissa* value (X axis) in the graph is expressed as:[38]

$$Abscissa = \left(\frac{L_{mol,i}}{G_{mol,i}}\right)\left(\frac{MW_L}{MW_G}\right)\sqrt{\frac{\rho_G}{\rho_L}} \tag{1.73}$$

The Ordinate value (Y axis) in the graph is expressed as:[38]

$$Ordinate = \frac{\left(G_{sfr,i}\right)^2 \Psi F_p \left(\frac{\mu_L}{2.42}\right)^{0.2}}{\rho_L \rho_G g_c} \tag{1.74}$$

where F_p is a packing factor, g_c is the gravitational constant (32.2), μ_L is the viscosity of the solvent (lb/ft-hr), 2.42 is the factor used to convert lb/ft-hr to centipoise, and Ψ is the ratio of the density of the scrubbing liquid to water. The value of F_p may be obtained from packing vendors. Typical values for commonly used packing materials are shown in Appendix B, Table B-1.

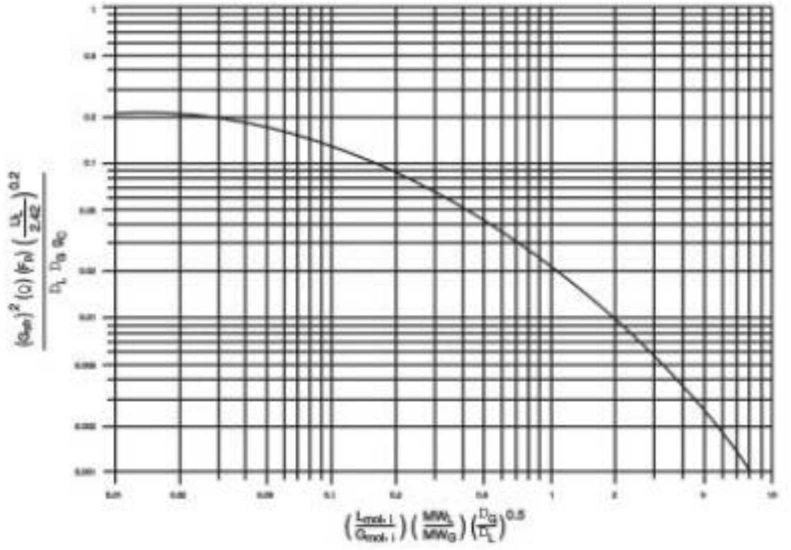

Figure 1.10: Eckert's Modification to the Generalized Correlation at Flooding Rate.

After calculating the *Abscissa* value, a corresponding *Ordinate* value may be determined from the flooding curve. The *Ordinate* may also be calculated using the following equation:[38]

$$Ordinate = 10^{\left[-1.668-1.085(\log Abscissa)-0.297(\log Abscissa)^2\right]} \tag{1.75}$$

Equation 1.75 may then be rearranged to solve for $G_{sfr,i}$:

$$G_{sfr,i} = \sqrt{\frac{\rho_l \ \rho_G \ g_c \ (Ordinate)}{F_p \ \Psi \left(\frac{\mu_L}{2.42}\right)^{0.2}}} \tag{1.76}$$

The cross-sectional area of the tower (ft^2) is calculated as:

$$A = \frac{G_{mol,i} MW_G}{3{,}600 G_{sfr,i} f} \tag{1.77}$$

where *f* is the flooding factor and 3,600 is the conversion factor from hours to seconds. To prevent flooding, the column is operated at a fraction of $G_{sfr,i}$. The value of *f* typically ranges from 0.60 to 0.75.[4]

The diameter of the column (ft) can be calculated from the cross-sectional area, by:

$$D = \sqrt{\frac{4}{\pi} A} \tag{1.78}$$

If a substantial change occurs between inlet and outlet volumes (i.e., moisture is transferred from the liquid phase to the gas phase), the diameter of the column will need to be calculated at the top and bottom of the column. The larger of the two values is then chosen as a conservative number. As a rule of thumb, the diameter of the column should be at least 15 times the size of the packing used in the column. If this is not the case, the column diameter should be recalculated using a smaller diameter packing.[38]

The superficial liquid flow rate entering the absorber, $L_{sfr,i}$ (lb/hr-ft) sectional area determined in Equation 1.64 is calculated from the equation:

$$L_{sfr,i} = \frac{L_{mol,i} MW_L}{A} \tag{1.79}$$

For the absorber to operate properly, the liquid flow rate entering the column must be high enough to effectively wet the packing so mass transfer between the gas and liquid can occur. The minimum value of $L_{sfr,i}$ that is required to wet the packing effectively can be calculated using the equation:[8, 41]

$$\left(L_{sfr,i}\right)_{min} = MWR \rho_L a \tag{1.80}$$

where MWR is defined as the minimum wetting rate (ft^2/hr), and a is the surface area to volume ratio of the packing (ft^2/ft^3). An MWR value of 0.85 ft^2/hr is recommended for ring packings larger than 3 inches and for structured grid packings. For other packings, an MWR of 1.3 ft^2/hr is recommended. [8, 41] Appendix B, Table B-1 contains values of *a* for common packing materials.

If $L_{sfr,i}$ (the value calculated in Equation 1.79) is smaller than $(L_{sfr,})_{min}$ (the value calculated in Equation 1.80), there is insufficient liquid flow to wet the packing using the current design parameters. The value of $G_{sfr,i}$, and A must be recalculated. See Appendix D for details.

1.3.2.4 Determining Tower Height and Surface Area

Tower height is primarily a function of packing depth. The required depth of packing (H_{pack}) is determined from the theoretical number of overall transfer units (N_{tu}) needed to achieve a specific removal efficiency, and the height of the overall transfer unit (H_{tu}):[7]

$$H_{pack} = N_{tu}H_{tu} \tag{1.81}$$

The number of overall transfer units may be estimated graphically by stepping off stages on the equilibrium-operating line graph from inlet conditions to outlet conditions, or by the following equation:[7]

$$N_{tu} = \frac{ln\left[\left(\frac{y_i - mx_i}{y_o - mx_i}\right)\left(1 - \frac{1}{AF}\right) + \frac{1}{AF}\right]}{1 - \frac{1}{AF}} \tag{1.82}$$

Where *ln* is the natural logarithm of the quantity indicated.

The equation is based on several assumptions:

1) Henry's law applies for a dilute gas mixture;
2) the equilibrium curve is linear from x_i to x_o; and
3) the pollutant concentration in the solvent is dilute enough such that the operating line can be considered a straight line.[7]

If $x_i \approx 0$ (i.e., a negligible amount of pollutant enters the absorber in the liquid stream) and $1/A \approx 0$ (i.e., the slope of the equilibrium line is very small and/or the L_{mol}/G_{mol} ratio is very large), Equation 1.82 simplifies to:

$$N_{tu} = ln\left(\frac{y_i}{y_o}\right) \tag{1.83}$$

There are several methods that may be used to calculate the height of the packed tower, all based on empirically determined packing constants. One commonly used method involves determining the overall gas and liquid mass transfer coefficients (K_G, K_L). A major difficulty in using this approach is that values for K_G and K_L are frequently unavailable for the specific pollutant-solvent systems of interest. The reader is referred to the book *Random Packing and Packed Tower Design Applications* in the reference section for further details regarding this method.[42]

For this chapter, the method used to calculate the height of the tower is based on estimating the height of the gas and liquid film transfer units, H_L and H_G, respectively:[7]

$$H_{tu} = H_G + \frac{1}{AF}H_L \tag{1.84}$$

The following correlations may be used to estimate values for H_L and H_G:[40]

$$H_G = \left[\alpha \frac{\left(3{,}600 f G_{sfr,i}\right)^{\beta}}{\left(L_{sfr,i}\right)^{\Gamma}}\right]\sqrt{\frac{\mu_G}{\rho_G D_G}} \tag{1.85}$$

$$H_L = \phi\left(\frac{L_{sfr,i}}{\mu_L}\right)^{b}\sqrt{\frac{\mu_L}{\rho_L D_L}} \tag{1.86}$$

The quantity $\mu/\rho\ D$ is the Schmidt number and the variables $ß$, b and Γ are packing constants specific to each packing type. Typical values for these constants are listed in Appendix B, Tables B-2 and B-3. The advantage to using this estimation method is that the packing constants may be applied to any pollutant-solvent system. One packing vendor offers the following modifications to Equations 1.84 and 1.85 for their specific packing:[43]

$$H_G = \left[\alpha \frac{(3{,}600 f G_{sfr,i})^{\beta}}{(L_{sfr,i})^{\Gamma}}\right] \sqrt{\frac{\mu_G}{\rho_G D_G}} \left(\frac{\mu_L^{\Gamma}}{\mu_G^{\beta}}\right) \tag{1.87}$$

$$H_L = \phi \left(\frac{L_{sfr,i}}{\mu_L}\right)^{b} \sqrt{\frac{\mu_L}{\rho_L D_L}} \left(\frac{T}{286}\right)^{-4.255} \tag{1.88}$$

Where T is the temperature of the solvent in Kelvin.

After solving for H_{pack} using Equation 1.81, the total height of the tower may be calculated from the following correlation:[44]

$$H_{tower} = 1.40 H_{pack} + 1.02 D + 2.81 \tag{1.89}$$

Equation 1.89 was developed from information reported by gas absorber vendors and is applicable for tower diameters from 2 to 12 feet and packing depths from 4 to 12 feet. The surface area (S) of the gas absorber can be calculated using the equation:[44]

$$S = \pi D \left(H_{tower} + \frac{D}{2}\right) \tag{1.90}$$

Equation 1.90 assumes the ends of the absorber are flat and circular.

1.3.2.5 Calculating Column Pressure Drop

Pressure drop in a gas absorber is a function of $G_{sfr,i}$ and properties of the packing used. The pressure drop in packed columns generally ranges from 0.5 to 1 inch of H_2O per foot of packing. The absorber may be designed for a specific pressure drop or pressure drop may be estimated using Leva's correlation: [8, 38]

$$\Delta P = c\ 10^{\left(\frac{j\ L_{sfr,i}}{3{,}600}\right)} \frac{(f G_{sfr,i})^2}{\rho_G} \tag{1.91}$$

The packing constants c and j are found in Appendix B, Table B-4, and 3,600 is the conversion factor from seconds to hours. The equation was originally developed for air-water systems. For other liquids, $L_{sfr,i}$ is multiplied by the ratio of the density of water to the density of the liquid.

1.3.2.6 Alternative Design Procedure

The diameter of a column can be designed for a specific pressure drop, rather than being determined based on a fraction of the flooding rate. Figure 1.10 presents a set of generalized correlations at various pressure drop design values. The *Abscissa* value of the graph is similar to Equation 1.73:[38]

$$Abscissa = \left(\frac{L_{mol,i}}{G_{mol,i}}\right)\left(\frac{MW_L}{MW_G}\right)\sqrt{\frac{\rho_G}{P_L-\rho_G}} \tag{1.92}$$

The *Ordinate* value is expressed as:[11]

$$Ordinate = \frac{(G_{sfr,i})^2 F_P\left(\frac{\mu_L}{2.42}\right)^{0.1}}{(\rho_L-\rho_G)\rho_G g_c} \tag{1.93}$$

For a calculated *Abscissa* value, a corresponding *Ordinate* value at each pressure drop can be read off Figure 1.10 or can be calculated from the following equation:[38]

$$\begin{gathered} Ordinate = exp[\,k_0 + k_1(1nAbscissa) + k_2(1nAbscissa)^2 \\ +k_3(1nAbscissa)^3 + k_4(1nAbscissa)^4] \end{gathered} \tag{1.94}$$

The constants k_0, k_l, k_2, k_3, and k_4, are shown in Table 1.6 for each pressure drop value.

Table 1.6: Values of Constants k_0 through k_4 for Various Pressure Drops

Δ P (inches water/ft packing)	**k_0**	**k_1**	**k_2**	**k_3**	**k_4**
0.05	-6.3205	-0.6080	-0.1193	-0.0068	0.0003
0.10	-5.5009	-0.7851	-0.1350	0.0013	0.0017
0.25	-5.0032	-0.9530	-0.1393	0.0126	0.0033
0.50	-4.3992	-0.9940	-0.1698	0.0087	0.0034
1.00	-4.0950	-1.0012	-0.1587	0.0080	0.0032
1.50	-4.0256	-0.9895	-0.0830	0.0324	0.0053

Equation 1.93 can be solved for $G_{sfr,i}$.

$$G_{sfr,i} = \sqrt{\frac{(\rho_L-\rho_G)\rho_G g_c(Ordinate)}{F_P\left(\frac{\mu_L}{2.42}\right)^{0.1}}} \tag{1.95}$$

The remaining calculations to estimate the tower diameter and $L_{sfr,i}$ are the same as presented in Section 1.3.2.3, except the flooding factor (*f*) is not used in the equations. The flooding factor is not required because an allowable pressure drop that will not cause flooding is chosen to calculate the diameter rather than designing the diameter at flooding conditions and then taking a fraction of that value. Such an approach allows for a column diameter that will permit typical operations with minimized potential for flooding.

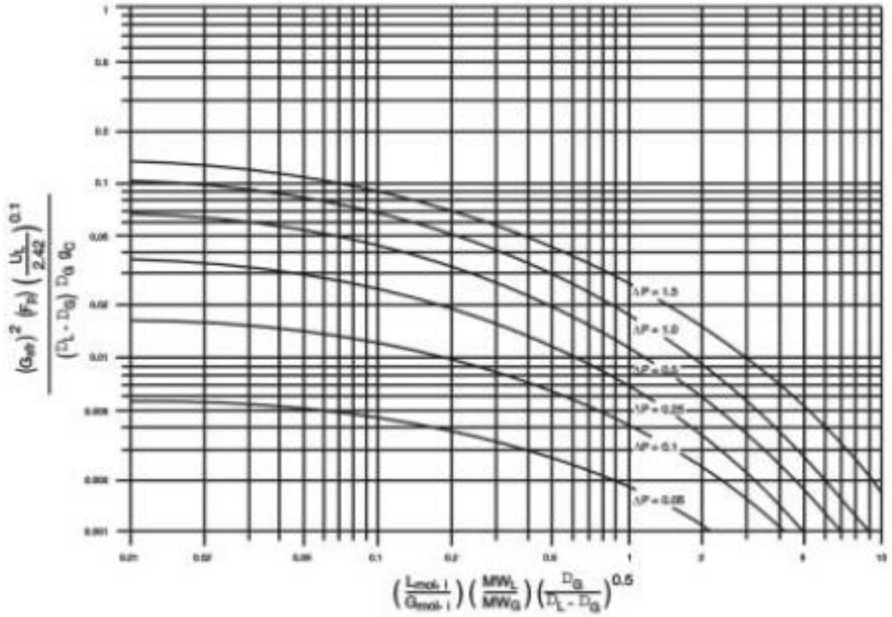

Figure 1.11: Generalized Pressure Drop Correlations [38]

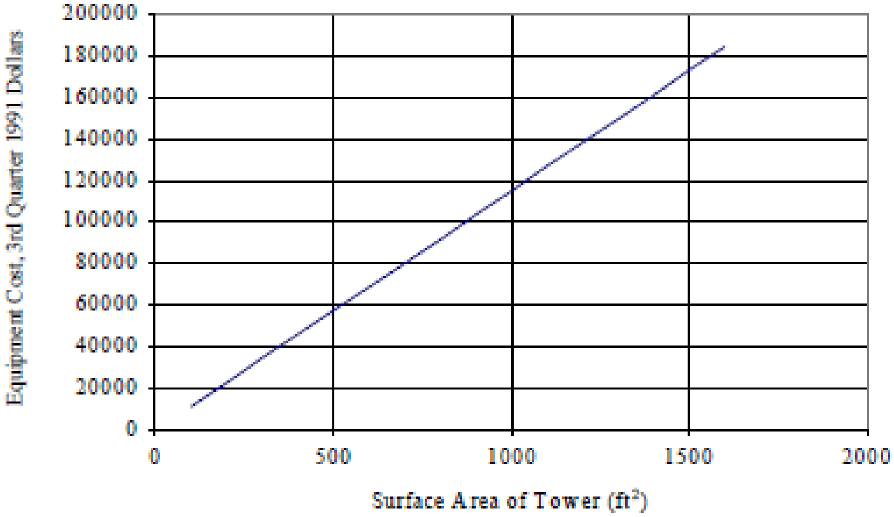

Figure 1.12: Packed Tower Equipment Cost [44]

1.3.3 Estimating Total Capital Investment

This section presents the procedures and data necessary for estimating capital costs for vertical packed bed gas absorbers using countercurrent flow to remove gaseous pollutants from waste gas streams. Equipment costs for packed bed absorbers are presented in Section 1.3.3.1, with installation costs presented in Section 1.3.3.2.

Total capital investment, *TCI*, includes equipment cost, *EC*, for the entire gas absorber unit, taxes, freight charges, instrumentation, and direct and indirect installation costs. All costs are presented in third quarter 1991 dollars. For escalating these and the other scrubber costs to more current year dollars, the EPA suggests that the Chemical Engineering Plant Cost Index (CEPCI) could be useful. This cost index covers construction and equipment costs for the chemical process industries, a likely sector for installation of these control devices. The CEPCI is available at http://www.chemengonline.com/pci.[14]

As for the rest of this chapter and the Control Cost Manual, the costs presented are study estimates with an expected accuracy of ± 30 percent. It must be kept in mind that even for a given application, design and manufacturing procedures vary from vendor to vendor, so costs vary. All of these costs are for new plant installations; no retrofit cost considerations are included.

1.3.3.1 Equipment Costs for Packed Tower Absorbers

Gas absorber vendors were asked to supply cost estimates for a range of tower dimensions (i.e., height, diameter) to account for the varying needs of different applications. The equipment for which they were asked to provide costs consisted of a packed tower absorber made of fiberglass reinforced plastic (FRP), and to include the following equipment components:

- absorption column shell;
- gas inlet and outlet ports;
- liquid inlet port and outlet port/drain;
- liquid distributor and redistributor;
- two packing support plates;
- mist eliminator;
- internal piping;
- sump space; and
- platforms and ladders.

The cost data the vendors supplied were first adjusted to put them on a common basis, and then were regressed against the absorber surface area (S). The equation shown below is a linear regression of cost data provided by six vendors in 1991 dollars. [40, 44]

$$TotalTowerCost(\$) = 115 \times S \tag{1.96}$$

where S is the surface area of the absorber, in ft^2. Figure 1.11 depicts a plot of Equation 1.96. This equation is applicable for towers with surface areas from 69 to 1,507 ft^2 constructed of FRP.

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[14] Please note that mention of the Chemical Engineering Plant Cost Index (CEPCI) in this Manual is not meant to serve as endorsement for commercial purposes. In addition, escalation of costs beyond 5 years is typically not recommended according to Section 1, Chapter 2 (Cost Estimation Methodology) of this Control Cost Manual. We escalated these costs from 1991 to 2016 costs given insufficient control cost data in more recent year dollars.

Costs for towers made of materials other than FRP may be estimated using the following equation:

$$TTC_M = CF \times TTC \tag{1.97}$$

where TTC_M is the total cost of the tower using other materials, and TTC is the total tower cost as estimated using Equation 1.96. The variable CF is a cost factor to convert the cost of an FRP gas absorber to an absorber fabricated from another material. Ranges of cost factors provided by vendors are listed for the following materials of construction:[40]

Material	CF
304 Stainless Steel	1.10 -1.75
Polypropylene:	0.80–- 1.10
Polyvinylchloride:	0.50–- 0.90

Auxiliary costs encompass the cost of all necessary equipment not included in the absorption column unit. Auxiliary equipment includes packing material, instruments and controls, pumps, and fans. Cost ranges for various types of packing materials are presented in Table 1.7. The cost of structured packings varies over a much wider range. In 1991, structured packings made of stainless steel ranged from \$45/ft^3 to \$405/ft^3 (escalating these costs to 2016 dollars using the CEPCI results in costs of \$62/ft^3 to \$562/ft^3), and those made of polypropylene ranged from \$65/ft^3 to \$350/ft^3 (\$90/ft^3 to \$485/ft^3 in 2016 dollars).[45]

Table 1.7: Typical Costs for Random Packing Materials[a]

Nominal Diameter (inches)	Construction Material	Packing Type	Packing cost (\$/ft^3)	
			<100 ft^3	>100 ft^3
1	304 stainless steel	Pall rings, Raschig rings, Ballast rings	70-109	65-99
1	Ceramic	Raschig rings, Berl saddles	33-44	26-36
1	Polypropylene	Tri-Pak®, Pall rings, Ballast rings, Flexisaddles	14-37	12-34
2	Ceramic	Berl saddles, Raschig rings	13-32	10-30
2	Polypropylene	Tri-Pac®, Lanpac®, Flexiring, Flexisaddle Tellerette®	3-20	5-19
3.5	304 stainless steel	Ballast rings	30	27
3.5	Polypropylene	Tri-pack®, Lanpac®, Ballast rings	6-14	6-12

[a] Provided by packing vendors based on 1991 dollars. [45]

® Denotes registered trademark.

Similarly, the cost of instruments and controls varies widely depending on the complexity required. Estimates provided to the EPA by vendors in 1991 ranged from \$1,000 to \$10,000 per column (\$1,390 to \$13,900 per column in 2016 dollars). A factor of 10% of the total equipment cost (EC) is used to estimate the cost of instruments and controls. Design and cost correlations for fans and pumps are presented in Section 2 (Generic Equipment and Devices) of this Manual.

However, cost data for auxiliary equipment are also available from the literature (e.g., see reference 45).

The EC is the sum of the component equipment costs, which includes costs for the tower, packing material and the auxiliary equipment.

$$EC = TTC + PackingCost + AuxiliaryEquipment \tag{1.98}$$

The purchased equipment cost (PEC) includes the cost of the absorber with packing and its auxiliaries (*EC*), instrumentation (0.10 *EC*), sales tax (0.03 *EC*), and freight (0.05 *EC*): [40, 47],

$$PEC = (1 + 0.10 + 0.03 + 0.05)EC = 1.18EC \tag{1.99}$$

1.3.3.2 Installation Costs

As discussed in Section 1, Chapter 2 of this Manual, the TCI is estimated from the total purchased equipment cost via direct and indirect installation cost factors and a contingency factor. A breakdown of the factors for wet packed tower absorbers is shown in Table 1.8. [47] The installation factors presented in Table 1.8 were confirmed by the absorber vendors. [40]

As indicated in Section 1.2, Chapter 2 of this Manual, the TCI also includes costs for land, working capital, and off-site facilities, which are not included in the direct/indirect installation factor. However, as these items are rarely required with wet scrubbers, they will not be included in the TCI here. Further, no factors have been provided for site preparation (SP) and buildings (Bldg.), as these site-specific costs depend very little on the purchased equipment cost.

Thus, the total capital investment, *TCI*, is obtained by multiplying the purchased equipment cost, *PEC*, by the total installation factor and adding an amount for contingencies:

$$TCI = 2.17 \times PEC + C \tag{1.100}$$

The contingency, C, accounts for unexpected costs associated with the fabrication and installation of the absorber and is calculated by multiplying the total direct and indirect costs by a contingency factor (CF). A default value of 10% of the direct and indirect costs is typically used for CF. However, values of between 5% and 15% may be used. More information can be found on contingency in Section 1, Chapter 2 of this Manual.

Table 1.8: Capital Cost Factors for Wet Packed Tower Absorbers [47]

Cost Item	Factor
Direct Costs	
Purchased equipment costs	
Absorber + packing + auxiliary equipment[a], EC	As estimated, A
Instrumentation[b]	0.10 A
Sales taxes	0.03 A
Freight	0.05 A
Purchased equipment cost, PEC	B = 1.18 A
Direct installation costs	
Foundations & supports	0.12 B
Handling & erection	0.40 B
Electrical	0.01 B
Piping	0.30 B
Insulation	0.01 B
Painting	0.01 B
Direct installation costs	0.85 B
Site preparation	As required, SP
Buildings	As required, Bldg.
Total Direct Costs, DC	1.85 B + SP + Bldg.
Indirect Costs (installation)	
Engineering	0.10 B
Construction and field expenses	0.10 B
Contractor fees	0.10 B
Start-up	0.01 B
Performance test	0.01 B
Total Indirect Costs, IC	0.32 B
Contingency Costs, C[c]	CF(DC + IC)
Total Capital Investment = DC + IC + C	2.17 B + SP + Bldg.+ C

[a] Includes the initial quantity of packing, as well as items normally not included with the unit supplied by vendors, such as ductwork, fan, piping, etc.

[b] Instrumentation costs cover pH monitor and liquid level indicator in sump.

[c] The default value for the contingency factor, CF, is 0.10. However, values of between 0.05 and 0.15 may be included to account for unexpected costs associated with the fabrication and installation of the control system. More information can be found on contingency in the cost estimation chapter of this Manual.

1.3.4 Estimating Annual Cost for Wet Packed Tower Absorbers

The total annual cost (TAC) is the sum of the direct and indirect annual costs.

1.3.4.1 Direct Annual Costs

Direct annual costs (DAC) are those expenditures related to operating the equipment, such as labor and materials. The suggested factors for each of these costs are shown in Table 1.9.

Table 1.9: Suggested Annual Cost Factors for Wet Packed Tower Absorbers

Cost Item	Factor
Direct Annual Costs, DAC	
Operating labor[a]	
Operator	0.5 hour per shift
Supervisor	15% of operator labor
Operating materials[b]	Application specific
Solvent	(throughput/yr) x (waste fraction)
Chemicals	Based on annual consumption
Wastewater disposal	(throughput/yr) x (waste fraction)
Maintenance[a]	
Labor	0.5 hour per shift
Material	100% of maintenance labor
Electricity	(consumption rate) x (hours/yr) x (unit cost)
Fan	
Pump	
Indirect Annual Costs, IAC	
Overhead	60% of total labor and material costs
Administrative charges	2% of Total Capital Investment
Property tax	1% of Total Capital Investment
Insurance	1% of Total Capital Investment
Capital recovery[c]	0.0527 x Total Capital Investment
Total Annual Cost	DAC + IAC

[a] These factors were confirmed by vendor contacts.

[b] If system does not use chemicals (e.g., caustic), this quantity is equal to annual solvent consumption.

[c] Assuming a 30-year life at 3.25%. See Chapter 2.

These factors were taken from Section 1 of this Manual and were confirmed by absorber vendors. The annual cost for each item is calculated by multiplying the number of units used annually (i.e., hours, pounds, gallons, kWh) by the associated unit cost.

Operating labor is estimated at 0.5-hour per 8-hour shift. The supervisory labor cost is estimated at 15% of the operating labor cost. Maintenance labor is estimated at 0.5-hour per 8-hour shift. Maintenance materials costs are assumed to equal maintenance labor costs.

Solvent costs are dependent on the total liquid throughput, the type of solvent required, and the fraction of solvent throughput wasted (often referred to as blow-down). Typically, the fraction of solvent wasted varies from 0.1% to 10% of the total solvent throughput.[40] For acid gas absorbers, the amount of solvent wasted is determined by the solids content, with bleed off occurring when solids content reaches 10% to 15% to prevent salt carry-over.[40]

The total annual cost of solvent (Cs) is given by:

$$C_s = L_i WF \left((60 \frac{min}{hr}) \begin{pmatrix} annual \\ operating \\ hours \end{pmatrix} \begin{pmatrix} solvent \\ unit\ cos\ t \end{pmatrix} \right) \tag{1.101}$$

where *WF* is the waste (make-up) fraction, and the solvent unit cost is expressed in terms of \$/gal.

The cost of chemical replacement (C_c) is based on the annual consumption of the chemical and can be calculated by:

$$C_c = \left(\frac{lbschemicalused}{hr} \right) \begin{pmatrix} annual \\ operating \\ hours \end{pmatrix} \begin{pmatrix} chemical \\ unit\ cos\ t \end{pmatrix} \tag{1.102}$$

where the chemical unit cost is in terms of \$/lb.

Solvent disposal (C_{ww}) costs vary depending on geographic location, type of waste disposed of, and availability of on-site treatment. Solvent disposal costs are calculated by:

$$C_{ww} = L_i WF \left(60 \frac{min}{hr} \right) \begin{pmatrix} annual \\ operating \\ hours \end{pmatrix} \begin{pmatrix} solvent \\ disposal\ cos\ t \end{pmatrix} \tag{1.103}$$

where the solvent disposal costs are in terms of \$/gal of waste solvent.

The electricity costs associated with operating a gas absorber derive from fan requirements to overcome the pressure drop in the column, ductwork, and other parts of the control system, and pump requirements to recirculate the solvent. The energy required for the fan can be calculated using Equation 1.104:

$$Energy_{fan} = \frac{1.17 \times 10^{-4} G_i DP}{\varepsilon} \tag{1.104}$$

where Energy (in kilowatts) refers to the energy needed to move a given volumetric flow rate of air (acfm), G_i is the waste gas flow rate entering the absorber, P is the total pressure drop through the system (inches of H_2O) and D is the combined fan-motor efficiency. Values for D typically range from 0.4 to 0.7. Likewise, the electricity required by a recycle pump can be calculated using Equation 1.105:

$$Energy_{pump} = \frac{(0.746)(2.52\times10^{-4})L_i(pressure)}{\varepsilon} \tag{1.105}$$

where 0.746 is the factor used to convert horsepower to kW, pressure is expressed in feet of water, and $\mathcal{E}$ is the combined pump-motor efficiency.

The cost of electricity (C_e) is then given by:

$$C_e = Energy_{fan+pump}\begin{pmatrix} annual \\ operating \\ hours \end{pmatrix}\begin{pmatrix} cos\,t\,of \\ electricity \end{pmatrix} \tag{1.106}$$

where cost of electricity is expressed in units of $/kW-hr.

1.3.4.2 Indirect Annual Costs

Indirect annual costs (IAC) include overhead, taxes, insurance, general and administrative (G&A), and capital recovery costs. The suggested factors for each of these items are shown in Table 1.9. Overhead is assumed to be equal to 60% of the sum of operating, supervisory, maintenance labor and maintenance materials. Overhead cost is discussed in Section 1, Chapter 2 of this Manual.

The system capital recovery cost, *CRC*, is based on an estimated 30-year equipment life, which is consistent with the estimate used for wet and dry FGD systems stated earlier in this chapter.[15] For a 30-year equipment life and an interest rate of 3.25%, the capital recovery factor (CF) is 0.0527. The system capital recovery cost is then estimated by:

$$CRC = 0.0527 \times TCI \tag{1.107}$$

G&A costs, property tax, and insurance are factored from the total capital investment (TCI), typically at 2%, 1%, and 1%, respectively.

1.3.4.3 Total Annual Cost

Total annual cost (TAC) is calculated by adding the direct annual costs and the indirect annual costs.

$$TAC = DAC + IAC \tag{1.108}$$

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[15] See Section 1, Chapter 2 of this Control Cost Manual for a discussion of the capital recovery cost.

1.3.4.4 Cost Effectiveness

The cost in dollars per ton of pollutant removed per year, is calculated as follows:

$$Cost\ Effectiveness\ = \frac{TAC}{Pollutant\ Removed(tons/yr)} \tag{1.109}$$

Where:

Cost Effectiveness = the cost effectiveness, \$/ton
SO_2 Removed/yr = annual mass of SO_2 removed by the absorber, ton/yr
TAC = total annual cost, \$/year

1.3.5 Example Problem for a Wet Packed Tower Absorber

The example problem presented in this section shows how to apply the sizing and costing procedures presented in Sections 1.3.2 through 1.3.4 to design and cost wet tower absorber for removing HCl from a waste gas stream consisting of HCl and air from an industrial process. This example problem uses the same outlet stream parameters presented in example problem found in Chapter 2 (Incinerators and Oxidizers) of this Manual for thermal and catalytic oxidizers. The waste gas stream entering the gas absorber is assumed to be saturated with moisture due to being cooled in the quench chamber. The concentration of HCl has also been adjusted to account for the change in volume.

1.3.5.1 Determine the Waste Stream Characteristics

The first step in the design procedure is to specify the conditions of the gas stream to be controlled and the desired pollutant removal efficiency. Gas and liquid stream parameters for this example problem are listed in Table 1.10.

The quantity of HCl can be written in terms of lb-moles of HCl per lb-moles of pollutant-free-gas (Y_i) using the following calculation:

$$Y_i = \frac{0.001871}{1-0.001871} = 0.00187\frac{lb-molesHCl}{lb-\text{molepollutantfreegas}}$$

The solvent, a dilute aqueous solution of caustic, is assumed to have the same physical properties as water.

1.3.5.2 Determine Gas and Liquid Stream Properties

Once the properties of the waste gas stream entering the absorber are known, the properties of the waste gas stream exiting the absorber and the liquid streams entering and exiting the absorber need to be determined. The pollutant concentration in the entering liquid (X_i) is assumed to be zero. The pollutant concentration in the exiting gas stream (Y_o) is calculated using Equation 1.57 and a removal efficiency of 99 percent.

$$Y_o = Y_i\left(1-\frac{\eta}{100}\right) = 0.00187\left(1-\frac{99}{100}\right) = 0.0000187$$

The liquid flow rate entering the column is calculated from the L_s/G_s ratio using Equation 1.58. Since Y_i, Y_o, and X_i are defined, the remaining unknown, X_o^*, is determined by consulting the equilibrium curve. A plot of the equilibrium curve-operating line graph for an HCl-water system is presented in Figure 1.12. The value of X_o^* is taken at the point on the equilibrium curve where Y_i intersects the curve. The value of Y_i intersects the equilibrium curve at an X value of 0.16.

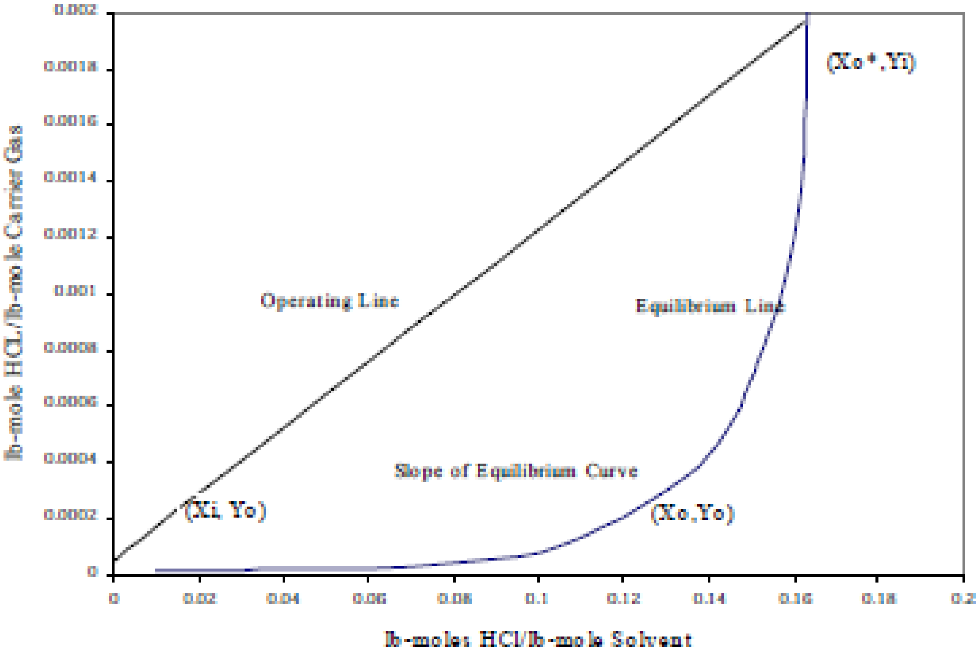

Figure 1.13: Equilibrium Curve Operating Line for the HCl-Water System [8]

Table 1.10: Waste Stream Characteristics and Packing Properties for Example Problem

Parameters	Values
Stream Properties	
Waste Gas Flow Rate Entering Absorber	21,377 scfm (22,288 acfm)
Temperature of Waste Gas Stream	100°F
Pollutant in Waste Gas	HCl
Concentration of HCl Entering Absorber in Waste Gas	1,871 ppmv
Pollutant Removal Efficiency	99% (molar basis)
Solvent	Water with caustic in solution
Density of Waste Gas[a]	0.0709 lb/ft^3
Density of Liquid [8]	62.4 lb/ft^3
Molecular Weight of Waste Gas[a]	29 lb/lb-mole
Molecular Weight of Liquid [8]	18 lb/lb-mole
Viscosity of Waste Gas[a]	0.044 lb/ft-hr
Viscosity of Liquid [8]	2.16 lb/ft-hr
Minimum Wetting Rate [8]	1.3 ft^2/hr
Pollutant Properties[b]	
Diffusivity of HCl in Air	0.725 ft^2/hr
Diffusivity of HCl in Water	1.02 x 10^{-4} ft^2/hr
Packing Properties[c]	
Packing type	2-inch ceramic Raschig rings
Packing factor: F_p	65
Packing constant: α	3.82
Packing constant: β	0.41
Packing constant: γ	0.45
Packing constant: φ	0.0125
Packing constant: b	0.22
Surface Area to Volume Ratio	28

[a] Reference [8], at 100°F

[b] Appendix C.

[c] Appendix B.

The operating line is constructed by connecting two points: (X_i, Y_o) and (X_o^*, Y_i). The slope of the operating line intersecting the equilibrium curve, $(L_s/G_s)_{min}$, is:

$$\left(\frac{L_s}{G_s}\right)_{min} = \frac{Y_i - Y_o}{X_o^* - X_i} = \frac{0.00187 - .0000187}{0.16 - 0} = 0.0116$$

The actual L_s/G_s ratio is calculated using Equation 1.60. For this example, an adjustment factor of 1.5 will be used.

$$\mathrm{G_s} = \frac{60\rho_{\mathrm{G}}\mathrm{G_i}}{\mathrm{MW_G}(1 + \mathrm{Y_i})}$$

$$G_s = \frac{\left(\left(60\frac{min}{hr}\right)\left(0.0709\frac{lb}{ft^3}\right)(22{,}288acfm)\right)}{\left(29\frac{lb}{lb - mole}\right)(1 + 0.00187)\frac{lb - moles}{hr}} = 3{,}263\ \frac{lb - moles}{hr}$$

The flow rate of the solvent entering the absorber may then be calculated using Equation 1.61.

$$L_s = \left(\frac{L_s}{G_s}\right)_{act} \times G_s = 0.0174\left(3{,}263\frac{lb - moles}{hr}\right) = 56.8\frac{lb - moles}{hr}$$

The values of $G_{mol,i}$ and $L_{mol,i}$ are calculated using Equations 1.62 and 1.63, respectively:

$$G_{mol,i} = G_s(1 + Y_i) = \left(3{,}263\frac{lb - moles}{hr}\right)(1 + 0.00187) = 3{,}269\frac{lb - moles}{hr}$$

$$L_{mol,i} = L_s(1 + X_i) = \left(56.8\frac{lb - moles}{hr}\right)(1 + 0) = 56.8\frac{lb - moles}{hr}$$

The pollutant concentration existing the absorber in the liquid is calculated using Equation 1.66.

$$x_o = \frac{Y_i - Y_o}{\left(\frac{L_s}{G_s}\right)} + X_i = \frac{0.00187 - 0.0000187}{0.0174} = \frac{0.106lb - molesHCl}{lb - molesolvent}$$

1.3.5.3 Calculate Absorption Factor

The absorption factor is calculated from the slope of the equilibrium line and the $L_{mol,i}/G_{mol,i}$ ratio. The slope of the equilibrium curve is based on the mole fractions of x_i, x_o, y_i, and y_o, which are calculated from X_i, X_o, Y_i and Y_o^* from Figure 1.12. From Figure 1.12, the value of Y_o^* in equilibrium with the X_o value of 0.106 is 0.0001. The values of Y_i and X_i are 0.

The mole fraction values are calculated from the concentration values using Equations 1.69 through 1.72.

$$x_o = \frac{0.106}{1 + 0.106} = 0.096$$

$$y_o^* = \frac{0.0001}{1 + 0.0001} = 0.0001$$

The slope of the equilibrium fine from x_i to x_o is calculated from Equation 1.68:

$$\mathrm{m} = \frac{\mathrm{y_o^* - y_i^*}}{\mathrm{x_o - x_i}} = \frac{0.0001 - 0}{0.096 - 0} = 0.00104$$

Since HCl is very soluble in water, the slope of the equilibrium curve is very small. The absorption factor is calculated from Equation 1.67.

$$\mathrm{AF} = \frac{\mathrm{L_{mol,i}}}{\mathrm{mG_{mol,i}}} = \frac{0.0174}{0.00104} = 17$$

1.3.5.4 Estimate Column Diameter

Once the inlet and outlet stream conditions are determined, the diameter of the gas absorber may be calculated using the modified generalized pressure drop correlation presented in Figure 1.9. The *Abscissa* value from the graph is calculated from Equation 1.73:

$$Abscissa = \left(\frac{L_{mol,i}}{G_{mol,i}}\right)\left(\frac{MW_L}{MW_G}\right)\sqrt{\frac{\rho_G}{\rho_L}}$$

$$Abscissa = 0.0174\left(\frac{18}{29}\right)\sqrt{\frac{0.0709}{62.4}} = 0.000364$$

Since this value is outside the range of Figure 1.9, the smallest value (0.01) will be used as a default value. The *Ordinate* is calculated from Equation 1.75.

$$Ordinate = 10^{\left[-1.668 - 1.085(log\,Abscissa) - 0.297(log\,Abscissa)^2\right]}$$

$$Ordinate = 10^{\left[-1.668 - 1.085(log\ \ 0.01) - 0.297(log\ \ 0.01)^2\right]} = 0.207$$

The superficial gas flow rate, $G_{sfr,i}$, is calculated using Equation 1.74. For this example calculation, 2-inch ceramic Raschig rings are selected as the packing. The packing factors for Raschig rings are listed in Appendix B.

$$\mathrm{G_{sfr,i}} = \sqrt{\frac{\rho_l\ \rho_G\ g_c\ (\mathrm{Ordinate})}{F_p\ \Psi\ \left(\frac{\mu_L}{2.42}\right)^{0.2}}}$$

$$G_{sfr,i} = \sqrt{\frac{(0.207)(62.4)\left(0.0709\,\frac{lb}{ft^3}\right)\left(32.2\,\frac{ft}{sec^2}\right)}{(65)(1)(0.893)^{0.2}}} = 0.681\,\frac{lb}{sec-ft^2}$$

Once $G_{sfr,i}$ is determined, the cross-sectional area of the column is calculated using Equation 1.77.

$$A = \frac{G_{mol,i}\ MW_G}{3{,}600\ G_{sfr,i}f} = \frac{\left(3{,}263\,\frac{lb-mol}{hr}\right)\left(29\,\frac{lb}{lb-mol}\right)}{\left(\left(3{,}600\,\frac{sec}{hr}\right)\left(0.681\,\frac{lb}{sec-ft^2}\right)(0.7)\right)^2} = 55.1\,ft^2$$

The superficial liquid flow rate is determined using Equation 1.79.

$$L_{sfr,i} = \frac{L_{mol,i}\ MW_L}{A} = \frac{\left(56.8\,\frac{lb-mol}{hr}\right)\left(18\,\frac{lb}{lb-mol}\right)}{55.1ft^2} = 18.6\,\frac{lb}{hr-ft^2}$$

At this point, it is necessary to determine if the liquid flow rate is sufficient to wet the packed bed. The minimum value of $L_{sfr,i}$ is calculated using Equation 1.80. The packing constant, *a*, is found in Appendix B, Table B-1.

$$\left(L_{sfr,i}\right)_{min} = MWR\rho_L a = \left(1.3\,\frac{ft^2}{hr}\right)\left(62.4\,\frac{lb}{ft^3}\right)\left(28\,\frac{ft^2}{ft^3}\right) = 2{,}271\,\frac{lb}{hr-ft^2}$$

The $L_{sfr,i}$ value calculated using the L/G ratio is far below the minimum value needed to wet the packed bed. Therefore, the new value, $(L_{sfr,i})_{min}$ will be used to determine the diameter of the absorber. The calculations for this revised diameter are shown in Appendix D. Appendix D shows that the cross-sectional area of the column is calculated to be 60 ft^2, $L_{mol,i}$ is 7572 lb-mole/hr, and $G_{sfr,i}$ is 0.627 lb/sec-ft^2. The diameter of the column is then calculated using Equation 1.78:

$$D = \sqrt{\frac{(4)(60ft)^2}{\pi}} = 8.74ft$$

The value of X_o is then:

$$x_o = \frac{0.00187 - 0.0000187}{\frac{7{,}572}{3{,}263}} = 0.0008$$

Expressed in terms of mole fraction:

$$x_o = \frac{0.0008}{1 - 0.0008} = 0.0008$$

The value of y_o in equilibrium with x_o cannot be estimated accurately. However, the value will approach zero, and the value of AF will be extremely large:

$$AF = \frac{7{,}572}{(3{,}263)(\approx 0)} \rightarrow \infty$$

1.3.5.5 Calculate Column Surface Area

Since $x_i = 0$ and *AF* is large, Equation 1.82 will be used to calculate the number of transfer units:

$$\mathrm{N_{tu}} = \frac{\mathrm{In}\left[\left(\frac{\mathrm{y_i - mx_i}}{\mathrm{y_o - mx_i}}\right)\left(1 - \frac{1}{\mathrm{AF}}\right) + \frac{1}{\mathrm{AF}}\right]}{1 - \frac{1}{\mathrm{AF}}} = ln\left(\frac{0.00187}{0.0000187}\right) = 4.61$$

The height of a transfer unit is calculated from, *AF*, H_L, and H_G. The values of H_G and H_L are calculated from Equations 1.87 and 1.88:

$$\mathrm{H_G} = \left[\alpha \frac{(3{,}600\mathrm{fG_{sfr,i}})^{\beta}}{(\mathrm{L_{sfr,i}})^{\Gamma}}\right]\sqrt{\frac{\mu_G}{\rho_G \mathrm{D_G}}}\left(\frac{\mu_L^{\Gamma}}{\mu_G^{\beta}}\right)$$

$$H_G = \frac{3.82[(3{,}600)(0.7)(0.627)]^{0.41}}{2{,}271^{0.45}}\sqrt{\frac{0.044}{(0.725)(0.0709)}} = 2.24ft$$

$$H_L = \phi\left(\frac{L_{sfr,i}}{\mu_L}\right)^{b}\sqrt{\frac{\mu_L}{\rho_L D_L}}\left(\frac{T}{286}\right)^{-4.255}$$

$$H_L = 0.0125\left(\frac{2{,}271}{2.16}\right)^{0.22}\sqrt{\frac{2.16}{(0.000102)(62.4)}} = 1.06ft$$

The height of the transfer unit is calculated using Equation 1.84:

$$H_{tu} = H_G + \frac{1}{AF} H_L = (2.24ft) + \frac{1}{\infty}(1.06ft) = 2.24ft$$

The depth of packing is calculated from Equation 1.81.

$$H_{pack} = N_{tu} \times H_{tu} = (4.61 \times 2.24ft) = 10.3ft$$

The total height of the column is calculated from Equation 1.89:

$$H_{tower} = 1.4H_{pack} + 1.02D + 2.81 = 1.40(10.3) + 1.02(8.74) + 2.81 = 26.1ft$$

The surface area of the column is calculated using Equation 1.90:

$$s = \pi D\left(H_{tower} + \frac{D}{2}\right) = (3.14)(8.74)\left(26.1 + \frac{8.74}{2}\right) = 836ft^2$$

1.3.5.6 Calculate Pressure Drop

The pressure drop through the column is calculated using Equation 1.91.

$$\Delta P = c\ 10^{\left(\frac{j\ L_{sfr,i}}{3{,}600}\right)} \frac{(fG_{sfr,i})^2}{\rho_G}$$

$$\Delta P = (0.24)10^{\left(\frac{(0.17)(2{,}271)}{3{,}600}\right)} \frac{[(0.7)(0.627)]^2}{0.0709}$$

= 0.83 inches water/foot packing

The total pressure drop (through 10.3 feet of packing) equals 8.55 inches of water.

1.3.5.7 Equipment Costs

Once the system sizing parameters have been determined, the equipment costs can be calculated. For this example, a gas absorber constructed of FRP will have its equipment costs estimated using Equation 1.97. The cost is converted to the current year dollars using the Chemical Engineering Plant Cost Index (CEPCI). In this example, the scaling factor for 2016 dollars is 541.7/390.6.[16]

$$TTC(\$) = 115S \times \left[\frac{CEPCI_{2016}}{390.6}\right] = 115(836) \times \left[\frac{541.7}{390.6}\right] = \$133{,}331$$

The cost of 2-inch ceramic Raschig rings can be estimated from packing cost ranges presented in Section 1.3.5. The volume of packing required is calculated as:

$$Volume\ ofpacking = (60ft^2)(10.3ft) = 618ft^3$$

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[16] Escalation of the equipment costs from 1991 dollars to 2016 dollars is not typically recommended according to Section 1, Chapter 2 of this Manual. For purposes of the example, however, we have escalated the costs since we do not have more recent equipment costs currently available. If they are available and documented, more recent equipment costs are preferred for cost estimates as they avoid escalation beyond five years, as mentioned in Chapter 1, Section 2 of the Control Cost Manual.

Using the average of the cost range for 2-inch ceramic packings, the total cost of packing is:

$$Packing\ cos\,t = (618ft^3)(\$20/ft^3)\left(\frac{541.7}{390.6}\right) = \$17{,}141$$

For this example problem, the cost of a pump will be estimated using vendor quotes. First, the flow rate of solvent must be converted into units of gallons per minute:

$$L(gpm) = \left(2{,}271\frac{lb}{h - ft^2}\right)(60ft^2)\left(\frac{gal}{8.34lb}\right)\left(\frac{hr}{60min}\right) = 272\ gpm$$

The average price for an FRP pump of this size is \$22.2/gpm at a pressure of 60 ft water, based on information from vendors (\$16/gpm in 1991 dollars escalated to 2016 dollars using CEPCI).[40] Therefore, the cost of the recycle pump is estimated as:

$$C_{pump} = (272gpm)\left(\frac{\$22.2}{gpm}\right) = \$6{,}038$$

For this example, the cost for a fan (FRP, backwardly-inclined centrifugal) can be calculated using the following equation:[46]

$$C_{fan} = 57.9d^{1.38} \times \left[\frac{541.7}{390.6}\right]$$

where *d* is the impeller (wheel) diameter of the fan expressed in inches. For this gas flow rate and pressure drop, an impeller diameter of 33 inches is needed. At this diameter, the cost of the fan is:

$$C_{fan} = 57.9(33)^{1.38} \times \left[\frac{541.7}{390.6}\right] = \$10{,}006$$

The cost of a fan motor (three-phase, carbon steel) with V-belt drive, belt guard, and motor starter can be computed as follows:[46]

$$C_{motor} = 104(hp)^{0.821} \times \left[\frac{547.1}{390.6}\right]$$

As will be shown in Section 1.3.5.8, the electricity consumption of the fan is 32.0 kW. Converting to horsepower, we obtain a motor size of 42.6 hp. The cost of the fan motor is:

$$C_{motor} = 104(42.6)^{0.821} \times \left[\frac{541.7}{390.6}\right] = \$3{,}139$$

The total auxiliary equipment cost is:

\$6,038 + \$10,006+ \$3,139 = \$19,183

The total equipment cost in 2016 dollars is the sum of the absorber cost, the packing cost, and the auxiliary equipment cost:

$$EC = TTC + Packing\ Cost + Auxiliary\ Equipment = 133{,}331 + 17{,}141 + 19{,}183 = \$169{,}655$$

The purchased equipment cost including instrumentation, controls, taxes, and freight is estimated using Equation 1.99:

$$PEC = 1.18\ EC = 1.18(169{,}655) = \$200{,}193$$

The total capital investment is calculated using Equation 1.100 and using a contingency factor of 10%:

$$TCI = 2.17PEC + C = 2.17(200{,}193) + (0.1)(2.17)(200{,}193) = \$477{,}861$$

1.3.5.8 Total Annual Cost

Table 1.11 summarizes the estimated annual costs using the suggested factors and unit costs for the example problem.

Direct annual costs for gas absorber systems include labor, materials, utilities, and wastewater disposal. Labor costs are based on 8,000 hr/yr of operation. Supervisory labor is computed at 15% of operating labor and operating and maintenance labor are each based on 0.5 hr per 8-hr shift.

The electricity required to run the fan is calculated using Equation 1.104 and assuming a combined fan-motor efficiency of 70%:

$$\mathrm{Energy_{fan}} = \frac{1.17 \times 10^{-4}\mathrm{G_i DP}}{\varepsilon} = \frac{(1.17 \times 10^{-4})(22{,}288)(8.55)}{0.70} = 32.0\ kW$$

The energy required for the liquid pump is calculated using Equation 1.105. The capital cost of the pump was calculated using data supplied by vendors for a pump operating at a pressure of 60 feet of water. Assuming a pressure of 60 ft of water a combined pump-motor efficiency of 70 percent:

$$Energy_{pump} = \frac{(0.746)(2.52 \times 10^{-4})L_i(pressure)}{\varepsilon} = \frac{(0.746)(2.52 \times 10^{-4})(272)(60)(1)}{0.70}$$
$$= 4.4\ kw$$

The total energy required to operate the auxiliary equipment is approximately 36.4 kW. The cost of electricity, C_e, is calculated using Equation 1.106 and with the cost per kWh shown in Table 1.11.

$$C_e = \left(Energy_{fan} + Energy_{pump}\right)(Operating\ Hours)(Cost\ of\ Electricty)$$
$$= (36.4kW)(8{,}000h/yr)(\$0.0674/kWh) = \$19{,}627/yr$$

The costs of solvent (water), wastewater disposal, and caustic are all dependent on the total system throughput and the fraction of solvent discharged as waste. A certain amount of solvent will be wasted and replaced by a fresh solution of water and caustic in order to maintain the system's pH and solids content at acceptable levels. Based on data provided by vendors, a maximum solids content of 10% by weight will be the design basis for this example problem.[46] The following calculations illustrate the procedure used to calculate how much water and caustic are needed, and how much solvent must be bled off to maintain system operability.

From previous calculations, $L_{mol,i}$ = 7,572 lb-moles/hr. The mass flow rate is calculated as:

$$L_{mass} = \left(7{,}572 \frac{lb - mole}{hr}\right)\left(18 \frac{lb}{lb - mole}\right) = 136{,}296 \frac{lb}{hr}$$

With $G_{mol,i}$ at 3,263 lb-moles/hr, the mass flow rate of the gas stream is calculated as:

$$G_{mass} = \left(3{,}263 \frac{lb - mole}{hr}\right)\left(29 \frac{lb}{lb - mole}\right) = 94{,}627 \frac{lb}{hr}$$

The amount of HCl in the gas stream is calculated on a molar basis as follows:

$$G_{mass} = \left(3{,}263 \frac{lb - mole}{hr}\right)\left(1{,}874 \frac{ppmv}{1 \times 10^6}\right) = 6.11 \frac{lb - molHCl}{hr}$$

On a mass basis:

$$G_{mass,HCl} = \left(6.11 \frac{lb - molHCL}{hr}\right)\left(36.5 \frac{lb}{lb - mole}\right) = 223.0 \frac{lbHCl}{hr}$$

For this example, the caustic is assumed to be Na_2O, with one mole of caustic required for neutralizing 2 moles of HCl. Therefore, 3.06 lb-moles/hr of caustic are required.

The unit cost of a 76% solution of Na_2O is given in Table 1.11. The annual cost is calculated from:

$$C_c = \left(3.06 \frac{lb{-}moles}{hr}\right)\left(62 \frac{lb}{lb{-}mole}\right)\left(\frac{8{,}000hr}{yr}\right)\left(\frac{ton}{2{,}000lb}\right)\left(\frac{1}{0.76}\right)\left(\frac{\$340}{ton}\right) = \$339{,}499/year$$

Mass of the salt formed in this chemical reaction, sodium chloride (NaC1), is calculated as:

$$Mass_{NaCl} = \left(223.0 \frac{lb - HCl}{hr}\right)\left(\frac{lb - mole}{36.5lbHCl}\right)\left(\frac{1lb - moleNaCl}{lb - moleHCl}\right)\left(\frac{58.5lbNaCl}{lb - moleNaCl}\right)$$
$$= 357.4lb\ NaCl/hour$$

If the maximum concentration of NaC1 in the wastewater (ww) is assumed to be 10 weight %, the wastewater volume flow rate is calculated as:

$$Wastewater_{flowrate} = \left(357.4\frac{lb\ \ NaCl}{hr}\right)\left(\frac{1lb\,ww}{0.1lb\ NaCl}\right)\left(\frac{gal\,ww}{8.34\,lb\,ww}\right)\left(\frac{1hr}{60\min}\right)$$

$$= 7.14\,gpm$$

Where 8.34 is the density of the wastewater.

The cost of wastewater disposal is:[17]

$$C_{ww} = (7.14gpm)\left(\frac{60min}{hr}\right)\left(\frac{8{,}000hrs}{yr}\right)\left(\frac{\$5.70}{1{,}000gal}\right) = \$19{,}535$$

The cost of solvent (water) is:

$$C_s = (7.14gpm)\left(\left(\frac{60min}{1hr}\right)\left(8{,}000\frac{hr}{yr}\right)\left(\frac{\$4.20}{1{,}000gal}\right)\right) = \frac{\$14{,}394}{year}$$

Indirect annual costs include overhead, administrative charges, property tax, insurance, and capital recovery. Total annual cost is estimated using Equation 1.108. For this example, the total annual cost is estimated to be $516,000 per year (Table 1.11).

Table 1.11: Annual Costs for Packed Tower Absorber Example Problem

Cost Item	Calculations	Cost
Direct Annual Costs, DAC		
Operating Labor		
Operator[(a)]	$\frac{0.5hr}{shift} \times \frac{shift}{8hr} \times \frac{8{,}000hr}{yr} \times \frac{\$26.61}{hr}$	$13,305
Supervisor	$15\%\ of\ operator = 0.15\ \times 13{,}305$	$1,996
Operating materials		
Solvent (water)[(b)]	$7.14gpm \times \frac{60\ min}{hr} \times \frac{8{,}000\ hr}{yr} \times \frac{\$4.20}{1{,}000\ gal}$	$14,394

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[17] Because the wastewater stream contains only NaC1, it probably will not require pretreatment before discharge to a municipal wastewater treatment facility. Therefore, the wastewater disposal unit cost shown here is just a sewer usage rate. This unit cost ($5.25/1,000 gal) is the average sewer rates for industrial facilities in 2013 compiled by Black & Veatch. See “50 Largest Cities Water/Wastewater Rate Survey.” Available at http://www.saws.org/who_we_are/community/RAC/docs/2014/50-largest-cities-brochure-water-wastewater-rate-survey.pdf.

Cost Item	Calculations	Cost
Caustic Replacement(c)	$\frac{3.06\ lb-mole}{hr}\times\frac{62\ lb}{lb-mole}\times\frac{8{,}000\ hr}{yr}\times\frac{ton}{2{,}000\ lb}\times\frac{1}{0.76}\times\frac{\$340}{ton}$	$339,499
Wastewater disposal(b)	$7.14\ gpm\left(\frac{60\ min}{hr}\right)\left(\frac{8{,}000\ hr}{yr}\right)\left(\frac{\$5.70}{1{,}000\ gal}\right)$	$19,535
Maintenance		
Labor(d)	$\frac{0.5}{shift}\times\left(\frac{shift}{8\ hr}\right)\times\left(\frac{8{,}000\ hr}{yr}\right)\times(\$29.27/hour)$	$14,635
Material	100% of maintenance labor	$14,635
Electricity(e)	$36.4\ kw\times\frac{8{,}000\ hr}{yr}\times\frac{\$0.0676}{kWh}$	$19,685
Total DAC		$437,684
Indirect Annual Costs, IAC(f)		
Overhead	60% of total labor and maintenance material: = 0.6(13,305 + 1,996 + 14,635 + 14,635)	$27,943
Administrative charges	2% of Total Capital Investment = 0.02 ($477,861)	$9,557
Property tax	1% of Total Capital Investment = 0.01 ($477,861)	$4,779
Insurance	1% of Total Capital Investment = 0.01 ($477,861)	$4,779
Capital recovery(g)	0.0527 x $477,861	$25,183
Total IAC		$72,241
Total Annual Cost (rounded)	DAC + IAC	$509,925

(a) Bureau of Labor Statistics, May 2016 National Occupational Employment and Wage Estimates – United States, May 2016 (https://www.bls.gov/oes/2016/may/oes_nat.htm). Hourly rates for operators based on data for Plant and System Operators – other (51-8099).

(b) Average water rates for industrial facilities in 2016 compiled by Black & Veatch. See "50 Largest Cities Water/Wastewater Rate Survey – 2018-2019." Available at https://www.bv.com/sites/default/files/2019-10/50_Largest_Cities_Rate_Survey_2018_2019_Report.pdf.

(c) Price of caustic soda based on July 2016 data reported in *Outlook '17: The Calm Before the Storm for European Caustic Soda*, ICIS, January 11, 2017 (available at https://www.icis.com/explore/resources/news/2017/01/11/10064199/outlook-17-the-calm-before-the-storm-for-european-caustic-soda/).

(d) Maintenance labor rate is estimated at 110% of the operator wage rate.

(e) U.S. Energy Information Administration. Electric Power Annual 2016. Table 2.10. Published December 2017. Available at: https://www.eia.gov/electricity/annual/pdf/epa.pdf.

(f) Indirect annual costs include overhead, administrative charges, property tax, insurance, and capital recovery. Total annual cost is estimated using Equation 1.108.

(g) The capital recovery cost factor, CRF, is a function of the absorber equipment life and the opportunity cost of the capital (i.e., interest rate). For this example, we assume a 30-year equipment life and a 3.25% interest rate.

1.3.5.9 Cost Effectiveness

The cost in dollars per ton of pollutant removed per year, is calculated as follows:

$$Cost\ Effectiveness\ = \frac{TAC}{HClRemoved/year} = \frac{\$509{,}925/year}{894tons/year} = \$570/ton$$

1.3.5.10 Alternate Example

In this example problem the diameter of a gas absorber will be estimated by defining a pressure drop. A pressure drop of 1 inch of water per foot of packing will be used in this example calculation. Equation 1.94 is used to calculate the ordinate value relating to an abscissa value. If the $L_{mole,i}/G_{mol,i}$ ratio is known, the Abscissa can be calculated directly. The Ordinate value is:

$$\mathrm{Ordinate} = \exp[\,\mathrm{k}_0 + \mathrm{k}_1(1\mathrm{nAbscissa}) + \mathrm{k}_2(1\mathrm{nAbscissa})^2 \\ +\mathrm{k}_3(1\mathrm{nAbscissa})^3 + \mathrm{k}_4(1\mathrm{nAbscissa})^4]$$

$$Ordinate = exp[-4.0950 - 1.00121n(0.0496) - 0.1587(1n0.0496)^2 + \\ 0.0080(1n0.0496)^3 + 0.0032(1n0.0496)^4] \\ = 0.084$$

The value of G_{sfr} is calculated using Equation 1.95:

$$\mathrm{G}_{\mathrm{sfr,i}} = \sqrt{\frac{(\rho_L - \rho_G)\rho_G g_c(\mathrm{Ordinate})}{\mathrm{F_P}\left(\frac{\mu_L}{2.42}\right)^{0.1}}}$$

$$G_{sfr,i} = \sqrt{\frac{(62.4 - 0.0709)(0.0709)(32.2)(0.084)}{65(0.893)^{0.1}}} = 0.43\frac{lb}{ft^2 - sec}$$

The remaining calculations are the same as in Section 1.3.5.4, except the flooding factor is not used in the equations.

Acknowledgments

The authors gratefully acknowledge the following companies and organizations for contributing data to this chapter:

- Air Plastics, Inc. (Cincinnati, OH)
- April, Inc. (Teterboro, NJ)
- Anderson 2000, Inc. (Peachtree City, GA)
- Calvert Environmental (San Diego, CA)
- Ceilcote Air Pollution Control (Berea, OH)
- Croll-Reynolds Company, Inc. (Westfield, NJ)
- Ecolotreat Process Equipment (Toledo, OH)
- Glitsch, Inc. (Dallas, TX)
- Institute of Clean Air Companies (ICAC) (Arlington, VA)
- Interel Corporation (Englewood, CO)
- Jaeger Products, Inc. (Spring, TX)
- Koch Engineering Co., Inc. (Wichita, KS)
- Lantec Products, Inc. (Agoura Hills, CA)
- Midwest Air Products Co., Inc. (Owosso, MI)
- Monroe Environmental Corp., (Monroe, MI)
- Norton Chemical Process Products (Akron, OH)

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Appendix A

Table A-1: List of Design Variables For Wet FGD Systems

Variable	Symbol	Units
Limestone consumption rate	$Q_{Limestone}$	tons/hour
Sulfur emission rate	S	lb/MMBtu
Unit size (gross)	A	MW
Heat rate factor	HRF	—
Removal efficiency	EF	fraction
Make-up water consumption rate	q_{water}	kgallons/hr
Coal factor	$CoalF$	—
Waste generation rate	q_{waste}	tons/hour
Electricity consumption	P	kW
Absorber island cost	$ABScost$	$
Reagent preparation cost	$RPEcost$	$
Waste handling equipment cost	$WHEcostl$	$
Balance of plant costs	$BOPcost$	$
Wastewater treatment plant cost	$WWTcost$	$
Retrofit factor	RF	—
Elevation factor	$ELEVF$	—
Number of hours of wastewater treatment	t_{op}	Hours/year
Capital recovery factor	CRF	—

Table A-2: List of Design Variables For Dry FGD Systems

Variable	Symbol	Units
Lime consumption rate	Q_{Lime}	tons/hour
Sulfur emission rate	S	lb/MMBtu
Unit size (gross)	A	MW
Heat rate factor	HRF	—
Removal efficiency	EF	fraction
Make-up water consumption rate	q_{water}	kgallons/hr
Coal factor	$CoalF$	—
Waste generation rate	q_{ww}	tons/hour
Electricity consumption	P	kW
Absorber island cost	$ABScost$	$
Balance of plant costs	$BOPcost$	$
Retrofit factor	RF	—
Elevation factor	$ELEVF$	—
Annual lime cost	$ARcost$	$
Number of operating hours	t_{op}	Hours/year
Capital recovery factor	CRF	—

Table A-3: List of Design Variables For Wet Packed Tower Absorbers

	Variable	Symbol	Units
►	Surface to volume ratio of packing	a	ft^2/ft^3
	Cross-sectional area of absorber	A	ft^2
	Abscissa value from plot of generalized press drop correlation	Abscissa	—
	Absorption factor	AF	—
	Diameter of absorber	D	feet
►	Diffusivity of pollutant in gas	D_G	ft^2/hr
►	Diffusivity of pollutant in liquid	D_L	ft^2/hr
►	Flooding factor	f	—
►	Packing factor	Fp	—
►	Waste gas flow rate entering absorber	G_i	acfm
	Waste gas flow rate exiting absorber	Go	acfm
	Waste gas molar flow rate entering absorber	$Gmol$	lb-moles/h
	Molar flow rate of pollutant free gas	Gs	lb-moles/h
	Waste gas superficial flow rate entering absorber	$Gsfr,i$	lb/sec-ft^2
	Height of gas transfer unit	H_G	feet
	Height of liquid transfer unit	H_L	feet
	Height of overall transfer unit	Htu	feet
	Height of packing	$Hpack$	feet
	Height of absorber	$Htower$	feet
	Pressure drop constants	$k0, k1, k2, k3, k_4$	—
	Liquid rate entering absorber	Li	gpm
	Liquid rate exiting absorber	Lo	gpm
	Liquid molar flow rate entering absorber	$Lmol,i$	lb-moles/h
	Molar flow rate of pollutant free solvent	Ls	lb-moles/h
	Liquid superficial flow rate entering absorber	$Lsfr,i$	lb/hr-ft^2
	Slope of equilibrium line	m	—
►	Molecular weight of gas stream	MW_G	lb/lb-mole
►	Molecular weight of the liquid stream	MW_L	lb/lb-mole
►	Minimum wetting rate	MWR	ft^2/hr
	Number of overall transfer units	Ntu	—
	Ordinate value from plot of generalized pressure drop correlation	Ordinate	—
	Surface area of absorber	S	ft^2
►	Temperature of solvent	T	K

Table A-3: List of Design Variables For Wet Packed Tower Absorbers (continued)

	Variable	Symbol	Units
	Maximum pollutant concentration in liquid phase in equilibrium with pollutant entering column in gas phase	X_o^*	$\frac{\text{lb} - \text{molespollutant}}{\text{lb} - \text{molespollutantfreesolvent}}$
	Pollutant concentration exiting absorber in liquid	X_o	$\frac{\text{lb} - \text{molespollutant}}{\text{lb} - \text{molespollutantfreesolvent}}$
	Mole fraction of pollutant entering absorber in waste gas	y_i	$\frac{\text{lb} - \text{molespollutant}}{\text{lb} - \text{moles}oftotalgas}$
	Mole fraction of pollutant in gas phase in equilibrium with mole fraction of pollutant entering in the liquid phase	y_i^*	$\frac{\text{lb} - \text{molespollutant}}{\text{lb} - \text{moles}oftotalgas}$
	Mole fraction of pollutant exiting scrubber in waste gas	y_o	$\frac{\text{lb} - \text{molespollutant}}{\text{lb} - \text{moles}oftotalgas}$
	Mole fraction of pollutant in gas phase in equilibrium with mole fraction of pollutant entering in the liquid phase	y_o^*	$\frac{\text{lb} - \text{molespollutant}}{\text{lb} - \text{moles}oftotalgas}$
►	Pollutant concentration entering scrubber in waste gas	Y_i	$\frac{\text{lb} - \text{molespollutant}}{\text{lb} - \text{molespollutantfree}gas}$
	Pollutant concentration entering scrubber in equilibrium with concentration in liquid phase	Y_i^*	$\frac{\text{lb} - \text{molespollutant}}{\text{lb} - \text{molespollutantfree}gas}$
	Pollutant concentration exiting scrubber in waste gas	Y_o	$\frac{\text{lb} - \text{molespollutant}}{\text{lb} - \text{molespollutantfree}gas}$
►	Pollutant removal efficiency	η	%
	Pollutant concentration exiting scrubber in equilibrium with concentration in liquid phase	Y_o	$\frac{\text{lb} - \text{molespollutant}}{\text{lb} - \text{moles}oftotalgas}$
►	Density of waste gas stream	ρ_G	lb/ft^3
►	Density of liquid stream	ρ_L	lb/ft^3
►	Viscosity of waste gas	μ_G	$lb/ft - hr$
►	Viscosity of solvent	μ_L	$lb/ft - hr$
	Ratio of solvent density to water	Ψ	—
	Pressure drop	ΔP	inches H_2O/feet of packing
►	Packing factors	$a, \alpha, \varphi, b, \beta, y, c, j$	—

► Denotes required input data.

Appendix B

Table B-1: Packing Factors for Various Packing Materials for Wet Packed Tower Absorbers [5, 8, 38, 41]

Packing Type	Construction Level	Nominal Diameter (inches)	Fp	a
Raschig rings	Ceramic	1/2	640	111
		5/8	380	100
		3/4	255	80
		1	160	58
		1 ½	95	38
		2	65	28
		3	37	
Raschig rings	Metal	½	410	118
		5/8	290	
		¾	230	72
		1	137	57
		1 ½	83	41
		2	57	31
		3	32	21
Pall rings	Metal	5/8	70	131
		1	48	66
		1 ½	28	48
		2	20	36
		3 1/2	16	
Pall rings	Polypropylene	58	97	110
		1	52	63
		1 ½	32	39
		2	25	31
Berl saddles	Ceramic	½	240	142
		¾	170	82
		1	110	76
		1 ½	65	44
		2	45	32
Intalox saddles	Ceramic	½	200	190
		¾	145	102
		1	98	78
		1 ½	52	60
		2	40	36
		3	22	
Tri-Packs®	Plastic	2	16	48
		3 ½	12	38

Table B-2: Packing Constants Used to Estimate H_G For Wet Packed Tower Absorbers [1, 5, 8, 41]

	Size	Packing Constants			Applicable Range[a]	
Packing Type	**(inches)**	**α**	**B**	**γ**	**G_{sfr}**	**L_{sfr}**
Raschig Rings	3/8	2.32	0.45	0.47	200-500	500-1,500
	1	7.00	0.39	0.58	200-800	400-500
	1	6.41	0.32	0.51	200-600	500-4,500
	1 1/2	1.73	0.38	0.66	200-700	500-1,500
	1 1/2	2.58	0.38	0.40	200-700	1,500-4,500
	2	3.82	0.41	0.45	200-800	500-4,500
Berl Saddles	1/2	32.4	0.30	0.74	200-700	500-1,500
	1/2	0.81	0.30	0.24	200-700	1,500-4,500
	1	1.97	0.36	0.40	200-800	400-4,500
	1 1/2	5.05	0.32	0.45	200-1,000	400-4,500
Partition Rings	3	640	0.58	1.06	150-900	3,000-10,000
LanPac®	2.3	7.6	0.33	-0.48	400-3,000	500-8,000
Tri-Packs®	2	1.4	0.33	0.40	100-900	500-10,000
	3 1/2	1.7	0.33	0.45	100-2,000	500-10,000

[a] Units of lb/hr-ft^2

Table B-3: Packing Constants Used to Estimate H_L For Wet Packed Tower Absorbers[1, 5, 41]

		Packing Constants		Applicable Range[a]
Packing Type	**Size (inches)**	**φ**	**b**	**L^a_{sfr}**
Raschig Rings	3/8	0.00182	0.46	400-15,000
	1	0.00357	0.35	400-15,000
	1 1/2	0.0100	0.22	400-15,000
	2 1/2	0.0111	0.22	400-15,000
	2	0.0125	0.22	400-15,000
Berl Saddles	1/2	0.00666	0.28	400-15,000
	1	0.00588	0.28	400-15,000
	1 1/2	0.00625	0.28	400-15,000
Partition Rings	3	0.0625	0.09	3,000-14,000
LanPac®	2.3	0.0039	0.33	500-8,000
	3.5	0.0042	0.33	500-8,000
Tri-Packs®	2	0.0031	0.33	500-10,000
	3 1/2	0.0040	0.33	500-10,000

[a] Units of lb/hr-ft^2

Table B-4-: Packing Constants Used to Estimate Pressure Drop For Wet Packed Tower Absorbers[1, 8, 41]

Packing Type	Construction Material	Nominal Diameter (inches)	c	j
Raschig rings	ceramic	1/2	3.1	0.41
		3/4	1.34	0.26
		1	0.97	0.25
		1 ¼	0.57	0.23
		1 ½	0.39	0.23
		2	0.24	0.17
Raschig rings	metal	5/8	1.2	0.28
		1	0.42	0.21
		11/2	0.29	0.20
		2	0.23	0.135
Pall rings	metal	5/8	0.43	0.17
		1	0.15	0.16
		11/2	0.08	0.15
		2	0.06	0.12
Berl saddles	ceramic	1/2	1.2	0.21
		3/4	0.62	0.17
		1	0.39	0.17
		11/2	0.21	0.13
Intalox saddles	ceramic	1/2	0.82	0.20
		3/4	0.28	0.16
		1	0.31	0.16
		11/2	0.14	0.14

[a] Units of lb/hr-ft^2

Appendix C

Table C-1: Physical Properties of Common Pollutants

Pollutant	**Molecular Weight** $\left(\frac{lb}{lb-mole}\right)$	**Diffusivity in Air at 25°C (cm^2/sec)[a]**	**Diffusivity in Water at 20°C ($cm^2/sec \times 10^5$)[a]**
Ammonia	17	0.236	1.76
Methanol	32	0.159	1.28
Ethyl Alcohol	46	0.119	1.00
Propyl Alcohol	60	0.100	0.87
Butyl Alcohol	74	0.09	0.77
Acetic Acid	60	0.133	0.88
Hydrogen Chloride	36	0.187	2.64
Hydrogen Bromide	36	0.129	1.93
Hydrogen Fluoride	20	0.753	3.33

[a] Diffusivity data taken from Reference [8, 48].

Appendix D

Minimum Wetting Rate Analysis

As explained in the design procedures, the liquid flow rate entering the column must be high enough to effectively wet the packing. If the liquid flow rate, as determined theoretically in Equation 1.79, is lower than the flow rate dictated by the minimum wetting rate, calculated in Equation 1.80, then the packing will not be wetted sufficiently to ensure mass transfer between the gas and liquid phases. The minimum liquid flow rate should then be used as a default value. The superficial gas flow rate, G_{sfr}, and cross-sectional area of the column must then be recalculated to account for the increased liquid flow rate. The approach is outlined below:

- The value of $L_{mol,i}$ must be recalculated from the value of $(L_{sfr,i})_{min}$ using the equation:

$$L_{mol,i} \frac{(L_{sfr,i})_{min}}{(MW)_L} \tag{D-1}$$

The value of *A* (the cross-sectional area of the absorber column) is the only unknown in the equation.

- The *Abscissa* value is calculated in terms of *A* by substituting the new $L_{mol,i}$ into Equation 1.73.

- The value of $G_{sfr,i}$ is recalculated by rearranging Equation 1.77, with A as the only unknown.

- The *Ordinate* value is calculated in terms of A from the new $G_{sfr,i}$ using the Equation 1.74.

- An iterative process is used to determine *A*, *Abscissa*, and *Ordinate*. Values of *A* are chosen and the *Abscissa* and *Ordinate* values are calculated. The *Ordinate* value corresponding to the *Abscissa* value is determined from Figure 1.9 (or Equation 1.75), and this value is compared to the *Ordinate* value calculated using Equation 1.74. This process is continued until both *Ordinate* values are equal.

Step 1: The first step is to recalculate the liquid flow rate. The liquid molar flow rate may be calculated using Equation 1.79.

$$L_{mol,i} = \left(2{,}271 \frac{lb}{hr-ft^2}\right)\left(\frac{lb-mole}{18lb}\right)A = \left(126.2 \frac{lb-mole}{hr-ft^2}\right)A \tag{D-2}$$

Step 2: The *Abscissa* value from Figure 1.9, and presented in Equation 1.73, is calculated as:

$$Abscissa = \frac{\left(126.2\frac{lb-mole}{hr-ft^2}\right)A}{3{,}263\frac{lb-mole}{hr}}\left(\frac{18}{29}\right)\sqrt{\frac{0.0709}{62.4}} \tag{D-3}$$

$$= 8.09x10^{-4}A$$

Step 3: The value of $G_{sfr,i}$ is then recalculated in terms of the cross-sectional area of the column.

$$G_{sfr} = \frac{\left(3{,}263\frac{lb-mole}{hr-ft^2}\right)\left(29\frac{lb}{lb-mole}\right)}{\left(3{,}600\frac{sec}{hr}\right)(0.7)(A)} = \frac{37.6}{A} \tag{D-4}$$

Step 4: The ordinate value from Figure 1.9, and presented in Equation 1.74, is calculated as:

$$Ordinate = \frac{\left(\frac{37.6}{A}\right)^2(65)(1)(0.893)^{0.2}}{(62.4)(0.0709)(32.2)} \tag{D-5}$$

$$= \frac{631}{A^2}$$

Step 5: At this point the simplest solution is an iterative approach. Choose a value for *A*, calculate the *Abscissa* value using Equation D-3, and find the corresponding *Ordinate* value off the flooding curve in Figure 1.9 (or use Equation 1.75 to calculate the Ordinate value). Compare the calculated *Ordinate* value from Equation D-5 to the value obtained from the graph or from Equation 1.75. By continuing this process until the *Ordinate* values converge the value of A is determined to be 60 ft^2. The following table illustrates the intermediate steps in the calculation process.

Table D-1: Results of Iterative Approach to Determining Column Cross-Sectional Area

Assumed Value of A	Abscissa Calculated From Equation D-1	Ordinate Calculated From Equation 9.81	Ordinate Calculated From Equation D-2
65	0.0526	0.1714	0.1493
62	0.0503	0.1740	0.1642
60	0.0485	0.1757	0.1752

The value of G_{sfr} is then:

$$G_{sfr} = \frac{37.6}{60} = 0.627 \frac{lb}{sec - ft^2}$$

The liquid molar flow rate is:

$$L_{mol,i} = (126.2)(60) = 7{,}572 \frac{lb}{sec - ft^2}$$

The diameter and height of the column using the results of this calculation are presented in the example problem shown in Section 1.3.5.