

CONTROLLING SO₂ EMISSIONS: A REVIEW OF TECHNOLOGIES

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Foreword

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E. Timothy Oppelt, Director
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Abstract

Sulfur dioxide (SO₂) scrubbers may be used by electricity generating units to meet the requirements of Phase II of the Acid Rain SO₂ Reduction Program. Additionally, the use of scrubbers can result in reduction of mercury and particulate matter emissions. It is timely, therefore, to review commercially available flue gas desulfurization (FGD) technologies that have an established record of performance.

The review of FGD technologies presented in this report describes these technologies, assesses their applications, and characterizes their performance. Additionally, the report describes some of the advances that have occurred in FGD technologies. Finally, the report presents an analysis of the costs associated with applications of limestone forced oxidation, lime spray dryer, and magnesium-enhanced lime FGD processes. The information presented in this paper should be useful to parties evaluating FGD technology applications.

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List of Symbols

<u>Symbol</u>	<u>Meaning</u>	<u>Unit</u>
A&S	Administration and support cost	dollars
ABSORBER	Absorber cost	dollars
ABSORBER 1	RLCS absorber cost	dollars
ABSORBER 2	Alloy absorber cost	dollars
ACFM	Flue gas flow into absorber	cfm
ACFM1	Flue gas flow out of absorber	cfm
ACFM2	Flue gas flow out of ID fans	cfm
ACFM3	Flue gas flow out of particulate control device	cfm
BARE MODULE _E	Support equipment area auxiliary cost	dollars
BARE MODULE _G	Flue gas handling area auxiliary cost	dollars
BARE MODULE _R	SO ₂ removal area auxiliary cost	dollars
BARE MODULE _W	Waste handling area auxiliary cost	dollars
BM	Capital cost for FGD system	dollars
BM _E	Capital cost component for support equipment area	dollars
BM _F	Capital cost component for reagent feed area	dollars
BM _G	Capital cost component for waste handling area	dollars
BM _R	Capital cost component for waste handling area	dollars
BM _W	Capital cost component for waste handling area	dollars
C _{B&H}	Cost of ball mill and hydroclones	dollars
C _{CaCO₃}	Cost of limestone	dollars
C _{CaO}	Cost of lime	dollars
C _{DBA}	Cost of DBA tank	dollars
C _{DL}	Cost of disposal with landfilling	dollars
C _{DS}	Cost of disposal with gypsum stacking	dollars
CF	Capacity factor	%
CHIMNEY	Cost of chimney	dollars
CREDIT	By-product credit	Dollars

<u>Symbol</u>	<u>Meaning</u>	<u>Unit</u>
F_{AFDC}	Allowance for funds during construction factor	-----
F_d	F-factor	scf/10 ⁶ Btu
F_{GPM}	Slurry flow rate	gal/min
Fixed _{O&M}	Fixed operation and maintenance cost	dollars
FR_L	Reagent feed rate	lb/hr
FR_{SO_2}	SO ₂ feed rate to the FDG System	lb/hr
F_{TCE}	Total cash expended factor	-----
HHV	Coal heating value	Btu/lb
HR	Plant heat rate	Btu/kWh
ID FANS	Cost of ID fans	dollars
L/G	Liquid-to-gas ratio	gal/1000 ft ³
ML&M	Maintenance, labor, and materials cost	dollars
N_a	Number of absorbers	-----
N_f	Number of fans	-----
N_p	Number of pumps	-----
OL	Operating labor cost	dollars
P	Percent oxygen in the stack	%
POWER	Cost of electrical energy	dollars
PUMP	Cost of pump	dollars
PUMPS	Cost of pumps	dollars
SPRAY DRYERS	Cost of spray dryers	dollars
SPRAY DRYERS1	Cost of RLCS spray dryers	dollars
SPRAY DRYERS2	Cost of alloy spray dryers	dollars
STEAM	Cost of steam	dollars
TER	Thermal energy required	Btu
THICKENER	Cost of thickener	dollars
Variable _{O&M}	Variable operation and maintenance cost	dollars
Wt%S	Coal sulfur content	Wt%

List of Acronyms and Abbreviations

CAAA	Clean Air Act Amendments of 1990
CFB	Circulating Fluidized Bed
CUECost	Coal Utility Environmental Cost Workbook
DBA	Dibasic Acid
DSI	Duct Sorbent Injection Process
EPA	United States Environmental Protection Agency
ESP	Electrostatic Precipitator
FGD	Flue Gas Desulfurization
FSI	Furnace Sorbent Injection Process
ID	Inside Diameter
L/G	Liquid-to-gas Ratio
LSD	Lime Spray Drying Process
LSFO	Limestone Forced Oxidation Process
LSIO	Limestone Inhibited Oxidation Process
MEL	Magnesium-Enhanced Lime Process
MW _e	Unit Electrical Generating Capacity
NAAQS	National Ambient Air Quality Standards
O&M	Operation and Maintenance
PM _{2.5}	Particulate Matter Less than 2.5 : m (Aerodynamic Diameter)
RLCS	Rubber-Lined Carbon Steel
SUSCM	State-of-the-art Utility Scrubber Cost Model
TCR	Total Capital Requirement
TPC	Total Plant Cost
TPI	Total Plant Investment
WESP	Wet Electrostatic Precipitator

Conversion Table – English Units to SI Units

<u>To Obtain</u>	<u>From</u>	<u>Multiply by</u>
m	ft	0.3048
m ²	ft ²	9.29 · 10 ⁻²
m ³	ft ³	2.83 · 10 ⁻²
°C	°F	5/9 (°F – 32)
kg	lb	0.454
J/kg	Btu/lb	1.33 · 10 ⁻⁴
m ³ /s	cfm	4.72 · 10 ⁻⁴
m ³ /s	gpm	6.31 · 10 ⁻⁵
J/kWh	Btu/kWh	1055.056
mills	\$	0.001
kg/m ²	in. Hg	345.31

CHAPTER 1

INTRODUCTION

Combustion of sulfur-containing fuels, such as coal and oil, results in sulfur dioxide (SO₂) formation. SO₂ emissions are known to cause detrimental impacts on human health and the environment. The major health concerns associated with exposure to high concentrations of SO₂ include breathing difficulties, respiratory illness, and aggravation of existing cardiovascular disease. In addition to the health impacts, SO₂ leads to acid deposition in the environment. This deposition causes acidification of lakes and streams and damage to tree foliage and agricultural crops. Furthermore, acid deposition accelerates the decay of buildings and monuments. While airborne, SO₂ and its particulate matter derivatives contribute visibility degradation.

Electric power generating units account for the majority of SO₂ emissions in the U.S. In 1998, these units contributed 64 percent of the national SO₂ emissions.¹ To mitigate SO₂ emissions from electric power generating units, the Acid Rain SO₂ Reduction Program² was established under Title IV of the Clean

Air Act Amendments of 1990 (CAAA). This two-phase program was designed to reduce SO₂ emissions from the power generating industry.

Phase I of the Acid Rain SO₂ Reduction Program began on January 1, 1995, and ended December 31, 1999. In 1997, 423 power generating units, affected under Phase I, emitted 5.4 million tons of SO₂ (1.7 million tons below the allowable 7.1 million tons of SO₂).³ Thus, the SO₂ emissions in 1997 reflect an output of 23 percent below the allowable amount.

Phase II of the Acid Rain SO₂ Reduction Program began on January 1, 2000. The nationwide cap for SO₂ will be 9.48 million tons from 2000 through 2009. In 2010, the cap will be reduced further to 8.95 million tons, a level approximately one-half of industry-wide emissions in 1980. To meet the requirements of this phase, some power generating units may use FGD technologies. Additionally, the use of these technologies can result in the reduction of fine particle precursor emissions and mercury emissions from combustion units. It is timely, therefore, to examine the current status of FGD (or SO₂ scrubbing) technologies.

This report presents a review of current FGD technologies. Following the introduction, Chapter 2 presents a concise review of commercially available FGD technologies. Technology applications on combustion units in the United States and abroad are discussed in Chapter 3. The performance and applicability of the most commonly occurring types of FGD technology installations is presented in Chapter 4. A review of recently reported technical advances to FGD technologies is provided in Chapter 5. Capital and operating costs of LSFO, LSD,

and MEL are analyzed in Chapter 6. Additional benefits achieved with wet limestone scrubbers and spray dryers are discussed in Chapter 7. References reviewed and utilized for the production of this report are given at the end.

It is expected that this review will be useful to a broad audience, including: (1) individuals responsible for developing and implementing SO₂ control strategies at sources, (2) persons involved in developing SO₂ and other regulations, (3) State regulatory authorities implementing SO₂ control programs, and (4) interested public at large. Moreover, persons engaged in research and development efforts aimed at improving cost-effectiveness of FGD technology may also benefit from this review.

CHAPTER 2

FGD TECHNOLOGY

Introduction

Various technologies exist that have been designed to remove SO₂ from flue gas produced by electricity generating plants. These technologies represent a varying degree of commercial readiness. Some can claim tens of thousand of hours of operational experience, while others have only recently been demonstrated at commercial plants. This report considers only commercially available FGD technologies that have an established record of reliable performance and sufficient quality and quantity of data to determine the cost of their deployment.

Commercially available FGD technologies can “conventionally” be classified as *once-through* and *regenerable*, depending on how sorbent is treated after it has sorbed SO₂.⁴ In once-through technologies, the SO₂ is permanently bound by the sorbent, which must be disposed of as a waste or utilized as a by-product (e.g., gypsum). In regenerable technologies, the SO₂ is released from the sorbent during the regeneration step and may be further processed to yield sulfuric acid, elemental sulfur, or liquid SO₂. The regenerated sorbent is recycled in the SO₂ scrubbing step. Both once-through and

regenerable technologies can be further classified as *wet* or *dry*. In wet processes, wet slurry waste or by-product is produced, and flue gas leaving the absorber is saturated with moisture. In dry processes, dry waste material is produced and flue gas leaving the absorber is not saturated with moisture.

Depending on process configuration and local market conditions at the plant site, once-through wet FGD processes can produce slurry waste or salable by-product. This waste/by-product must be dewatered in some fashion prior to disposal or sale (in case of a salable by-product). The “conventional” classification of FGD processes is shown in Figure 2-1.

A review of FGD technology applications was conducted based on the information provided in CoalPower3 Database, available from the International Energy Agency's Coal Research Centre in London, England. This database lists commercial FGD applications. The review reveals that regenerable FGD processes are being used only marginally, with once-through FGD processes involved in the vast majority of applications. Therefore, for this work, FGD technologies were grouped into the following three major categories:

- Wet FGD (composed of once-through wet FGD)
- Dry FGD (composed of once-through dry FGD)
- Regenerable FGD (composed of wet and dry regenerable FGD)

The above grouping of FGD technologies is consistent with other evaluations of FGD,⁵ and will be used in the remaining chapters of this report. Accordingly, when wet FGD is mentioned in the remainder of this report, it is

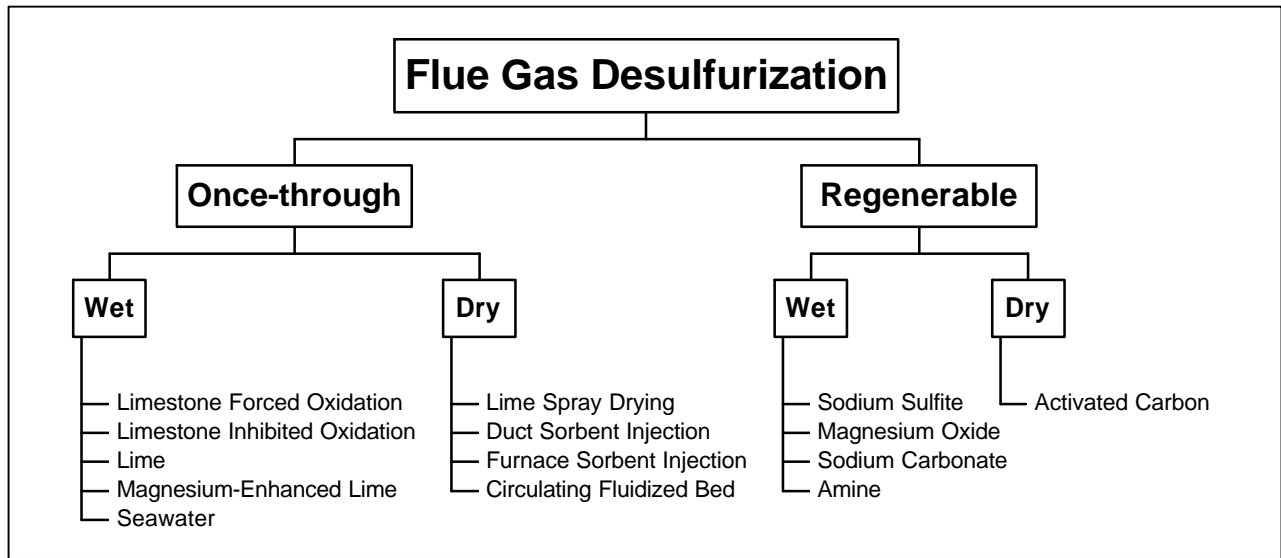


Figure 2-1. FGD technology tree.

meant as once-through wet FGD. Similarly, when dry FGD is mentioned, it is meant as once-through dry FGD. Moreover, as regenerable technologies are used only marginally, their coverage in this report is limited.

Wet FGD Technologies

In wet FGD processes flue gas contacts alkaline slurry in an absorber. The absorber may take various forms (spray tower or tray tower), depending on the manufacturer and desired process configuration. However, the most often-used absorber application is the counterflow vertically oriented spray tower. A diverse group of wet FGD processes have evolved to take advantage of particular properties of various sorbents and/or by-products. All wet FGD processes discussed here are once-through (i.e., non-regenerable). A generalized flow diagram of a baseline wet FGD system is shown in Figure 2-2. SO₂-containing flue gas is contacted with limestone slurry in an absorber. Limestone slurry is prepared in two consecutive steps.

First, limestone is crushed into a fine powder with a desired particle size distribution. This takes place in a crushing station; e.g., ball mill (fine crushing maximizes the dissolution rate of a given limestone). Next, this fine powder is mixed with water in a slurry preparation tank. Sorbent slurry from this tank is then pumped into the absorber reaction tank.

As mentioned before, the absorber is most often a counterflow tower with flue gas flowing upwards, while limestone slurry is sprayed downwards by an array of spray nozzles. In the absorber, SO₂ is removed by both sorption and reaction with the slurry. Reactions initiated in the absorber are completed in a reaction tank, which provides retention time for finely ground limestone particles to dissolve and to react with the dissolved SO₂.

FGD TECHNOLOGY

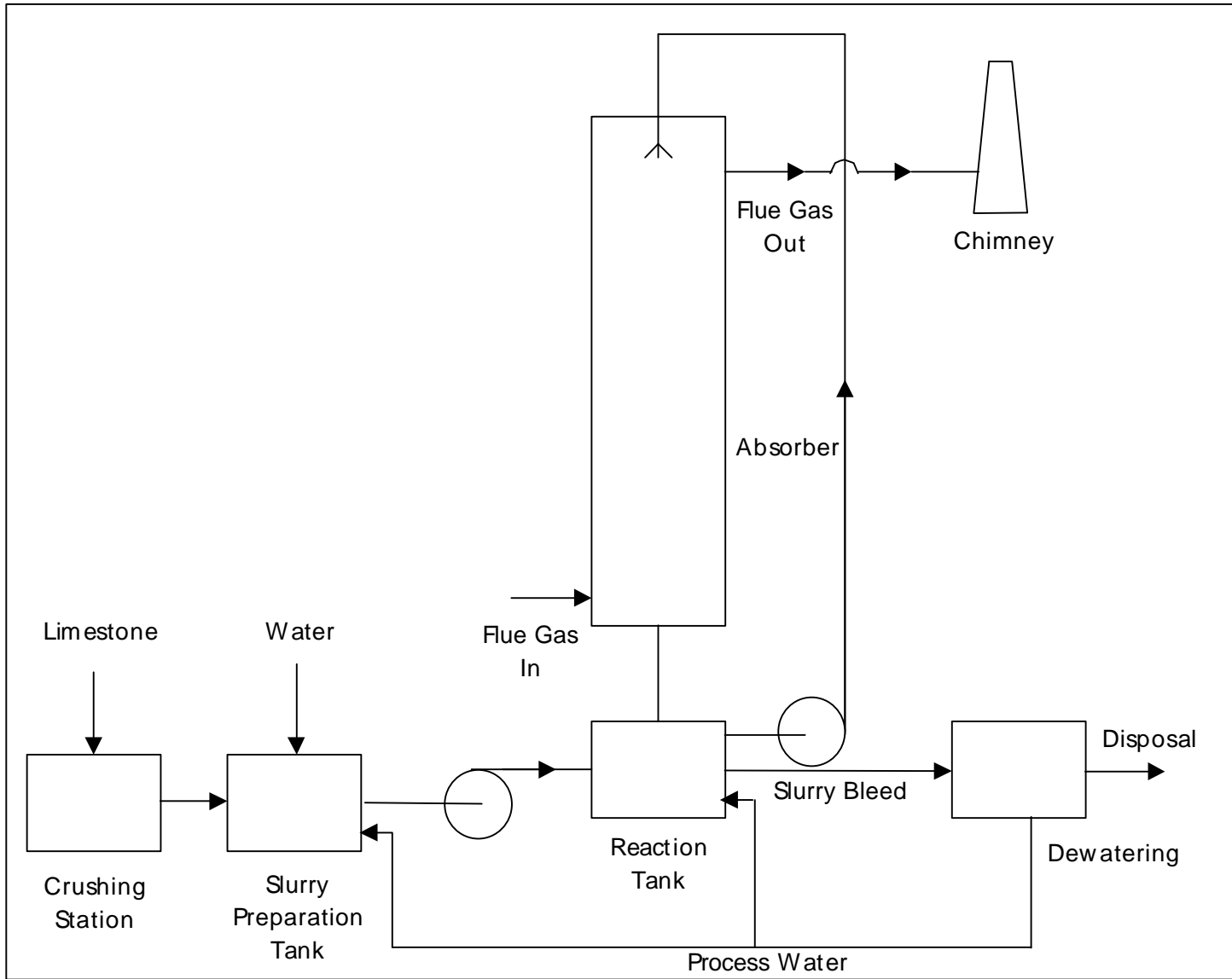
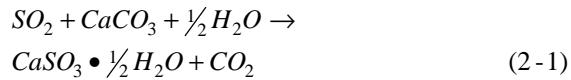
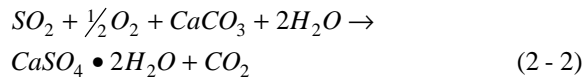


Figure 2-2. Baseline wet FGD system.

The overall reactions in the absorber and in the reaction tank can be summarized by:



and



The complex chemistry summarized by the above equations involves SO₂-CO₂-H₂O equilibrium relationships in the absorber, limestone dissolution, and sulfite/sulfate crystallization (occurring mostly in the reaction tank)⁶. If the oxidation of sulfite to sulfate is not controlled, the wet limestone system is operating under the so-called natural oxidation. Depending on SO₂ concentration and the excess air in the flue gas, as well as on slurry pH, some systems may be operated in the natural oxidation mode. However, for most applications, it is beneficial to control oxidation.

The dissolution and crystallization reactions in the reaction tank are, to a large extent, controlled by the pH of the liquid, which is a function of limestone stoichiometry (number of mols of Ca added per mol of SO₂ removed). Both pH and limestone stoichiometry are preset parameters for the operation of an absorber. Normally, the required stoichiometry of a limestone wet FGD system varies from 1.01 to 1.1 moles of CaCO₃ per mole of SO₂ (1.01 to 1.05 for modern scrubbers) and pH is in the range 5.0 to 6.0. A gradual decrease in a preset operating value of pH indicates increased limestone consumption and triggers the fresh limestone feed. Spent sorbent from the reaction tank (slurry bleed) is dewatered and

disposed of in a waste slurry pond (ponding). The complexity of the dewatering process is determined by the chemical composition and crystal habit of the spent sorbent, and whether the end product is to be utilized or discharged. For example, CaSO₄ is easier to dewater than CaSO₃.

Entrained slurry droplets that escaped from the absorber's spray zone and were carried out by the flue gas are separated in an impaction-type mist eliminator. Mist eliminator design parameters include style (chevrons, mesh pads, baffles, etc.), blade number and spacing, and wash system configuration. The mist eliminator plays an important role in preventing corrosion of downstream equipment and ducts, as well as deposition of stack effluent in the immediate vicinity of the plant. Mist eliminators can be designed for either a vertical or horizontal configuration. A horizontal configuration offers several advantages over a vertical configuration; e.g., better drainage. However, the drawbacks of horizontal mist eliminators include increased flue gas pressure drop and more difficult maintenance.

Wet FGD process variables include: flue gas flow rate, liquid-to-gas ratio (L/G), recycle slurry pH, flue gas SO₂ concentration, and solids concentration and retention time. The effect of these variables on the operation of a wet FGD system is discussed below.

Flue gas velocity optimization considerations depend on the type of wet absorber used. Normally, the upper limit for flue gas velocity in a counterflow absorber depends on the capability of the mist eliminators to prevent droplet carryover.⁷ Droplet carryover, or droplets escaping from the unit eliminator, can increase duct corrosion downstream of the absorber. Some absorbers

have a perforated tray added for the improvement of SO₂ capture. In such cases, the optimum flue gas velocity is determined by the tray design. For this type of absorber, excessive flue gas velocity will cause an absorber to "flood," whereas too low a velocity will prevent slurry holdup on the tray. For a given scrubber, trays are designed for a maximum gas velocity, so as not to flood.

Another type of wet FGD absorber that could be used for SO₂ control is a packed absorber. Packed absorber utilizes a material placed in it to provide a surface over which scrubbing solution is distributed. In this manner, gas/liquid contact surface area is generated. As far as a mist eliminator's operation is concerned, higher flue gas velocities could be used for a packed absorber without causing its failure and a subsequent droplet carryover. Packed absorbers can be used only for clear solution systems (systems with a scrubbing medium being a solution rather than a slurry).

L/G is usually expressed in terms of gallons of slurry per 1000 ft³ of flue gas at actual conditions leaving the absorber. The amount of surface system available for the reaction with SO₂ is determined by L/G. For a counterflow spray absorber operated at a given flue gas flow rate, L/G approximates the surface area of droplets and is one of the main design variables available to obtain a desired SO₂ removal in the absorber. The amount of available alkalinity for the reaction with SO₂ increases with the increasing L/G. L/G also affects the oxidation rate of sulfite/bisulfite reaction products in the absorber by affecting the absorption rate of O₂ from the flue gas. As will be explained further in this report, oxidation rate affects the potential for scaling absorber internals.

Slurry pH also has a significant effect on SO₂ removal efficiency in a wet FGD system. In addition, pH is likely the single most important control variable for absorber operation. It determines the amount of limestone added to the system. Within the operational range, increasing the amount of limestone added increases the amount of SO₂ removal. This is because of the increased concentration of soluble alkaline species and undissolved reagent. This reagent is then available for dissolution and renewal of alkalinity in the liquid phase.

At constant operating conditions of a scrubber, increasing the concentration of SO₂ (increasing sulfur content of fuel) will decrease SO₂ removal efficiency by a wet absorber. This decreased efficiency is observed because increasing SO₂ concentration causes a more rapid depletion of liquid phase alkalinity causing the increase of liquid phase resistance.

Solids concentration and retention time affect the reliability of wet FGD operation. Solids concentration in the slurry is typically maintained at 10 to 15 percent solids by weight. It is controlled by removing a part of the slurry from the reaction tank for subsequent dewatering. Proper solids concentration in the slurry is necessary to ensure scale-free operation of the absorber. Correct solids retention time in the reaction tank is essential to achieving high utilization of limestone and maintaining correct handling and dewatering properties of solids. Typical solids retention time for wet FGD is 12 to 14 hours.⁷

Limestone Forced Oxidation

As described above, wet FGD can be operated reliably in a natural oxidation mode under certain favorable conditions. However,

for the majority of applications, it is necessary to control the extent of oxidation in order to improve operational reliability of the system. Over the years, several process variations have been designed to improve the operational reliability of wet FGD technology. Consequently, the *limestone forced oxidation process* (LSFO) has become the preferred FGD technology worldwide.

First-generation wet limestone FGD systems were plagued with scaling problems, resulting from oxidation of the reaction products to calcium sulfate (gypsum) that would deposit throughout the absorber, mist eliminator, and piping. Gypsum scale typically forms via natural oxidation when the fraction of calcium sulfate in the slurry (slurry oxidation level) is greater than 15 percent. Initially, gypsum scaling was combated by installation of extra capacity.

One way to prevent the scaling problem is to blow air into the absorbent slurry to encourage controlled oxidation outside of the absorber. This type of FGD system, limestone forced oxidation, provides rapid calcium sulfate crystal growth on seed crystals. LSFO minimizes scaling in the scrubber and also results in slurry that can be more easily dewatered. Consequently, the LSFO system has become the preferred technology worldwide. The most often used configuration is for the air to be blown into the reaction tank (*in-situ* oxidation). Alternatively, air can be blown into an additional hold tank (*ex-situ* oxidation). LSFO requires compressors/ blowers and additional piping, compared to a system without forced oxidation.

The prime benefit of scale control derived from forced oxidation is greater scrubber absorber availability. As a result, the need

for redundant capacity is greatly reduced. The added benefits are the formation of a stable product, a salable by-product (which eliminates the need for landfilling), and smaller dewatering equipment. Nearly complete (99 percent plus) oxidation is required for a commercial quality by-product.

This level of oxidation can be accomplished in a modern wet FGD system. However, the salability of the wet FGD by-product (FGD gypsum) is also a function of the demand for gypsum. Depending on site-specific conditions, LSFO may produce a salable by-product in the form of commercial grade gypsum that could be used for wallboard manufacturing. When salable gypsum is not attainable, dry FGD waste is piled (gypsum stacking) or landfilled. Gypsum stacking is the procedure where a gypsum slurry is sent to the stacking area, allowed to have the solids to separate from the water, and then removing the water and leaving the solids as a pile.

The solids handling system for LSFO consists of primary and secondary dewatering, solids modification unit, and ultimate waste disposal, regardless whether a part or all of the by-product will be sold as commercial quality gypsum. The objective of primary dewatering is to increase the solids concentration of spent limestone slurry from the reaction tank discharge conditions (10 to 15 percent by weight) to between 30 and 50 percent by weight. Primary dewatering is accomplished by hydroclones. The process water recovered during primary dewatering is recirculated to the absorber. Solids discharged from the primary dewatering unit are directed to the underflow storage tank.

The objective of secondary dewatering is to reduce the moisture content (increase solids

content) beyond the setpoint of primary dewatering. The solids content of the material leaving this stage will be 45 to 90 percent. This relatively wide range of solids concentration in the product of secondary dewatering is a result of different disposal methods for the product. For an LSFO absorber aimed at commercial gypsum production, solids concentration in the product will be in the high end of the range. However, for an absorber operated as LSFO, but without product recovery, the solids concentration will be at the low end of the range.

The types of equipment most often used for secondary dewatering are belt and/or drum vacuum filters and centrifuges. The selection of the equipment depends on the quality of product desired. If commercial quality gypsum is desired, then belt vacuum filters may be selected over drum filters because of their ability to provide superior cake washing capabilities (important to achieve gypsum specifications). The process water recovered during secondary dewatering is recirculated to the absorber.

Solids discharged from the secondary dewatering unit are directed either to the modification unit of solids handling or to the temporary storage system. During the modification, solids are stabilized or fixated to improve their strength bearing, landfill, and leachate characteristics. This is most often accomplished by mixing dewatered solids with fly ash and lime in a pug mill to promote the pozzolanic reaction. Pozzolanic reaction occurs when lime and silica react in the presence of water to form hydrated calcium silicates. The degree of solids modification is dependent on the final use for the solids (e.g., road-base, concrete aggregate, or structural fill). By-product

solids can be used as a road-base, concrete aggregate, or structural fill. These applications utilize improved properties of FGD by-product mixed with fly ash: increased unconfined strength and decreased permeability. These improved properties are the result of pozzolanic reaction. Sometimes, when commercial quality gypsum is made, pelletization is employed. The selection of the ultimate disposal method is highly site-specific and depends on, among other factors, land availability, hydrogeology, and topography. In general, three options exist for the ultimate disposal of waste FGD solids: landfills, ponds, and gypsum stacks.

In addition to technical issues, several market issues are involved in the decision of wallboard manufacturers to use FGD gypsum. These market issues are presented below. Normally, the use of the quantity of FGD gypsum produced by a representative LSFO (hundreds of thousands of tons per year) would be possible only if a dedicated wallboard plant was built for this feed source, or was shared by several existing wallboard plants.⁸ The proximity of the wallboard plant to the FGD by-product plant is important because the transportation cost of FGD gypsum to the wallboard plant can be a significant percentage of its market value. Since most existing wallboard plants in the United States were designed to use mined rock gypsum as feed material, the solids handling equipment at these plants can use only a limited quantity of FGD gypsum, which has different handling properties.

Another potential obstacle to the marketability of FGD gypsum is the fact that the operating schedule of a power plant and that of a wallboard plant often do not coincide. Wallboard plants generally have storage capacity to buffer the flow of gypsum

in and out of the plant. Unlike power plants, wallboard plants do not operate 24 hours per day and 7 days a week. Similarly, power plants do not operate year round, whereas wallboard plants do.⁷

Limestone Inhibited Oxidation

A variation of the wet limestone process is the *limestone inhibited oxidation process* (LSIO). This process has been designed to control oxidation in the absorber. The LSIO is particularly well suited for applications with high sulfur coals. Because of LSIO chemistry, the difficulty in inhibiting the oxidation generally increases with the decreasing amount of sulfur content in coal.⁹

Several factors influence the performance of LSIO. Flue gas composition, most notably oxygen concentration, affects the extent of sulfite oxidation to sulfate. Other flue gas factors affecting LSIO are: SO₂ concentration, fly ash content in the inlet gas to the scrubber, and flue gas temperature and humidity. Changing mass transfer characteristics of the system (the ratio of SO₂/O₂ absorbed) can alter the extent of natural oxidation and, therefore, determine how difficult it will be to inhibit the oxidation. The change in mass transfer characteristics of the system can result from adjusting the L/G. Chemical characteristics of the system, such as pH and liquid-phase composition, can also alter the difficulty of oxidation inhibition.

In the LSIO, emulsified sodium thiosulfate (Na₂S₂O₃) is added to the limestone slurry feed to prevent the oxidation to gypsum in the absorber's internals by lowering the slurry oxidation ratio to below 15 percent.¹⁰ Typically, a design oxidation ratio of between 4 and 10 percent is used in LSIO. The amount of additive necessary to inhibit oxidation depends on the chemistry and

operating conditions of a given absorber and is, therefore, site specific.

Because of economic considerations, sulfur is often added to the limestone slurry in lieu of thiosulfate. Sulfur is added directly to the limestone reagent tank. However, conversion to thiosulfate occurs in the reaction tank when sulfur contacts sulfite. The overall conversion of sulfur to thiosulfate is between 50 and 75 percent. The amount of thiosulfate (or sulfur) required to achieve inhibited oxidation is a function of system chemistry and operating conditions.

An additional benefit of using LSIO may be an increased limestone solubility, which enhances sorbent utilization. The waste product, calcium sulfite, is landfilled. The dewatering characteristics of the waste are improved for LSIO compared to the waste from natural oxidation operation of a wet FGD absorber. This is because the calcium sulfite product from the LSIO tends to form larger crystals, similar to gypsum solids.

Lime and Magnesium-Lime

The *lime process* uses hydrated calcitic lime slurry in a countercurrent spray tower. This slurry is more reactive than limestone slurry, but is more expensive. The *magnesium-enhanced lime process* (MEL) is a variation of the lime process in that it uses a special type of lime, magnesium-enhanced lime (typically 5 – 8 percent magnesium oxide) or dolomitic lime (typically 20 percent magnesium oxide).¹¹ The operational pH value for lime processes is normally in the range 6.0 to 7.0 because of their increased alkalinity and solubility, compared to limestone processes. The lime process may be designed to utilize the alkalinity of fly ash in addition to the alkalinity of a sorbent.

Lime used in the MEL contains magnesium in addition to its calcitic component. Because of the greater solubility of magnesium salts compared to calcitic sorbents, the scrubbing liquor is significantly more alkaline. Therefore, MEL is able to achieve high SO₂ removal efficiencies in significantly smaller absorber towers than the limestone scrubbers. Additionally, MEL allows for a significant decrease of L/G, compared to LSFO for a given target SO₂ removal.¹²

Because waste solids from MEL have poorer dewatering characteristics than solids from calcitic limestone slurry processes, the best dewatering operation of MEL occurs when low solids concentration is maintained along with moderate-to-low sulfite oxidation levels.¹³ Forced oxidation, external to the absorber, can be used in MEL to improve the quality of their solids. This results in the production of commercial quality gypsum.⁷ Commercial grade gypsum produced from MEL is, in fact, brighter than gypsum produced by a conventional LSFO. Brighter gypsum, potentially, has a higher commercial value.¹⁴

Seawater Process

The seawater process utilizes the natural alkalinity of seawater to neutralize SO₂. The chemistry of the process is similar to the LSFO chemistry except that the limestone comes completely dissolved with the seawater and that the chemistry does not involve any dissolution or precipitation of solids. Seawater is available in large amounts at the power plant as cooling medium in the condensers. It is used as a sorbent downstream of the condensers for the purpose of FGD. Seawater is alkaline by nature, and has a large neutralizing capacity with respect to SO₂.

The absorption of SO₂ takes place in an absorber, where seawater and flue gas are brought into close contact in a counter-current flow. The scrubber effluent flows to the treatment plant where it is air-sparged to oxidize absorbed SO₂ into sulfate before discharge.¹⁵ The sulfate is completely dissolved in seawater, so as a result there is no waste product to dispose of. Sulfate is a natural ingredient in seawater, and typically there is only a slight increase of sulfate in the discharge. This increase is within variations naturally occurring in seawater. The difference from the background level normally is not detectable within even a short distance from the point of discharge.

Since the utilization of seawater for SO₂ scrubbing introduces a discharge to the ocean, it is necessary to make an assessment based on local conditions. Typically, the assessment includes: effluent dilution and dispersion calculations, description of effluent, comparison of effluent data with local quality criteria, description of local marine environment, and evaluation of possible effects from the discharge. High chloride concentrations, characteristic of systems using seawater, result in a requirement for construction materials with increased corrosion resistance.¹⁶

Dry FGD Technologies

In these technologies, SO₂-containing flue gas contacts alkaline (most often lime) sorbent. As a result, dry waste is produced with handling properties similar to fly ash. The sorbent can be delivered to flue gas in an aqueous slurry form [*lime spray drying process* (LSD)] or as a dry powder [*duct sorbent injection process* (DSI), *furnace*

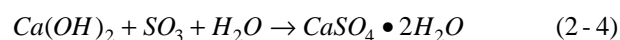
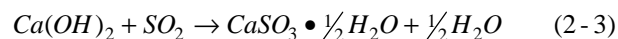
sorbent injection process (FSI), and circulating fluidized bed process (CFB)]. The LSD and the CFB require dedicated absorber vessels for sorbent to react with SO₂, while in DSI and FSI new hardware requirements are limited to sorbent delivery equipment. In dry processes, sorbent recirculation may be used to increase its utilization. All dry FGD processes discussed here are once-through (i.e., non-regenerable) and, in general, limited to SO₂ removals below those attainable with wet once-through FGD.

Lime Spray Drying

LSD for the control of SO₂ emissions is used for sources that burn low- to medium-sulfur coal, with occasional applications for coals with higher sulfur content. Some issues that limit the use of spray dryers with high-sulfur coals include the potential impact of chloride contained in the coal on the spray dryer performance, and the ability of the existing particulate control device to handle the increased loading and achieve the required efficiency.

The LSD is shown schematically in Figure 2-3. Hot flue gas mixes in a spray dryer vessel with a mist of finely atomized fresh lime slurry. Fresh lime slurry is prepared in a slaker (most often a ball mill) at a nominal concentration of solids. Rotary atomizers or two-fluid nozzles are used to finely disperse lime slurry into flue gas. Typically, spray dryers are operated at lime stoichiometry of 0.9 for low sulfur coals and 1.3 to 1.5 for high sulfur coals. Simultaneous heat and mass transfer between alkali in a finely dispersed lime slurry and SO₂ from the gas phase result in a series of reactions and a drying of process waste. The amount of water fed into the spray dryer is carefully controlled to avoid complete saturation of the flue gas.

While a close approach to adiabatic saturation (from 10 to 15 °C for coal-derived flue gas) is required to achieve high SO₂ removal, complete saturation impairs operation of a spray dryer because of wet solids adhering to vessel walls and within the particulate collector. Primary reactions in the spray dryer are as follows:



Some of the dry reaction product solids are collected at the bottom of the spray dryer. The remaining solids, suspended in the flue gas, travel to the particulate control device where the separation occurs. For a process configuration where the particulate control device is a baghouse, a significant additional SO₂ removal may occur in the filter cake on the surface of bags. Dry solids from the particulate control device's hopper and from the bottom of the spray dryer are disposed of.

The extent of alkali usage in a spray dryer is limited by its available residence time for a gas-solid reaction. Typical residence time in a spray dryer is 8 to 12 seconds. In order to increase sorbent utilization, part of the dry solids from the bottom of the spray dryer and the particulate collector's hopper are sent to the recycle solids slurry tank. The recirculated stream (shown with a broken line in Figure 2-3) contains partially reacted alkali from previous passes through the system. The additional exposure of a sorbent to SO₂ afforded by the recycle promotes increased sorbent utilization.

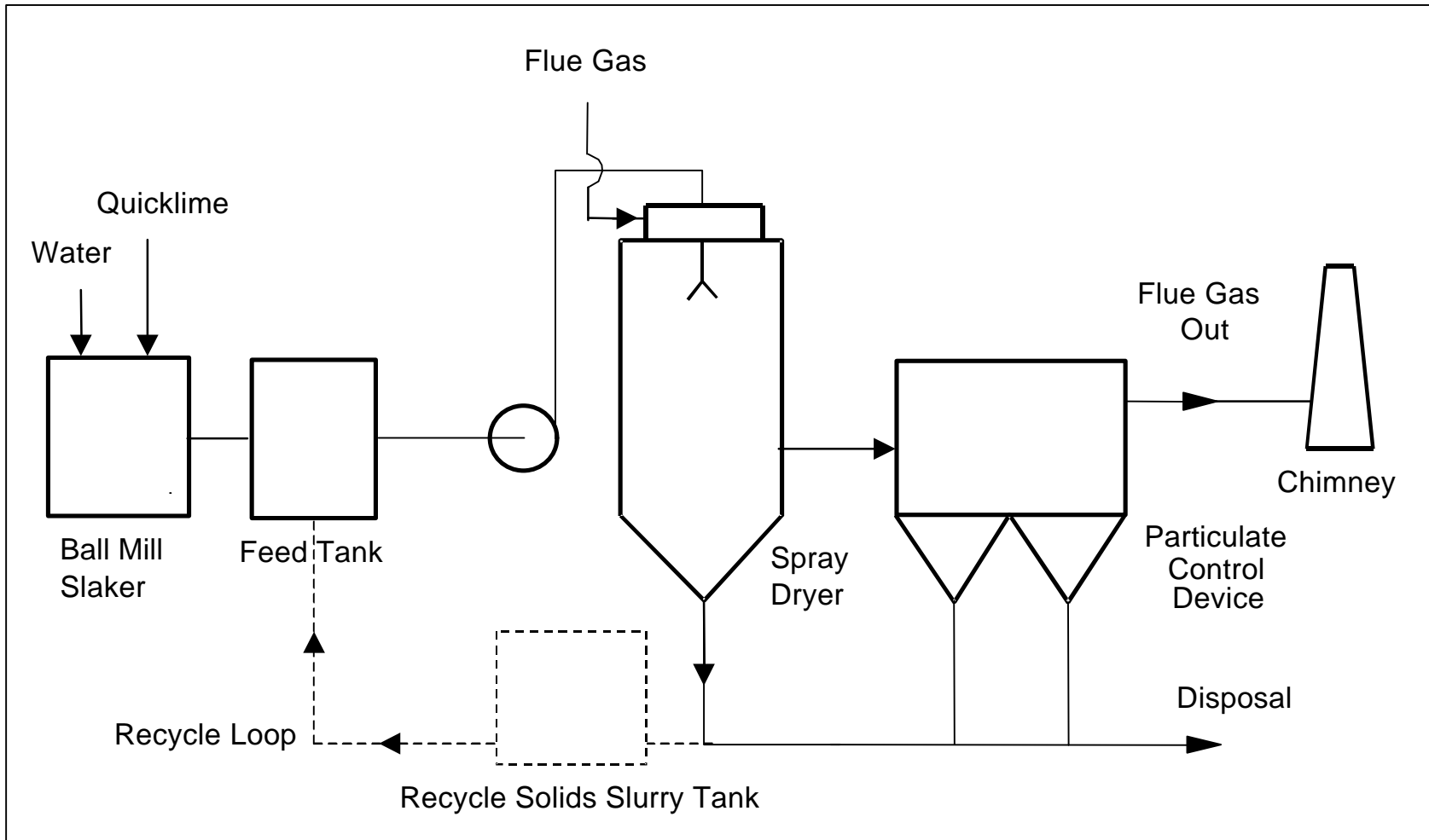


Figure 2-3. Lime spray dryer FGD system.

Mass transfer during a spray drying process occurs in two discrete phases: moist and dry.¹⁷ During the moist phase, SO₂ diffuses from the bulk gas to the moisture layer on the surface of lime particles and reacts with dissolved lime. The reaction product precipitates on the surface of the lime particle. During the dry phase, SO₂ diffuses through the products of the lime and SO₂ reaction and causes a gas-solid reaction with the unreacted core of lime particle.

Studies indicated that a majority of SO₂ capture in the spray dryer occurs during the moist phase. Any increase in the duration of the moist phase would therefore increase the amount of captured SO₂. Deliquescent salt additives sometimes are added to the lime slurry to be atomized in a spray dryer to achieve this effect. A similar effect is achieved when spray dryers are used on coals with elevated chloride content.

Duct Sorbent Injection

DSI for SO₂ emission control is intended to enable the control directly in the flue gas duct between the air preheater and the particulate control device. Since no dedicated absorber vessel is required, the amount of hardware needed to control SO₂ is minimized for DSI.

DSI utilizes the contacting of finely dispersed sorbent with the flue gas. Sorbent used in DSI is typically hydrated lime or, occasionally, sodium bicarbonate.¹⁸ In the DSI shown schematically in Figure 2-4, dry hydrated lime sorbent is injected into the flue gas downstream of the boiler's air preheater. Water may be injected separately from the sorbent either downstream or upstream of the dry sorbent injection point to humidify the flue gas. The relative position of dry sorbent and water injection is optimized to maximally promote the so-called

droplet scavenging or impacts between sorbent particles and water droplets, both suspended in the flue gas. Fly ash, reaction products, and any unreacted sorbent are collected in the particulate control device. Additionally, recycling solids from the particulate control device can boost the utilization of alkaline material.¹⁹

A variation of DSI is duct spray drying process, in which slurry is atomized and, subsequently, evaporated in the duct.

Furnace Sorbent Injection

In the FSI, a dry sorbent is injected directly into the furnace in the optimum temperature region above the flame.²⁰ FSI is shown schematically in Figure 2-5. As a result of the high temperature (approximately 1000 °C), sorbent particles (most often calcium hydroxide, but sometimes calcium carbonate) decompose and become porous solids with high surface systems,²¹ according to the reaction below:



SO₂ in the flue gas reacts with the nascent CaO as given below:



Calcium sulfate, and any remaining unreacted sorbent, leave the furnace with the flue gas. In some systems, the flue gas is humidified downstream of the air preheater or a humidifier vessel is installed to improve reagent utilization. *Ex-situ* spent sorbent reactivation (wetting) is also used occasionally as an integral part of the FSI. Sorbent reactivated *ex-situ* is then injected downstream of the air preheater. Such a configuration should probably be considered as a furnace/duct injection hybrid.

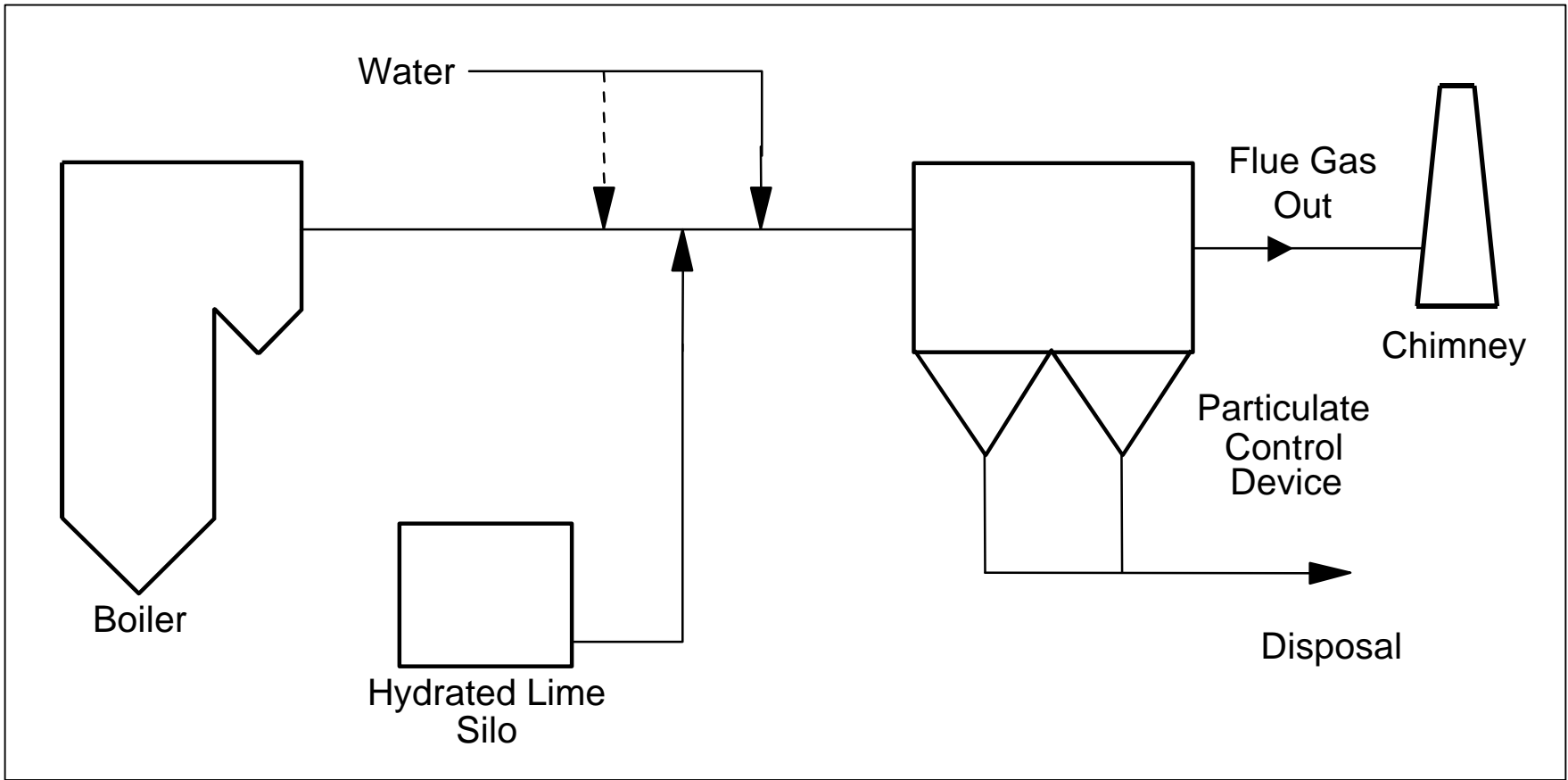


Figure 2-4. Schematic of DSI.

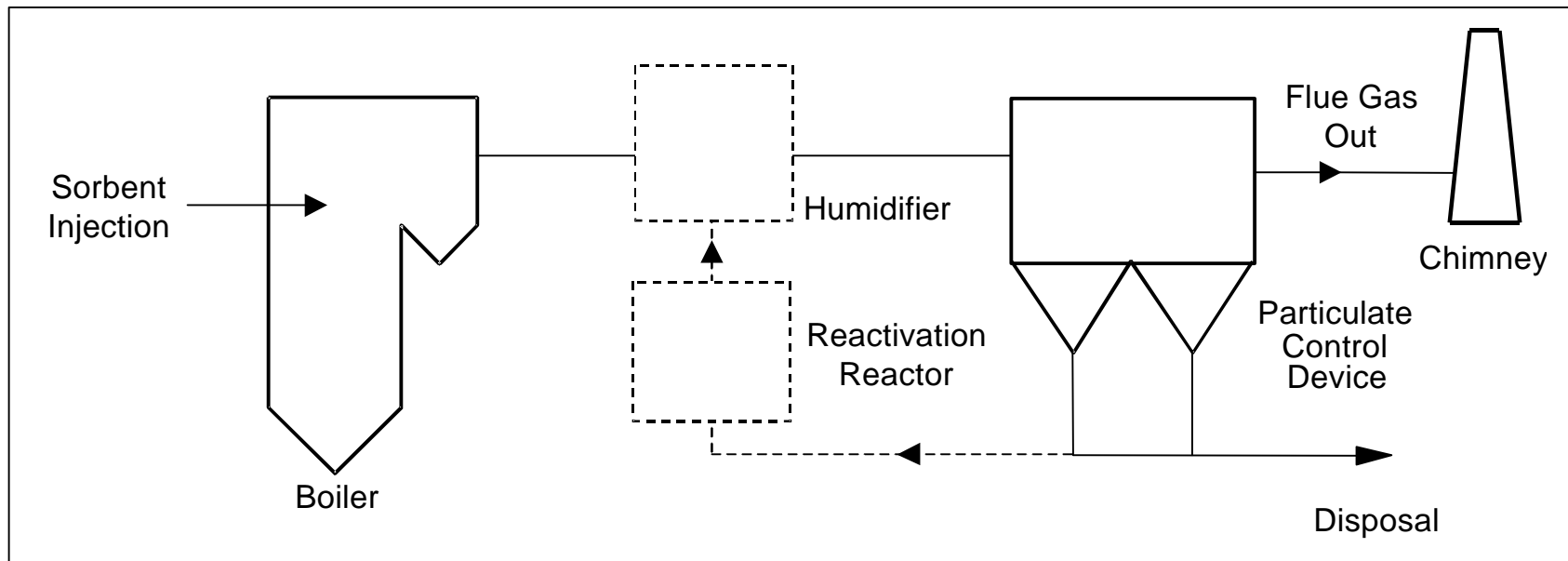


Figure 2-5. Schematic of FSI.

Circulating Fluidized Bed

In CFB, dry sorbent [most often $\text{Ca}(\text{OH})_2$] is contacted with a humidified flue gas in a circulating fluidized bed. CFB is shown schematically in Figure 2-6. The fluidized bed is formed as a result of flue gas flowing upward through a bed of sorbent solids. The CFB provides a long contact time between the sorbent and flue gas because sorbent passes through the bed several times. The flue gas laden with reaction products then flows to a particulate control device. Some of the particulate control device's catch is recirculated into the bed to increase the utilization of sorbent, while the remaining fraction is sent to disposal.

The CFB is characterized by good SO_2 mass transfer conditions from the gas to the solid phase. This is achieved as a result of intimate mixing of the solids with the gas as well as a high slip velocity between the two phases. An additional benefit of the fluidized bed is continuous abrasion of sorbent particles, resulting in the exposure of fresh, unreacted alkali.²²

The CFB is not widely used in the United States, and the bulk of its operating experience comes from Germany for units ranging from 50 to 250 MW_e .²³ This process uses hydrated lime rather than the less expensive and less reactive limestone commonly used in wet FGD technology processes. Additionally, due to a higher particulate matter concentration downstream of the fluidized bed, a larger ESP (or an additional precollector) may be needed to maintain the required particulate emission levels compared with a non-circulating sorbent.

Regenerable FGD Technologies

Regenerable FGD technologies discussed in this section include four wet regenerable processes (sodium sulfite, magnesium oxide, sodium carbonate, and amine) and one dry regenerable process (activated carbon). These processes are characterized by their product, a concentrated stream of SO_2 . As will be discussed in the following section, regenerable FGD technology finds only marginal application in the United States and throughout the world. These processes have a comparatively high O&M cost relative to other FGD processes, and the return from sale of the product does not offset a significant portion of the increased process cost. Product marketability may be a major problem.²⁴ As a result, some of the existing regenerable FGD-technology-equipped units have been converted to advanced limestone wet FGD.²⁵

Wet Regenerable FGD

Sodium Sulfite

The sodium sulfite, or Wellman-Lord process, absorbs SO_2 in a wet scrubber where pretreated flue gas is contacted with sodium sulfite solution. The product of the reaction is sodium bisulfite liquor heavily loaded with SO_2 . The liquor is subsequently regenerated in evaporators that crystallize sodium sulfite. Concentrated SO_2 is suitable for sulfuric acid production.

Magnesium Oxide

In the magnesium oxide process, SO_2 is removed in a wet scrubber. In this process, hydrogen chloride and hydrogen fluoride are removed in a prescrubber. The magnesium sulfite/sulfate product results from SO_2 absorption in a scrubber. The absorbed product is dried and calcined in a kiln to regenerate magnesium oxide. SO_2 captured during calcination is suitable for sulfuric acid production.

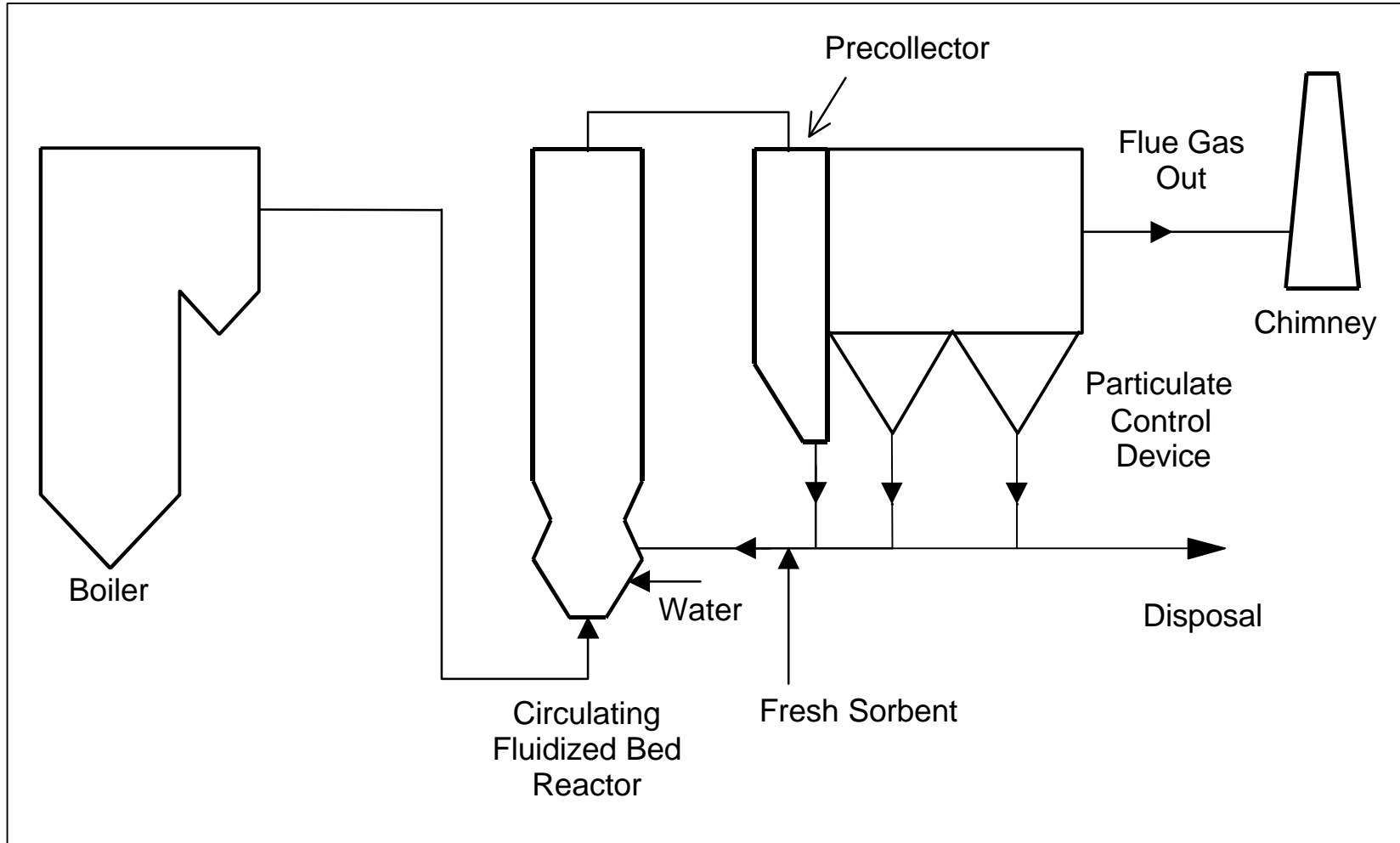


Figure 2-6. Schematic of CFB.

Sodium Carbonate

In this process, SO_2 is contacted with a spray of sodium carbonate solution. Products of the reaction are sodium sulfite and sodium sulfate, which are reduced to sodium sulfide. Following the reaction of sodium sulfide with carbon dioxide and water, sodium carbonate is regenerated and hydrogen sulfide is converted to sulfur.²⁶

Amine

The amine process involves absorption of SO_2 with an aqueous amine absorbent. The amine is regenerated thermally to release a concentrated water-saturated SO_2 stream. SO_2 may then be treated by conventional technologies to produce sulfuric acid.

Dry Regenerable FGD

Activated Carbon

The activated carbon process adsorbs SO_2 on a moving bed of granular activated carbon. Activated carbon is thermally regenerated to produce a concentrated SO_2 stream. SO_2 may then be treated by conventional technologies to produce sulfuric acid.

CHAPTER 3

TECHNOLOGY APPLICATIONS

Introduction

As discussed before, FGD technology applications were reviewed based on the information in CoalPower3 Database, available from the International Energy Agency's Coal Research Centre in London, England and released in November 1998.²⁷ This database has not been modified or otherwise amended. Findings of this review are described below.

Historical Applications

Applications of FGD technologies over the last three decades are shown in Figures 3-1 and 3-2 for the United States and the world, respectively. In the United States, wet FGD technology has dominated throughout the '70s and early '80s with over 90 percent of the overall installed FGD capacity. This same period also saw a considerably high rate of FGD installation: approximately 25,000 MW_e from 1976 through 1980. The mid-to-late '80s saw a lower rate of FGD capacity increase, compared to that of the '70s. It was in the '80s that the first dry and regenerable systems were installed. The early '90s saw a slow increase of installed FGD capacity, in wet and dry FGD technologies. A significant

increase of the FGD capacity occurred from 1994 through 1998. During this period, as much as a 20,000 MW_e increase was accomplished, almost all of it in wet FGD. No significant increase in regenerable FGD capacity has taken place since the early '80s.

A somewhat different pattern for the rate of application of FGD technology could be observed throughout the world, as shown in Figure 3-2. With approximately 30,000 MW_e of installed FGD capacity in 1980, the capacity has been increasing at an approximate rate of 100,000 MW_e per decade. Similar to the trend in the United States, no significant increase in regenerable FGD capacity has taken place worldwide since the early '80s. Also, the rest of the world has seen a smaller percent of dry FGD-controlled capacity than the United States.

Since the wet FGD technology has historically dominated both U.S. and worldwide applications, it is of interest to analyze application data in terms of specific wet FGD processes. An illustration of U.S. applications is presented in Figure 3-3. The initial installed FGD capacity in the early '70s was dominated by limestone processes.

Shortly thereafter, lime processes (lime and MEL) were applied. The sodium carbonate process was first applied in late '70s, and this application has not seen any significant growth through 1998. The growth of FGD during the mid-to-late '80s, as well as the early '90s, was almost entirely due to the increase of the wet limestone process capacity. From 1994 through 1998, there was a step increase in the installed FGD capacity with most of this being attributed to wet limestone processes and the dolomitic lime process in the United States.

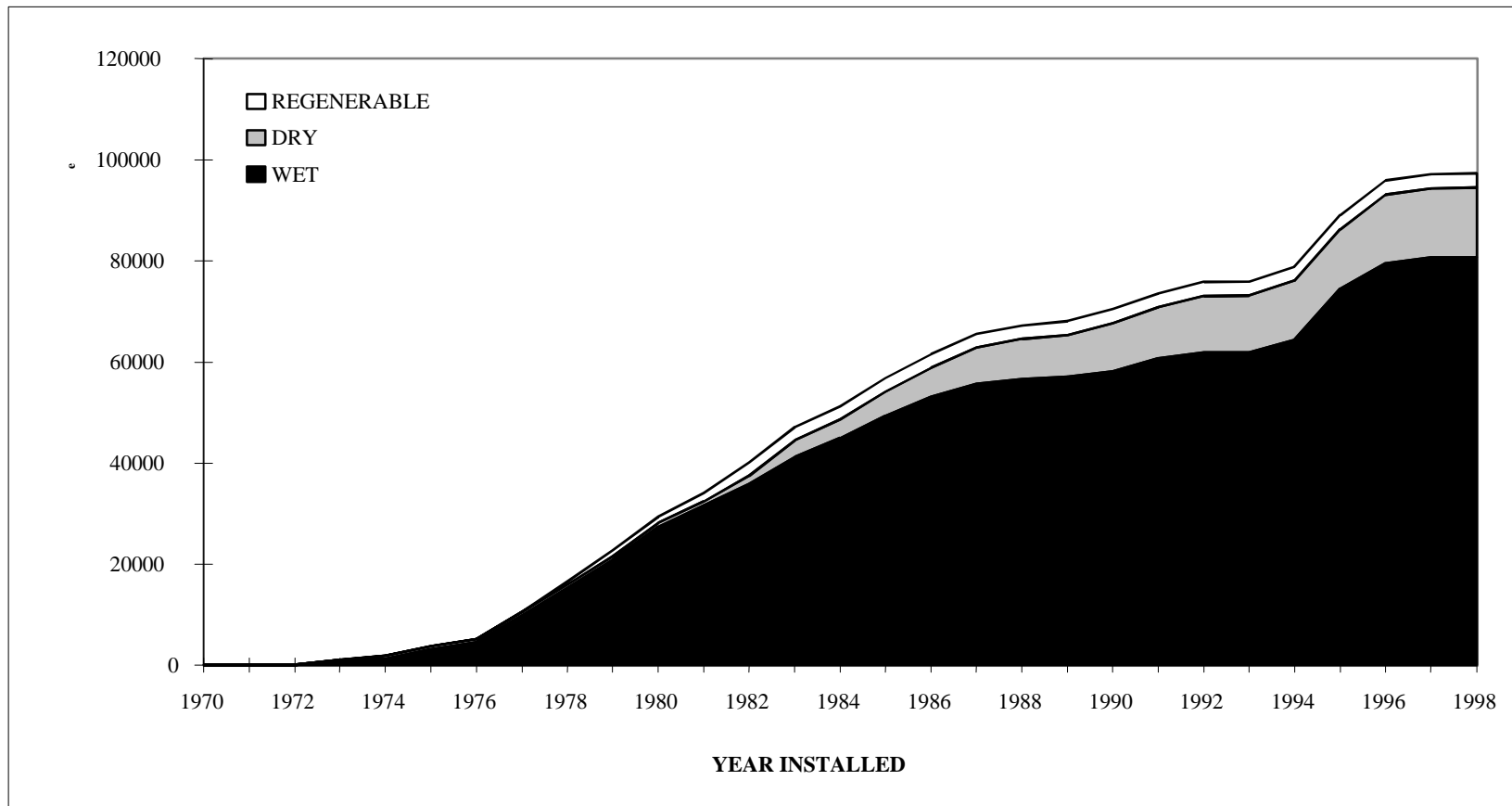


Figure 3-1. Historical application of FGD technology in the United States.

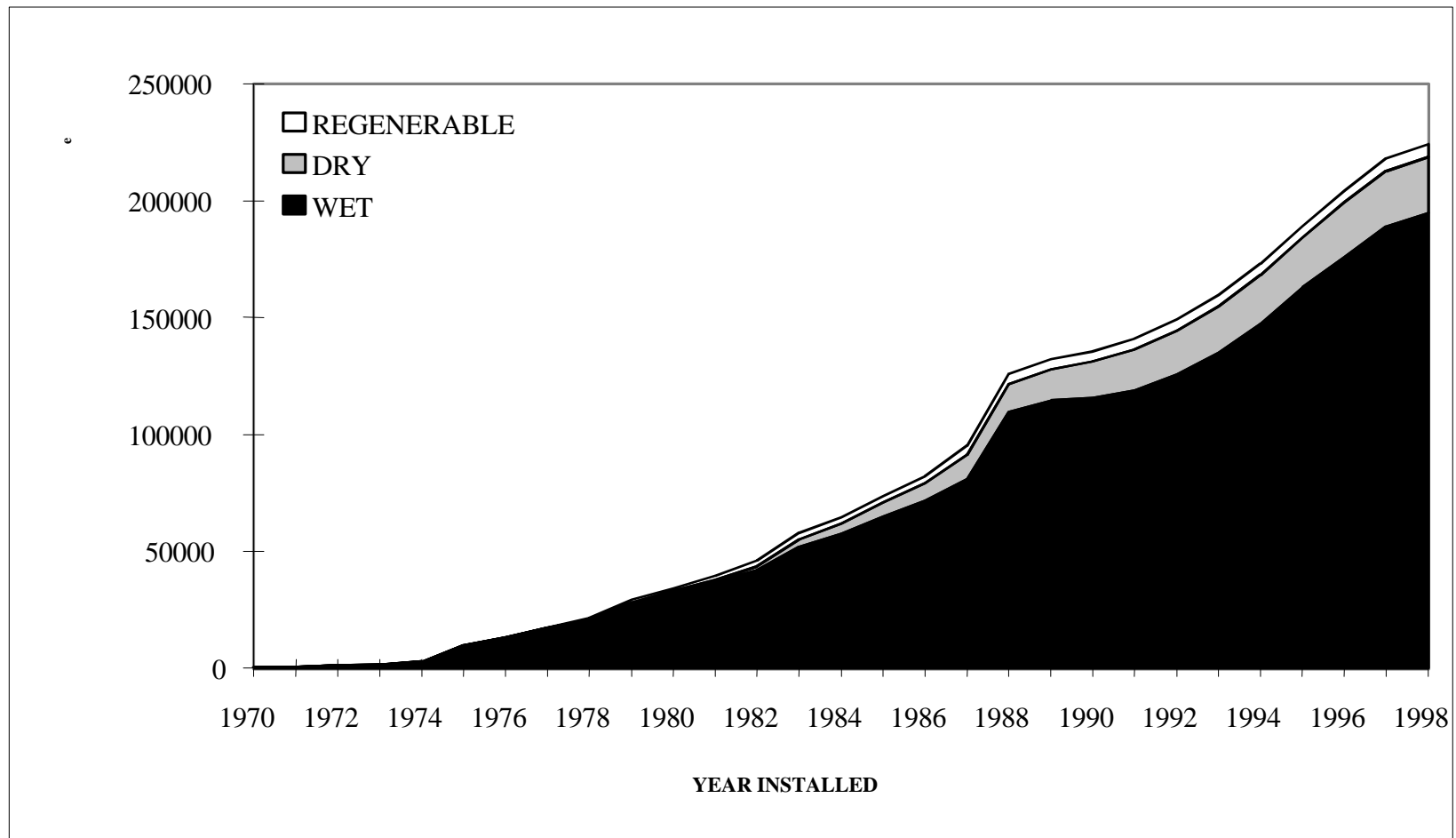


Figure 3-2. Historical application of FGD technology throughout the world.

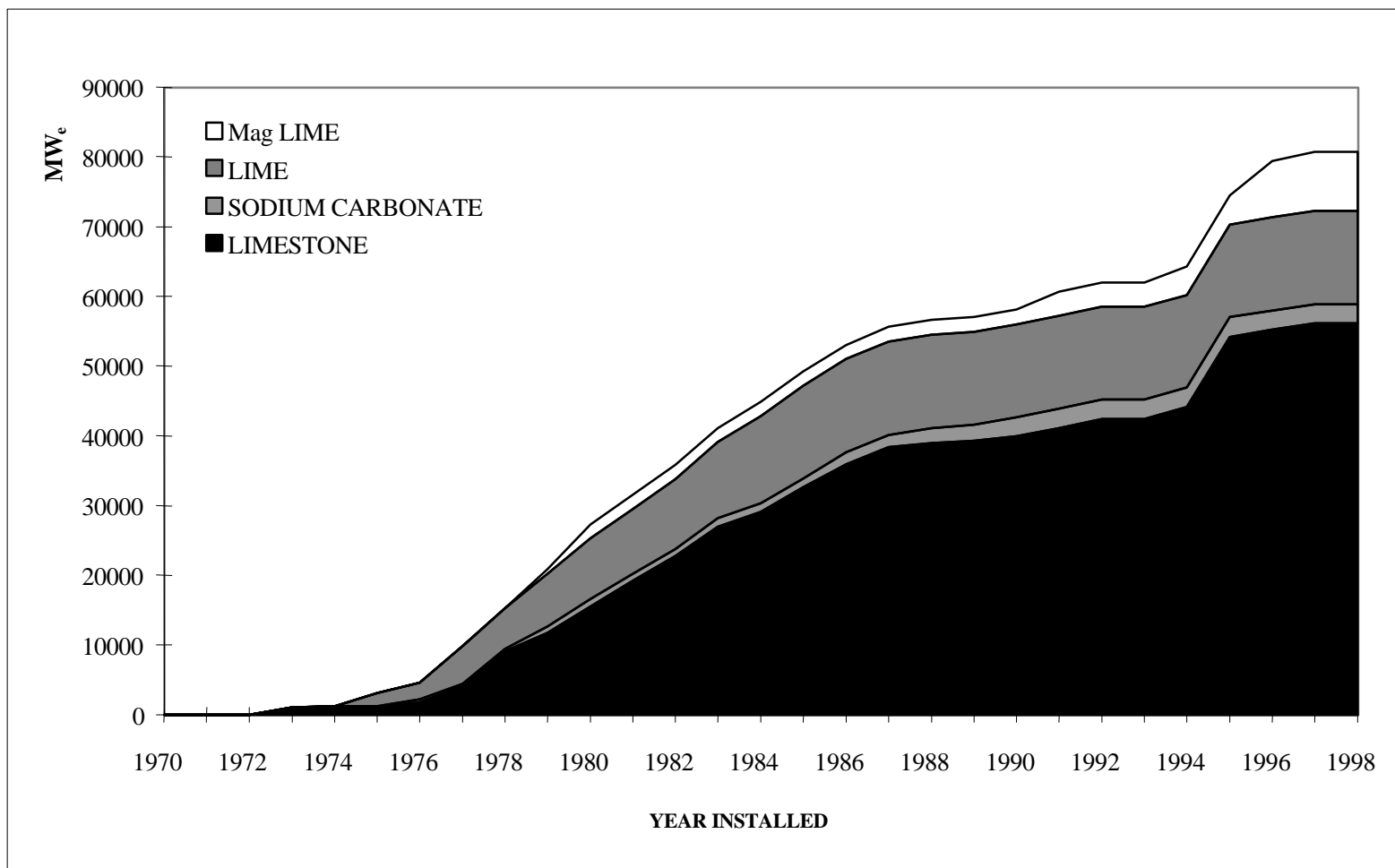


Figure 3-3. Wet FGD technology application in the United States.

Historical applications of dry technologies in the United States are shown in Figure 3-4. As presented in this figure, the spray drying process has historically dominated applications in the United States throughout the '80s and '90s. The late '80s and early '90s saw a mild increase of the installed capacity of duct sorbent injection. There were also a few furnace sorbent injection commercial applications during the early '90s and CFB applications in the mid '90s. Clearly, the spray drying process has been popular among the dry FGD technology processes.

Finally, historical applications of regenerable processes in the United States are shown in Figure 3-5. Regenerable processes (e.g., sodium sulfite, magnesium oxide, sodium carbonate, activated carbon, amine) have not seen any increase in their installed capacity following their initial application.

In summary, the majority of historical applications of FGD technology in the United States, as well as throughout the world, have utilized wet limestone and spray drying processes. Wet FGD technology, other than the wet limestone process, either uses a more expensive sorbent (lime) or is limited by the local availability of the specific sorbent used by the process (e.g. sodium carbonate process). Dry FGD technology, other than LSD, either does not enjoy significant commercial experience (e.g., CFB and FSI) or offers only limited sorbent utilization (e.g., DSI).

The LSD has enjoyed a relatively steady increase in installed capacity in the United States since its initial application in the early '80s. Wet limestone installed capacity increased sharply during the '80s, stagnated during early '90s, then experienced a step increase during the late '90s (due to the

impact of the Clean Air Act Amendments of 1990).

Current Application

Table 3-1 shows statistics describing the installation of FGD systems at fossil-fuel-fired electric power plants through 1998. FGD systems were installed to control SO₂ emissions from over 226,000 MW_e generating capacity, worldwide. Of FGD systems installed on this capacity, 86.8 percent consist of wet FGD technology, 10.9 percent consist of dry FGD technology, and the balance consist of regenerable FGD technologies. Through 1998, almost 100,000 MW_e of capacity in the United States had FGD technology. Of these FGD systems installed, 82.9 percent consist of wet FGD technology, 14.2 percent consist of dry FGD technology, and the balance consist of regenerable FGD technologies. The percent shares of the three FGD technology categories installed are shown in Figure 3-6.

The pattern of installations in the U.S. and abroad reflects that wet FGD technologies predominate over other FGD technologies. It is generally recognized that high SO₂ removal efficiency, coupled with cost effectiveness, has been responsible for the overwhelming popularity of wet FGD technologies, particularly wet-limestone-based FGD technologies. While the earlier wet FGD systems produced only waste by-product sludge, recent systems produce salable by-product gypsum. This has likely increased the attractiveness of wet FGD technologies. Limited application of dry FGD technologies, compared to wet FGD technologies, is likely the result of their higher reagent cost and limited choices for by-product disposal.

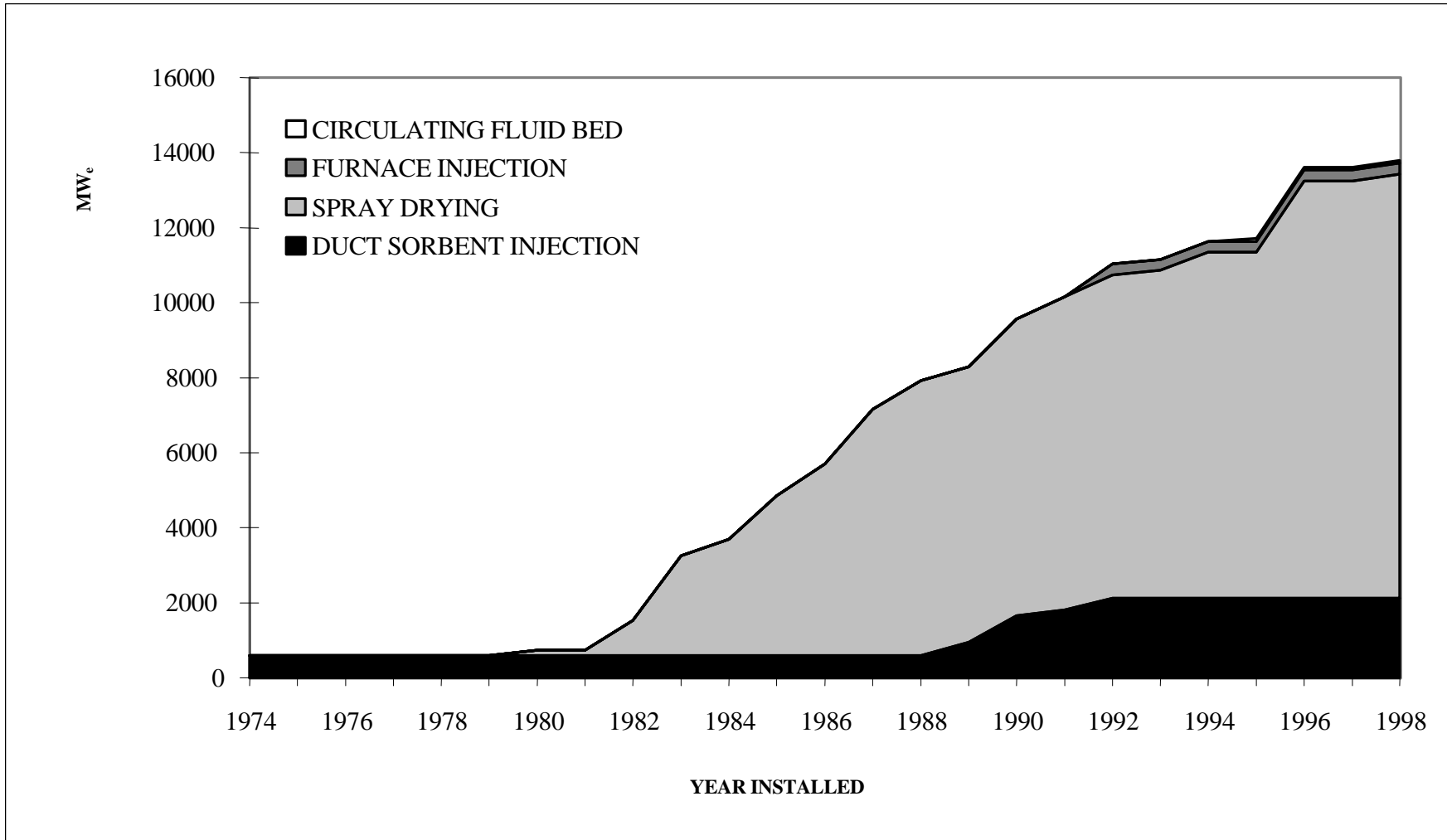


Figure 3-4. Dry FGD technology application in the United States.

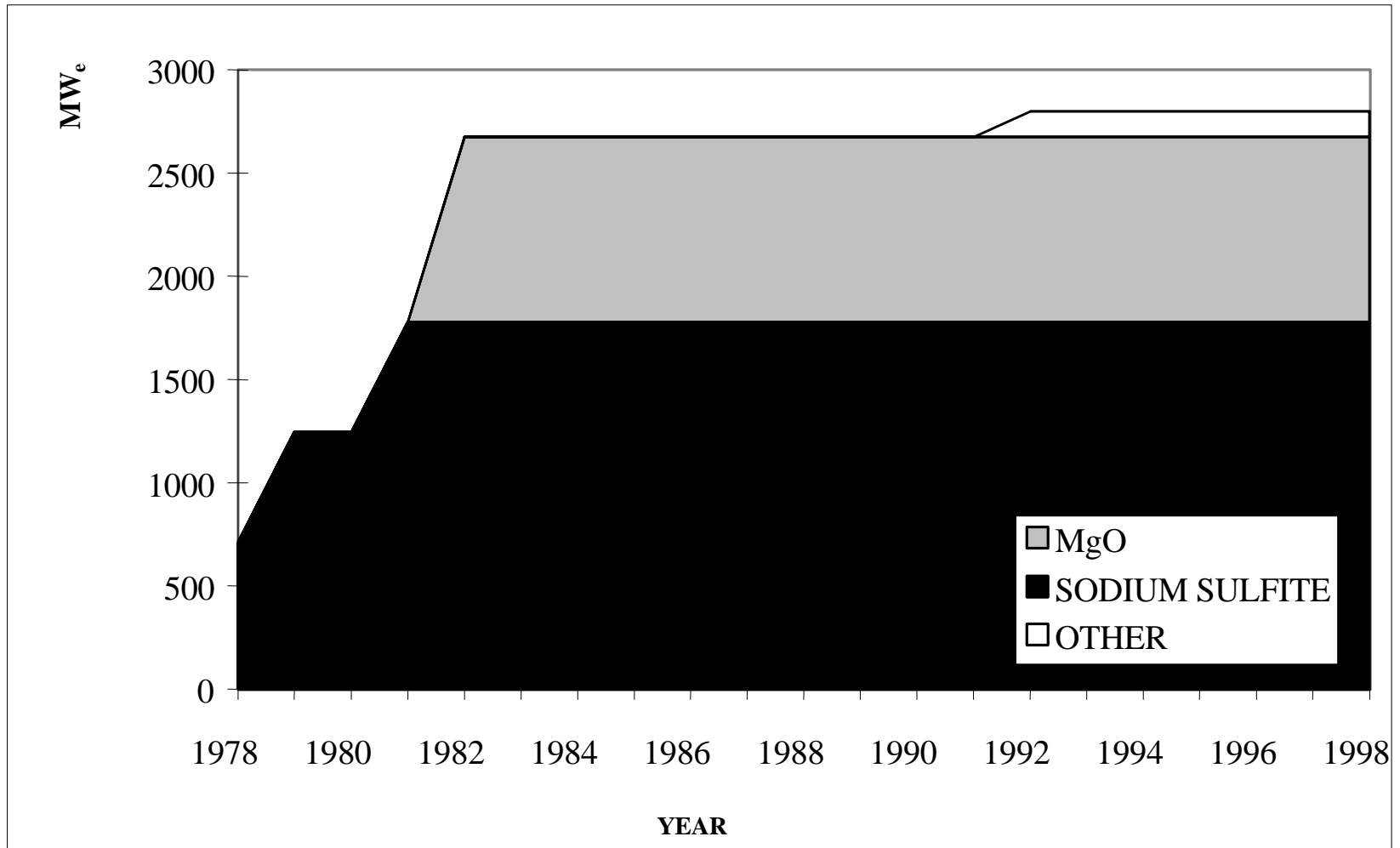


Figure 3-5. Regenerable FGD technology application in the United States.

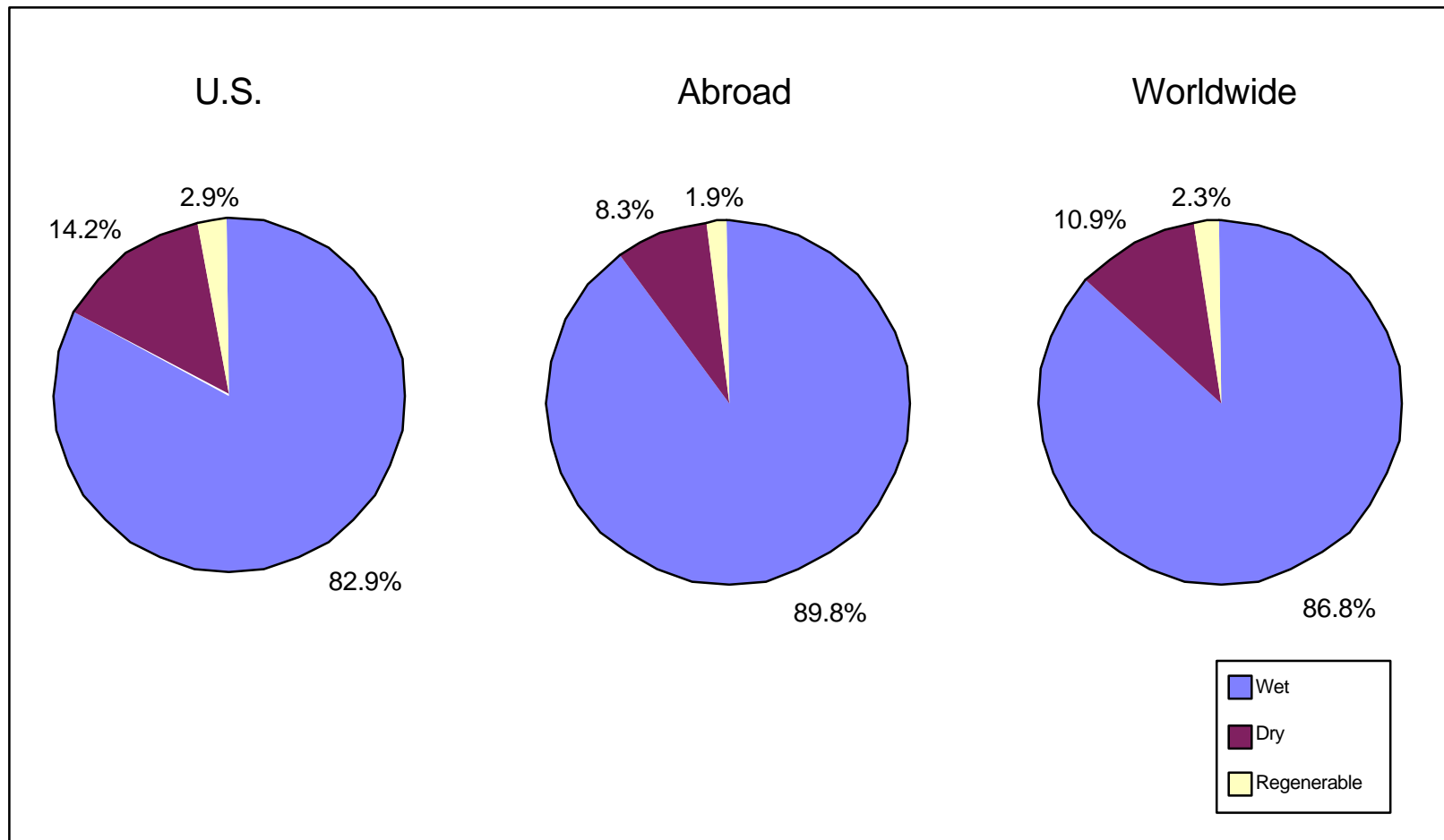


Figure 3-6. Percent shares (capacity) of the three FGD technologies installed.

Table 3-1. Coal-fired Electrical Generation Capacity (MW_e) Equipped with FGD Technology (1998)

Technology	United States	Abroad	World Total
Wet	82,092	114,800	196,892
Dry	14,081	10,654	24,735
Regenerable	2,798	2,394	5,192
Total FGD	98,971	127,848	226,819

Table 3-2 shows capacities of various wet FGD technology systems at power plants in the United States and abroad. Of the United States wet FGD technology installations, 68.9 percent use limestone processes. Abroad, limestone processes are used on as much as 93.2 percent of the total wet FGD technology installations. This trend is shown in Figure 3-7, which shows the division of wet FGD technology applications into limestone and non-limestone ones. The main difference in the pattern of wet FGD technology use in the United States and abroad is the extent of the application of dolomitic lime and sodium carbonate processes. The attractiveness of these processes depends on the local availability of the special sorbents they require. Limited availability of these special sorbents abroad has likely limited the application of the two processes. In the U.S., dolomitic lime and sodium carbonate processes have been applied on some units due to reagent availability at particular sites.

Table 3-3 shows statistics describing the pattern of use of dry FGD technologies. Of the worldwide capacity equipped with dry FGD technology, 73.7 percent use the spray drying process. This compares with 80.4 percent equipped with the spray drying

process in the U.S. Almost all of the remaining installations of dry FGD technology use sorbent injection, which includes furnace (with and without a downstream humidifier) and duct (calcium compound as well as sodium compound) injection. The dominance of the spray drying process within the dry FGD technology category is because this process is more economical for low-to-moderate-sulfur coal applications than wet FGD technology.

These processes have been used commercially in the U.S. since the early '80s and abroad since the mid '80s. Other dry technology processes are considered to be niche applications for retrofit systems, where only limited SO₂ removal is required.

Further understanding of recent FGD technology selections made by the U.S. electricity generating industry can be gained by examining the recent FGD technology installations in the U.S. Between 1991 and 1995, 19,154 MW of U.S. electric generating capacity were retrofitted with FGD technologies. Of this capacity 75, 17.5, and 7.5 percent were equipped with LSFO, MEL, and LSD, respectively.

Table 3-2. Total Capacity (MW_e) Equipped with Wet FGD Technology (1998)

Process	United States	Abroad	World Total
Limestone	56,560	106,939	163,499
Lime	14,237	4,338	18,575
MEL	8,464	50	8,514
Sodium Carbonate	2,756	-	2,756
Seawater	75	1,050	1,125
Regenerable (other)	-	2,423	2,423
Total Wet FGD	82,092	114,800	196,892

Table 3-3. Total Capacity (MW_e) Equipped with Dry FGD Technology (1998)

Process	United States	Abroad	World Total
Spray Drying	11,315	6,904	18,219
Dry Sorbent Injection	2,400	1,125	3,525
CFB	80	517	597
FSI	286	2,108	2,394
Total Dry FGD	14,081	10,654	24,735

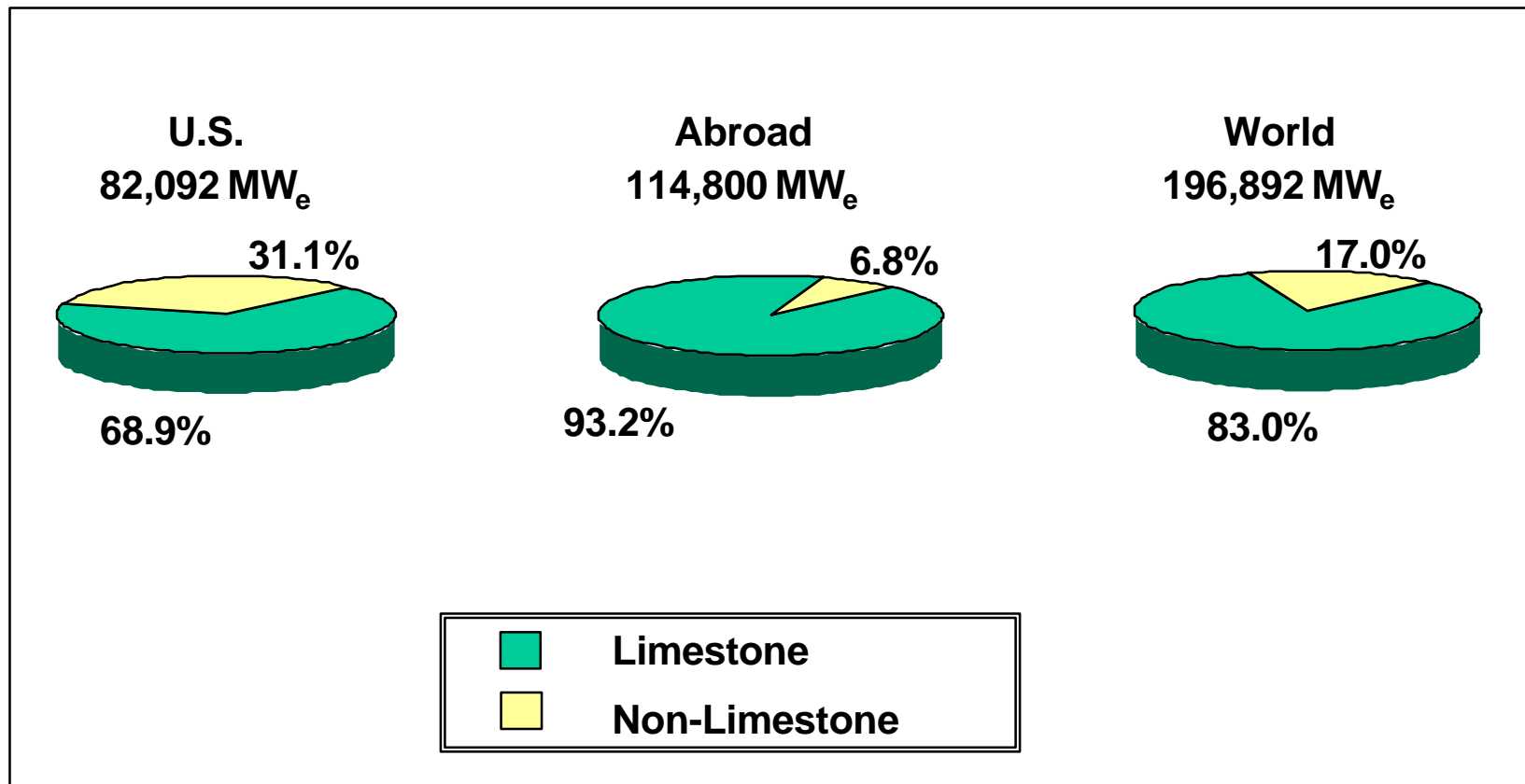


Figure 3-7. Comparison of limestone and non-limestone wet FGD applications.

Table 3-4 shows additional statistics describing the worldwide installation of FGD systems on electric power plants. Through 1998, 668 FGD systems have been installed. Of the installed FGD systems, 522 were wet FGD technology, 124 were dry FGD technology, and the balance consisted of regenerable FGD technologies. Through 1998, 236 FGD technology systems were installed in the U.S. Of the installed FGD systems, 174 were wet FGD technologies, 54 were dry FGD technologies, and the balance consisted of regenerable FGD technologies.

Combining the data from Table 3-4 with those from Table 3-1 allows calculation of

representative sizes of FGD systems for each of the technologies considered. These representative sizes are shown in Table 3-5. These average sizes were arrived at by dividing the MW_e shown in Table 3-1 by the pertinent number of FGD systems shown in Table 3-4.

As seen in Table 3-5, the installations of wet FGD technology in the U.S., as well as those abroad, appear to be larger than installations of dry or regenerable categories of FGD technologies. Additionally, the average FGD system size in the United States is considerably larger than abroad.

Table 3-4. Number of Installed FGD Technology Systems (1998)

Technology	United States	Abroad	World Total
Wet	174	348	522
Dry	54	70	124
Regenerable	8	14	22
Total FGD	236	432	668

Table 3-5. Average Size (MW_e) of FGD Technology Systems (1998)

Technology	United States	Abroad	World Total
Wet	472	330	377
Dry	261	152	199
Regenerable	350	171	236

CHAPTER 4

PERFORMANCE

Introduction

As discussed in Chapter 3, LSFO, MEL, and LSD have been the dominant processes in terms of the electric generating capacity equipped with FGD over the last 30 years. Therefore, the remainder of this report will focus on issues related to these processes.

SO₂ Removal Efficiency

An estimate of SO₂ removal performance of FGD processes can be obtained by examining the design SO₂ removal efficiencies of these processes reported in the CoalPower3 Database. Table 4-1 shows design SO₂ removal efficiencies for wet limestone and LSD processes. These data reflect that wet limestone systems have been designed for high levels of SO₂ removal, up to 98 percent. However, most wet limestone systems appear to be designed for 90 percent SO₂ removal. All LSFO systems installed after 1990 have design SO₂ removal greater than 90 percent. The units with low design efficiencies are generally associated with plants burning low sulfur fuels.²⁸ Also, the units with the design efficiency at the low end of the range given in

Table 4-1 are reported by the CoalPower3 Database to have been installed in the 70s. It is likely that the low design efficiencies are a result of unit specific requirements for permitting purposes, rather than technology limitations. It is also likely that new regulatory requirements were a catalyst for technology improvements by creating a market for more stringent SO₂ control.

Even though the median design efficiency for all units with wet limestone processes in CoalPower3 Database is 90 percent, it should be emphasized that advanced, state-of-the-art wet scrubbers are capable of routinely achieving SO₂ removal efficiencies of over 95 percent. The high velocity LSFO process, with state-of-the-art design options, is reportedly capable of removing more than 99.6 percent of SO₂ under test conditions.²⁹

As seen in Table 4-1, the range and median of SO₂ reduction efficiency at LSD installations are 70-96 and 90 percent, respectively. Spray dryers often achieve greater than 90 percent SO₂ removal on coals with up to 2 percent sulfur.^{30,31} CoalPower3 data also indicate that all spray dryers installed during the period from 1991 to 1995 have a design SO₂ removal efficiency of between 90 and 95 percent.

The performance of wet limestone and LSD processes has improved significantly over the period of their application. To investigate this improvement, the median design SO₂ removal efficiency was determined for the pertinent populations of wet limestone and LSD installations for each of the three decades: 1970-1979, 1980-1989, and 1990-1999. The design efficiencies reported in the CoalPower3 Database were used to determine median design SO₂ removal efficiency.

Table 4-1. Design SO₂ Removal Efficiencies

FGD Technology	Range of Design Efficiency, percent	Median Design Efficiency, ^a percent
Wet Limestone Processes	52-98	90
LSD Processes	70-96	90

^aDerived based on CoalPower3 reported data. Application conditions for wet limestone and LSD processes may differ (e.g., coal sulfur percent).

Since the LSD did not become commercial until the early '80s, no median efficiency could be characterized for the '70s for this process. For each of the last three decades, median design SO₂ removal efficiencies, as well as ranges of reported design SO₂ removal efficiencies, for the wet limestone and LSD are shown in Figure 4-1. A steady increase of the design SO₂ removal efficiency can be noted for wet limestone and spray drying processes. This improvement may be due, in part, to the increasing need to better control SO₂ emissions. However, the trends do reflect that the SO₂ removal efficiency for the processes considered has improved with time.

Energy Requirements

As described previously, once-through wet FGD technology (and specifically, LSFO) has enjoyed the largest extent of application among all FGD technologies. Therefore, it would be reasonable to expect any efforts undertaken to improve energy efficiency of FGD to be initiated on once-through wet FGD systems. A review of the existing literature reveals numerous efforts aimed at increasing energy efficiency of wet FGD systems. Both, design and operational issues were considered in order to improve the energy efficiency.

Modern LSFO absorbers operate at high flue gas velocities in order to achieve improved mass transfer and decrease absorber capital cost at the same time. Flue gas velocity as high as 20 ft/s was achieved under test conditions. In an effort to improve the energy efficiency, a new inlet design has been implemented that incorporates the inlet duct/absorber transition into the flared section of the absorber. It is claimed that this new design allows for a 33 percent pressure drop reduction for absorbers operated at as much as 20 ft/s gas velocity.³²

In a recent survey of LSFO O&M cost,³³ pumping of sorbent slurry was consistently ranked as the most energy intensive component in the operation of wet FGD systems. Pumping sorbent slurry raises the slurry from tank to spray header level and provides pressure necessary for fine atomization. A decrease in the efficiency of droplet/flue gas mixing must be compensated for by increasing L/G in order to maintain the target efficiency for SO₂ removal. Therefore, it is important to utilize a spray that has been atomized within the spray tower for maximum contact with the flue gas. In-depth computational fluid dynamics studies, coupled with field tests, have revealed a radial gradient of SO₂ concentration in a

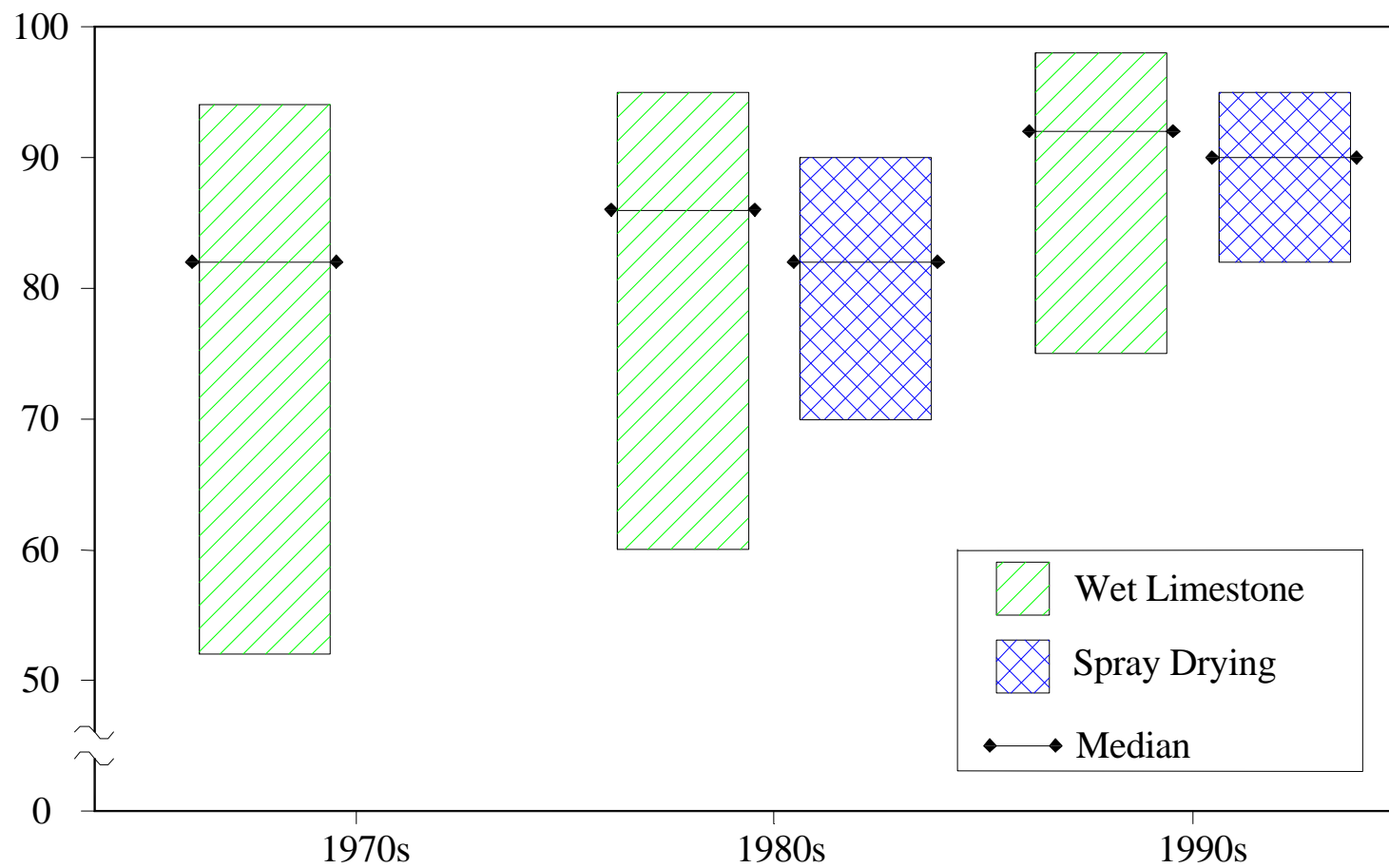


Figure 4-1. Design SO₂ removal efficiencies for wet limestone and spray drying processes.

wet limestone absorber.³⁴ To remedy this undesirable occurrence, guide vanes along the perimeter of the tower could be used. When installed on a 250 MW_e absorber, the guide vanes allowed for a 30-percent L/G reduction. This reduction in L/G cut energy consumption by as much as 20 percent.

Another energy intensive system in the operation of LSFO system is limestone pulverization. The quality and fineness of grinding are critical operational parameters that affect mass transfer properties in an absorber. Horizontal ball milling is a preferred method to pulverize limestone for wet FGD. It is well suited to FGD service because it offers a large reduction capability, resistance to abrasion, and relatively low operation, control, and maintenance requirements.³⁵ Depending on the mode of grinding, a horizontal ball mill consumes 32 and 25 kWh/dry ton of limestone for the dry and wet mode of operations, respectively. Attrition grinding, a new method being considered, has allowed for a reduction of approximately 50 percent in energy consumption, and uses only 15 kWh/dry ton of limestone. An attrition grinder involves a stationary vessel and internally stirred grinding media (balls). Continuous attrition grinders have been demonstrated that are capable of grinding 6 mm limestone down to 95 percent minus 325 mesh.³⁵

EPA's recently published cost estimation algorithm, CUECost, estimates energy power requirements for LSFO and LSD. CUECost estimates energy consumption for LSFO without DBA addition at 2 percent of the net generating capacity of the unit prior to adding pollution controls. With DBA addition, the LSFO power consumption estimate is reduced to 1.65 percent of the net generating capacity. The LSD power consumption is

estimated at 0.7 percent of the net generating capacity.

Applicability

There are some technical constraints to using the spray drying process on applications with high sulfur coal. In the U.S., this process has typically been used in applications on units burning low-to-medium-sulfur coal.³⁶ There has been a great deal of discussion regarding the use of this process on units with high sulfur coal requiring removal efficiencies of over 80 percent. For each spray dryer, there exists a maximum solids concentration (sorber slurry concentration) above which the slurry cannot be easily atomized. High sulfur coal applications may require sorber slurry concentrations in excess of the maximum, since the amount of water that can be evaporated is limited by the desired approach to adiabatic saturation and temperature of the flue gas leaving the absorber.

Another technical constraint may be the unit's physical size, which is a function of the amount of flue gas to be treated. Typically, spray drying has been applied to generating units smaller than 300 MW_e.³⁶ However, spray dryers have also been installed on larger units using multiple absorbers. Successful operation of a spray dryer is dependent on a uniform mixing of finely atomized sorber slurry with flue gas. In large spray dryer vessels, the limited penetration of the atomized sorber slurry may compromise control efficiency.

CHAPTER 5

ADVANCES

Introduction

Over the last 30 years, significant advances have been made in wet limestone FGD processes. As discussed before, once-through dry FGD is a newer technology (applications began in early '80s) and only a few applications were seen in the United States during the late '80s and during the '90s. Since once-through wet FGD has been involved with the bulk of FGD technology applications during this period, no significant advances in once-through dry FGD have been reported. Therefore, only recent advances in wet FGD will be discussed in this report. Some of these advances have been aimed at improving the performance and cost-effectiveness of established processes, while others have focused on developing new processes. The initial part of this chapter discusses once-through wet FGD technology advances. It discusses both advances that can be used to increase the performance of existing once-through wet FGD systems and advances that can be used in the construction of new once-through wet FGD systems. The chapter then concludes with discussion of a new technology - ammonia scrubbing.

Once-through Wet FGD Technology

At present, several technical options exist for upgrading the performance of existing installations using wet limestone processes.

These options include:³⁷

- increasing the sorbent amount used per mole of SO₂;
- increasing the reactivity of the limestone slurry with organic acid (e.g., dibasic acid) addition;
- using more reactive sorbents;
- increasing L/G by increasing the recycle slurry flow rate (requires more pumping power);
- installing a perforated tray or other device to increase mass transfer;
- reducing the amount of gas that is bypassed (requires more fan power); and
- improving gas/liquid hydrodynamics (e.g. guide vanes).

In general, selecting from the above options, the existing installations may be upgraded to achieve removal efficiencies of 95 percent or more.

When considering the feasibility of upgrade scenarios, interrelations between increased SO₂ removal efficiency and many physical and technical parameters require a thorough evaluation. For example, the addition of more sorbent may require the expansion of the reagent preparation capacity and may require better or increased sorbent preparation (milling) capacity. Any increase in efficiency will result in increased waste output, slurry transport, dewatering, and waste disposal capacity.³⁸

The economics of FGD processes, affected by technical advances and regulatory

requirements, are driving numerous conversions of existing older wet FGD systems to more advanced ones. These conversions are aimed at achieving improved SO₂ removal efficiencies and/or waste minimization. Limestone wet FGD systems can be converted to MEL systems to increase SO₂ removal efficiency. For example, an inhibited oxidation limestone wet scrubber designed for 85 percent SO₂ removal at an L/G of 70 (gal/1,000 ft³) and 10 ft/s velocity has been converted to MEL lime.³⁹ Following the conversion, SO₂ removal efficiency increased to 96.7 percent at an L/G of 23 (gal/1,000 ft³).

In another example of a vintage wet FGD system upgrade, conversion of an inhibited

oxidation wet FGD process to a LSFO system was initiated in 1997.⁴⁰ The objective of this conversion was to initiate production of commercial-grade gypsum in place of calcium sulfite waste, which used to be fixated via pozzolanic reaction with lime and fly ash prior to disposal in a landfill.

Several advanced design, process, and sorbent options are now available for new wet FGD scrubbers.⁴¹ These options are shown in Table 5-1. If implemented, some of these advanced design options are capable of providing high SO₂ removal and/or improving the operational efficiency of wet scrubbers while at the same time, reducing cost.

Table 5-1. Advanced Options for New Wet FGD Scrubbers

Option	Approach
Design	large capacity modules
	increased flue gas velocity in scrubber
	concurrent flow
	improved mist eliminator
	improved hydraulics
	superior materials of construction
	low-energy spray nozzles
Sorbent	organic acid buffering
	ultrafine limestone grind
Process	wet stack
	<i>in-situ</i> oxidation
	<i>ex-situ</i> oxidation with MEL
	wastewater evaporation system
	gypsum stacking for final disposal

Among design improvement options, construction of large capacity modules (single module per unit) results in significant capital savings (up to 35 percent) compared to the baseline multi-module design.^{42,43} A single tower absorber serving 890 MW_e and two units has recently been reported as being under construction.⁴⁴ The FGD system went into operation in January 2000 as the largest absorber in the United States and one of the largest in the world.

Increased flue gas velocity in the scrubber allows for a reduced vessel size. The reduced vessel size is possible because of increased mass transfer coefficients resulting from higher gas/liquid relative velocity, increased turbulence, and increased percentage of droplets suspended within the scrubber.^{12, 29, 45, 46} Utilization of a concurrent flow pattern provides a benefit in the form of a reduced pressure drop across the vessel.⁴⁷

A considerable amount of computational fluid dynamics modeling effort has been invested in design advances for mist eliminators.⁴⁸ Modifications include shape (forward tilt into the gas flow), spacing (additional drainage), and orientation (horizontal better than vertical for high velocity scrubbers). These modifications benefit the user with improved mist eliminator cleaning, reduced liquid/particulate matter carryover, and minimized droplet re-entrainment.

Design modifications also include improved hydraulics intended to intensify gas/liquid contact throughout the system. Intensified gas/liquid contact results in improved gas velocity profiles across the spray tower.⁴⁹ Improvements include: optimized placement and selection of nozzles as well as installation of wall rings to eliminate sneakage close to the wall.^{29, 50} Hydraulic model tests have

revealed that an optimum positioning of the flue gas inlet in the flared section of the absorber can significantly reduce its overall pressure drop.³² In order to provide better mixing of air and slurry as well as to improve air distribution in the reaction tank, a rotary air sparger has been used for LSFO.⁵¹

Finally, the advanced wet absorber system design includes new materials of construction, such as alloys, clad carbon steel, and fiberglass, to provide corrosion resistance at an optimum cost.⁵² In addition, electrochemical protection is being used to minimize the corrosion in reaction tanks for systems with high fluoride concentrations. This type of corrosion protection has been determined as the most cost effective for such applications.⁵³

Among improved sorbent options, the use of organic acid buffering allows a reduced vessel size and/or increased efficiency through increased sorbent utilization. Organic acids, such as dibasic acid (DBA), can be added to limestone slurry in a wet limestone process to improve SO₂ removal, sorbent utilization, and/or a particular system's operation. The increased SO₂ removal efficiency in the presence of DBA is a result of its buffering action (limiting the pH drop) at the liquid/gas interface.⁵⁴

An ultrafine limestone grind improves limestone dissolution in the reaction tank (reaction tank size reduction) and even in a spray zone.^{29, 55} An additional option is to implement the direct use of pulverized limestone, eliminating the need for on-site grinding.

Some process modifications are aimed at increasing the energy efficiency of the process and include operation with a wet

stack (no gas reheat) and a wastewater evaporation system. The latter is accomplished by liquid purge injection into the hot flue gas upstream of the electrostatic precipitator (ESP). In this option, wastewater from solids handling/dewatering operation is evaporated in the flue gas. Other process options include *in-situ* forced oxidation, which results in waste with better dewatering characteristics for disposal.⁵⁶

Recent process advances in MEL FGD technology on full commercial-scale incorporate *ex-situ* oxidation to produce gypsum with excellent purity and bright white color. By-product $Mg(OH)_2$ can be produced optionally for in-plant use or sale. This $Mg(OH)_2$ can be used for boiler injection for SO_3 control, to minimize air preheater fouling, and/or $PM_{2.5}$ -related stack emission.

MEL can offer some advantages over LSFO. It can operate with high SO_2 removal efficiency (98 percent plus) in high sulfur coal applications, low L/G ratio, smaller scrubbers and recirculating pumps, and lower energy requirement.

Ammonia Scrubbing

Over the last few years, a promising wet FGD process has been under development. This process, wet ammonia FGD, has the potential to improve waste management in conjunction with providing SO_2 removal efficiency in excess of 95 percent.⁵⁷ Operators of conventional wet limestone FGD processes may be confronted with saturated markets for commercial-grade gypsum of FGD origin. At present, the wet ammonia FGD process offers the advantage of an attractive ammonium sulfate $[(NH_4)_2SO_4]$ by-product that can be used as fertilizer.

This process also has the potential for becoming a promising option for units burning high sulfur coal, as it is also capable of removing other acid gases (e.g., sulfur trioxide $[SO_3]$ and hydrogen chloride $[HCl]$) in addition to SO_2 . While HCl emissions can be reduced concurrently with SO_2 emissions using currently commercial FGD technology, the removal of SO_3 and control of sulfuric acid (H_2SO_4) aerosol is not as straightforward. Depending on the type of FGD technology, a considerable portion of H_2SO_4 aerosol may exit the stack as a respirable fine particulate emission and may cause a visible plume.⁵⁸

Ammonia scrubbing of SO_2 offers an alternative for maximizing the value of the by-product produced in a wet FGD system.⁵⁹ With the ongoing deregulation in the electric utility industry, the cost of generation for large power generation units is continually under scrutiny. For units that utilize wet, limestone-based FGD, producing a salable by-product, such as gypsum, is a means for reducing the cost of operation. However, the United States has an abundant supply of natural gypsum and, as a result, the price for FGD gypsum produced by LSFO system has historically been very low.⁶⁰ Apparent problems related to the economics of FGD gypsum can potentially be overcome by ammonia scrubbing. The reaction of ammonia, SO_2 , and oxygen in an absorber installed in an ammonia scrubbing system produces $(NH_4)_2SO_4$ fertilizer. Recently, prices of ammonia are reported to have decreased considerably,⁶⁰ making the use of ammonia as a reagent much more economically favorable compared to a few years ago. The best opportunities to apply ammonia scrubbing technology will likely be found at power plants in a proximity to

navigable water or good rail access, and a location with high $(\text{NH}_4)_2\text{SO}_4$ prices.

The ammonia scrubbing process, as currently envisioned,⁶¹ employs a counterflow spray tower design that is similar in configuration to the existing wet limestone-based FGD systems. In some cases, prescrubber may be used to humidify the flue gas and/or remove HCl prior to the main absorption stage. The flue gas then enters the counterflow spray tower where it is contacted by a solution of $(\text{NH}_4)_2\text{SO}_4$ liquor.

The ammonia is stored in a pressurized or refrigerated vessel and pumped as a liquid to a vaporizer. The vaporizer typically uses steam to vaporize the ammonia prior to introducing it into the oxidation air or directly into the absorber reaction tank.

Ammonia is added with the oxidation air to maintain the recycle liquor at the desired pH to ensure that the required SO_2 removal is achieved. The cleaned flue gas passes through mist eliminators to remove any entrained droplets. The absorber is operated in a pH range selected to eliminate ammonia slip and aerosol formation. Conversion of ammonium sulfite $[(\text{NH}_4)_2\text{SO}_3]$ and ammonium bisulfite $(\text{NH}_4\text{HSO}_3)$ to $(\text{NH}_4)_2\text{SO}_4$ takes place in the reaction tank via injection of compressed air. $(\text{NH}_4)_2\text{SO}_4$ solution (10 - 25 wt percent dissolved solids) is bled from the absorber. Fresh makeup water required by the process is added to the absorber reaction tank to maintain the liquid level in the tank.⁶¹

Aerosol emissions are a concern with ammonia scrubbing processes and must be addressed carefully in the design. The simultaneous presence of ammonia, SO_2/SO_3 , and water vapor in flue gas can result in the

formation of ammonia/sulfur aerosols. The aerosols are very small (0.1 to 0.3 micrometer) and, once formed, are emitted from the absorber, causing a visible plume at the stack discharge.

Although it is theoretically possible to operate an ammonia scrubbing system in a plume-free mode with very precise control of absorber pH, temperature, solution concentration, etc., wet electrostatic precipitators (WESPs) are generally employed downstream of the absorber to eliminate concerns related to aerosol emissions. Even with excellent process control, aerosol emissions can occur as a result of load changes, sulfur inlet changes, non-ideal gas/liquid contact, and pH control problems. An added benefit of the WESP is the control of SO_3 mist and droplet emissions.

$(\text{NH}_4)_2\text{SO}_4$ fertilizer production can be accomplished in two ways:

- Remote crystallization and drying
- *In-situ* crystallization, dewatering, and granulation

In the case of remote crystallization/drying, the absorber loop operates with a clear solution of $(\text{NH}_4)_2\text{SO}_4$ (approximately 30 to 35 percent). The solution is sent to an adjacent by-product processing plant, which consists of a crystallizer, centrifuge, and dryer. Thermal energy (steam) or vapor recompression/evaporation is used to concentrate the solution to the point where crystallization takes place.

Advantages to clear solution operation include:

- Discreet $(\text{NH}_4)_2\text{SO}_4$ crystals are formed in a device specifically designed for that purpose, so that the size can be carefully controlled to meet the required product specifications
- The monolithic crystals are not subject to attrition or dusting during shipping or handling
- The $(\text{NH}_4)_2\text{SO}_4$ solution can be filtered prior to crystallization, thus eliminating any concern with solid contaminants (e.g. fly ash) in the byproduct
- The entire absorber loop operates with clear solutions and is not subject to the plugging and erosion concerns associated with slurry scrubbing

With *in-situ* crystallization processes, slurry from the pre-absorber is passed to a dewatering hydroclone, where the slurry solids concentration is increased to about 35 weight percent. The purpose of the hydroclone is two-fold: to dewater the slurry from the prescrubber to optimize the centrifuge feed slurry density; and to separate the fine particles (primarily ash from the boiler) from the product, and thus maintain product purity. The slurry is next pumped to a series of centrifuges where the slurry is dewatered to 97 - 98 percent solids. Centrifuges discharge the material immediately into a rotary drum dryer where heated air is passed over the crystals to further dry the material to less than 1 percent moisture.⁶¹

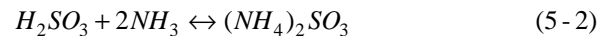
To maximize the by-product value, the $(\text{NH}_4)_2\text{SO}_4$ material must be converted to larger granular crystals. To accomplish this,

raw $(\text{NH}_4)_2\text{SO}_4$ material is transferred to a compaction system. In this system, fresh feed $(\text{NH}_4)_2\text{SO}_4$ material is mixed with the raw $(\text{NH}_4)_2\text{SO}_4$ in a pug mill mixer. Finally, the material from the mixer is compacted into hard flakes subsequently discharged into a flake breaker. The flake breaker crushes the large flakes into smaller pieces, later sized in a series of sizing mills. The final acceptably sized product is transported to the storage area.

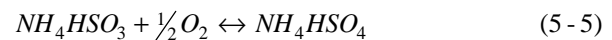
The chemistry of the production of $(\text{NH}_4)_2\text{SO}_4$ from boiler flue gas is very similar to the chemistry of wet limestone FGD. SO_2 from the flue gas is absorbed in the spray tower by water according to the equation:



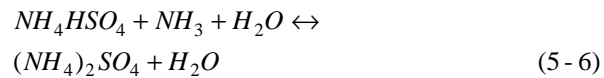
The H_2SO_3 is then reacted in a reaction tank with ammonia to form $(\text{NH}_4)_2\text{SO}_3$ and NH_4HSO_3 :



$(\text{NH}_4)_2\text{SO}_3$ and NH_4HSO_3 are also oxidized in the absorber (forced oxidation) to form $(\text{NH}_4)_2\text{SO}_4$ and NH_4HSO_4 :



The NH_4HSO_4 is neutralized in the presence of ammonia and water to form $(\text{NH}_4)_2\text{SO}_4$:



Because of the relatively high value of $(\text{NH}_4)_2\text{SO}_4$ fertilizer, the economics of ammonia scrubbing improve as the sulfur content of the fuel increases. Consequently, ammonia scrubbing offers a potential to the plant to use high sulfur fuel such as high sulfur coal or petroleum coke. Thus, the need for use of more expensive low sulfur coal could be avoided. Petroleum coke has been identified as a low cost, high sulfur fuel that can be burned in many boilers. A recent study⁶¹ concluded that, as more refineries use crude with higher sulfur content, both the quantity and sulfur content of the coke will increase. Since many refineries are expected to have difficulty disposing of the coke with the high sulfur content, this fact should lead to attractive prices for the material.

The attractiveness of the ammonia scrubbing process appears to depend on the ability of the plant to sell $(\text{NH}_4)_2\text{SO}_4$ fertilizer. An evaluation of the price of $(\text{NH}_4)_2\text{SO}_4$ over a period of 11 years has indicated a sustained increase.⁶¹ This has been explained by its value as a nutrient for selected crops and its ability to replenish the sulfur deficiency in soils.

A successful demonstration of 90-95 percent SO_2 removal and aerosol-free operation has recently been reported⁶⁰ for a 130 MW_e system installed on boilers burning 2 to 3.5 percent sulfur coal.

CHAPTER 6

FGD COST

General Approach

As discussed before, LSFO, LSD, and MEL have been the processes of choice in recent U.S. applications. Therefore, in this work, state-of-the-art cost models were developed for these processes. These state-of-the-art models are collectively called State-of-the-art Utility Scrubber Cost Model (SUSCM) and are expected to provide budgetary cost estimates for future applications. In the ensuing paragraphs, descriptions and results are provided for the state-of-the-art LSFO, LSD, and MEL cost models developed in this work.

The Air Pollution Prevention and Control Division (APPCD) of EPA's National Risk Management Research Laboratory (NRMRL) has recently published the Coal Utility Environmental Cost Workbook (CUECost).⁶² CUECost provides budgetary cost estimates (± 30 percent accuracy) for between 100 and 2000 MW_e net LSFO and LSD applications based on user-defined design and economic criteria. CUECost algorithms provided the starting point for the LSFO and LSD cost models developed in this work. For each of these models, first, a sensitivity analysis was conducted to determine those

variables that have a minor impact on cost (i.e., a deviation of less than 5 percent over the selected baseline). Then, these variables were fixed at typical values to arrive at a simplified cost model. Next, the simplified LSFO and LSD cost models were validated with published data. Finally, these models were further adjusted with cost-effective design decisions to arrive at state-of-the-art LSFO and LSD cost models.

For costing purposes, MEL can be considered to be a combination of LSFO and LSD. In the MEL, sorbent (magnesium-enhanced slurry) is prepared in a similar manner to that used in LSD, and this sorbent is contacted with flue gas in an absorber similar to a typical LSFO absorber. However, because MEL sorbent is more reactive than LSFO sorbent, less flue gas residence time is needed in the MEL absorber. As such, a MEL absorber is significantly smaller than a corresponding LSFO absorber. Further MEL waste handling equipment operates in a fashion similar to that in LSFO, producing gypsum by-product. Considering these characteristics of the MEL, for costing purposes this process can be considered to be a combination of LSFO and LSD. Therefore, the LSFO and LSD algorithms developed as described above were used appropriately to develop the MEL cost model. As for LSD and LSFO, cost-effective design choices were made to arrive at a state-of-the-art MEL cost model.

Limestone Forced Oxidation

For the sensitivity analysis, the baseline consisted of an LSFO application on a 500 MW_e unit with a 10,500 Btu/kWh heat rate, burning 3.4 percent sulfur (S) Jefferson, OH, coal (heating value of 11,922 Btu/lb), and presenting medium retrofit difficulty.

The primary design elements fixed in this baseline LSFO application were materials for construction of the absorbers, addition of DBA, a wet stack, and gypsum stacking disposal. The choice of materials for construction is known to have a major cost impact. Selection of rubber-lined carbon steel (RLCS) had the largest cost impact, saving nominally 0.65 mills/kWh over alloy construction. Other variables were fixed to CUECost default values for the baseline LSFO, including 95 percent removal of SO₂. This baseline LSFO retrofit requires one absorber serving a maximum size of 700 MW_e. Thus defined, baseline LSFO has an annual operating cost of 10.31 mills/kWh.

Sensitivity analyses were performed using the CUECost outputs resulting from single-variable perturbations from the baseline. The results of these analyses are summarized in Table 6-1. Perturbations in the variables were selected to span the range of realistic values (e.g., unit size ranged between 100 and 2000 MW_e). The high and low values of variables were selected and the corresponding costs were then determined for each single-variable perturbation. Next, the prediction differences were calculated between baseline and high, as well as low, values for each perturbed variable.

Based on the results of sensitivity analyses, shown in Table 6-1, it was determined that the majority of cost impacts (cost impacts greater than ± 5 percent) can be captured with capacity, heat rate, coal sulfur content, coal heating value, capacity factor, and disposal mode.

The remaining variables were determined to have a minor impact on cost and, therefore, they were fixed at typical values. The list of variables that have minor impacts on the cost,

as predicted by the sensitivity analyses, is given in Table 6-2. Furthermore, the values selected to fix these minor variables are also shown in Table 6-2. For example, the air heater outlet temperature that was shown to have between 1.4 and -0.5 percent impact when varied between 360 and 280 °F respectively, was fixed at 300 °F, as shown in Table 6-2. These fixed values are based on the CUECost defaults.

Fixed operation and maintenance (O&M) cost in the simplified LSFO cost model accounts for the cost associated with operating labor, maintenance labor and materials, and administration and support labor. The variable O&M cost is composed of reagent cost, disposal cost, steam cost, and energy cost. The assumptions used in calculating these costs are based on the default values provided in CUECost and the suggested values in Electric Power Research Institute's Technical Assessment Guide (EPRI TAG).

CUECost determines capital cost for FGD system as Total Capital Requirement (TCR). The cost estimation begins with the installed equipment capital cost (BM). Following the EPRI TAG's methodology, the installed BM cost is then multiplied by appropriate factors to incorporate costs of general facilities, engineering fees, contingencies, and the prime contractor's fee, resulting in an estimate of Total Plant Cost (TPC). Financial factors related to the time required to construct the FGD equipment are applied to TPC to estimate Total Plant Investment (TPI). TCR is the sum of TPI, inventory cost, and pre-production costs. Pre-production cost incorporates one-twelfth of the projected annual O&M expenses and 2 percent of the TPI estimate.

Table 6-1. Sensitivity Analysis of LSFO Annual Operating Cost (baseline cost of 10.31 mills/kWh)

Variable, units	Baseline	Variable's High Value	Variable's Low Value	Cost for High Value of Variable, mills/kWh	Cost for Low Value of Variable, mills/kWh	High Value Difference, ^d %	Low Value Difference, ^e %
Capacity, MW _e	500	2000	100	6.57	22.62	-36.3	119.4
Heat Rate, Btu/kWh	10,500	11,000	8,000	12.25	9.13	18.8	-11.4
Coal Sulfur Content, %	3.43	4.0	1.5	10.60	9.71	2.8	-5.8
Coal Heating Value ^a , Btu/lb	11,922	14,000	10,500	9.56	10.92	-7.3	5.9
Air Heater Outlet, °F	300	360	280	10.45	10.26	1.4	-0.5
SO ₂ Removal, %	95	98	90	10.36	10.22	0.5	-0.9
L/G	125	160	60	10.36	10.22	0.5	-0.9
Slurry Concentration, % solids	15	20	10	10.31	10.31	0	0
Capacity Factor, %	65	90	40	8.04	15.41	-22.0	49.5
DBA ^c Addition	no	N/T ^b	yes	N/T	10.20	N/T	-1.1
Disposal Mode	stacking	landfill	wallboard	12.51	10.23	21.3	-0.1
Absorber Material	alloy	N/T	RLCS	N/T	9.66	N/T	-6.3
No. of Absorbers	1	2	N/T	10.41	N/T	1.0	N/T
Reheat	yes	N/T	no	N/T	9.94	N/T	-3.6

^aCoal Data: Form EIA-767, DOE ²⁸

^bN/T=not tested

^cDibasic Acid

^dDifference = (Cost for High Value of Variable – 10.31) / 10.31 • 100%

^eDifference = (Cost for Low Value of Variable – 10.31) / 10.31 • 100%

Table 6-2. Representative Values for LSFO Variables with Minor Cost Impacts

Variable	Units	Value	Comments
Coal Heating Value ^a	Btu/lb	11,900	Baseline 11,922
Limestone Composition	% CaCO ₃	95.3	
SO ₂ Control Efficiency	%	95	
L/G	gal/1,000 ft ³	125	70 with DBA
Ambient Pressure	in. Hg	29.4	
Air Heater Outlet Temperature	°F	300	
Moisture in the Flue Gas	%	14.0	
Max Fan Capacity	cfm	1,600,000	Either 2, 4, or 8 fans
Chimney Inlet Gas Temperature	°F	127	

^a Not a minor impact; value is set to 11,900 Btu/lb.

Capital Cost

LSFO systems consist of five major equipment areas: reagent feed, SO₂ removal, flue gas handling, waste handling, and support equipment. As described before, capital cost algorithms for these areas in CUECost were simplified to be functions of capacity, heat rate, coal sulfur content, and coal heating value only. Summation of these adjusted algorithms provides the total capital cost in the simplified LSFO cost model.

The above five areas are shown schematically for LSFO in Figure 6-1. Accordingly, in cost considerations the capital cost of each area is represented as: Reagent Feed (BM_F), SO₂ Removal (BM_R), Flue Gas Handling (BM_G), Waste Handling (BM_W), and Support Equipment (BM_E). The estimation methods used for the five major equipment areas are described below.

The BM_F, BM_R, and BM_W cost estimates were explicitly determined by the SO₂ feed rate to the FGD system. This feed rate was determined by the coal sulfur content and

coal use rate, with no provision for sulfur retention in the ash. SO₂ flow rate to the FGD system (FR_{SO₂}) was estimated from the amount of sulfur in the coal as well as the coal burn rate at full load:

$$FR_{SO_2} = \frac{Wt\%S \cdot 1000}{HHV} \cdot \left[\frac{64}{32} \right] \cdot MW_e \cdot HR \quad (6-1)$$

where Wt%S is coal sulfur content (wt%), MW_e is LSFO size (MW_e), HR is plant heat rate (Btu/kWh), and HHV is coal heating value (Btu/lb).

Reagent Feed Area

The BM_F cost (including receiving, storage, and grinding) – a fourth order polynomial in limestone addition rate was used based on CUECost. The limestone addition rate was determined based on the SO₂ feed rate to the absorber, reagent addition rate, SO₂ removal requirement, and limestone CaCO₃ content. In CUECost (and in this simplified model), all the sulfur in the coal was assumed to be delivered to the FGD system as SO₂.

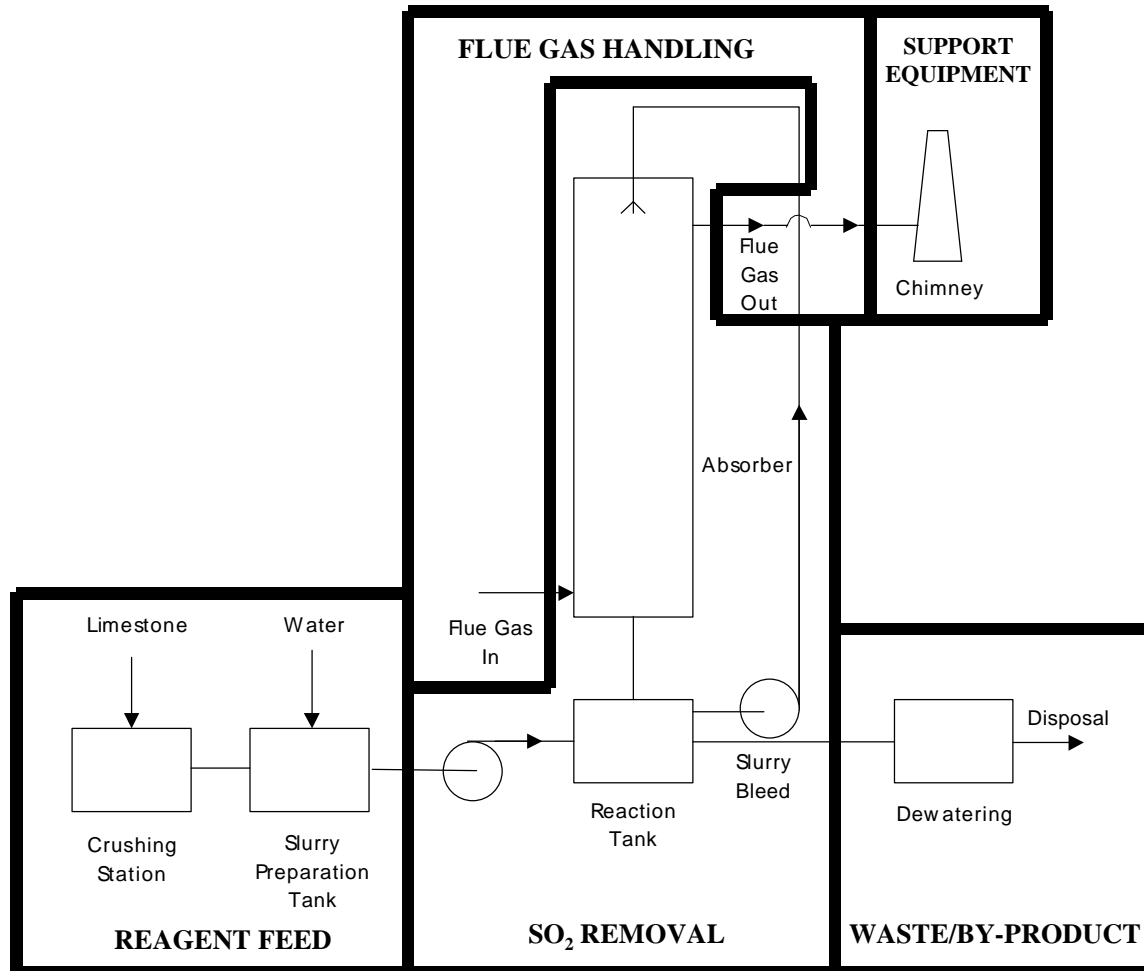


Figure 6-1. Schematics of LSFO system's equipment areas.

CUECost adjusts the reagent feed ratio to ensure the CaCO_3 present is sufficient to remove all the chlorine in the coal as CaCl_2 , in addition to the specified SO_2 removal. However, chlorine removal has been eliminated from this model based on the assumption that it has a negligible cost impact. Specifically, the cost of limestone and reagent addition was calculated as follows:

- Cost of ball mill and hydrocyclones - second order polynomial on limestone addition rate.
- Cost of DBA supply tank - power law on DBA addition rate, which is, in turn, proportional to the rate of SO_2 removal.

The BM_F cost was estimated based on the limestone feed rate. Limestone composition (purity) has been fixed in this model at 95.3 percent CaCO_3 , which is the default composition used in CUECost. The limestone addition rate has been fixed in this model at 1.05 times the reagent feed ratio.

Reagent feed rate (FR_L) was estimated as:

$$\text{FR}_L = \text{FR}_{\text{SO}_2} \cdot 1.05 \cdot \frac{100}{64} \cdot \frac{0.95}{0.953} \quad (6-2)$$

These parameters allow BM_F to be estimated as follows:

$$\begin{aligned} \text{BM}_F = & \left(-0.0034 \cdot \left(\frac{\text{FR}_L}{1000} \right)^4 \right) + \left(2.1128 \cdot \left(\frac{\text{FR}_L}{1000} \right)^3 \right) \\ & - \left(494.55 \cdot \left(\frac{\text{FR}_L}{1000} \right)^2 \right) + 68164.7 \cdot \frac{\text{FR}_L}{1000} + 7118470 \\ & + C_{B\&H} + C_{DBA} \end{aligned} \quad (6-3)$$

where $C_{B\&H}$ is the cost of the ball mill and hydrocyclones as given by:

$$\begin{aligned} C_{B\&H} = & 32.9 \cdot \left(\frac{\text{FR}_L}{2000} \right)^2 + 22412 \\ & \cdot \left(\frac{\text{FR}_L}{2000} \right) + 1854902 \end{aligned} \quad (6-4)$$

and C_{DBA} is the cost of the DBA tank as given by:

$$C_{DBA} = 364627 \cdot \left(\frac{\text{FR}_{\text{SO}_2} \cdot \frac{0.95 \cdot 20}{2000}}{8.34 \cdot (1 + 0.5)} \right)^{0.283} \quad (6-5)$$

C_{DBA} was added only for LSFO systems with the DBA addition.

SO_2 Removal Area

BM_R cost (including absorbers, tanks, and pumps) – a third order polynomial on SO_2 rate to the scrubber was used based on CUECost. These cost components were calculated as follows:

- Cost of absorbers - power law on flue gas flow rate to each absorber inlet multiplied by the number of absorbers. Different power laws were used depending on absorber construction materials. Maximum absorber size was limited in CUECost to treat 700 MW_e ; larger units required multiple, equal size absorbers.
- Cost of spray pumps - power law applied to the slurry flow rate per absorber per pump multiplied by the number of pumps. The slurry flow rate (gpm) was calculated based on the gas flow rate per absorber at the exhaust temperature, but at 1 in. H_2O less than the inlet pressure (typical absorber inlet pressure drop). L/G

was fixed at 70 for LSFO with DBA and, otherwise, was fixed at 125. CUECost default is L/G of 125 for 95 percent SO₂ removal.

BM_R cost required estimation of flue gas flow through the LSFO system. The absorber cost was estimated based on inlet flue gas flow rate and type of construction materials. The spray pump cost was estimated based on flue gas flow rates exhausting the absorber.

The flue gas flow was calculated in CUECost using the coal analysis in addition to unit size, heat rate, excess air, and air inleakage. This approach was analogous to computing the F-factor (*F_d*) for each fuel. As it was not considered practical to calculate an *F_d* for each fuel, gas flow was estimated using the methodology employed in 40CFR75 Appendix F. An *F_d* of 9,780 scf/10⁶ Btu was applied for all coals, as the differences in coal rank (e.g., 9,860 scf/10⁶ Btu for lignite) were expected to have negligible impact on the estimated scrubber cost. Flue gas flow into the absorber (ACFM) was calculated as follows:

$$ACFM = \frac{1000}{10^6} \cdot \frac{9780}{60} \cdot \frac{(460 + 295)}{528} \cdot \frac{100}{(100 - 6)} \cdot MW_e \cdot HR \cdot \left(\frac{0.04}{P} + \frac{0.209}{P} \cdot \frac{(P - 0.04)}{(0.209 - P)} \right) \quad (6-6)$$

where P is %O₂ in the stack (9 percent O₂ in the stack was assumed).

The pressure at the absorber inlet was fixed at 10 in. H₂O gauge, the CUECost default. Ambient pressure was fixed at the CUECost default of 29.4 in. Hg. Temperature of the flue gas entering the absorber might have been varying significantly for different units but was expected to have minimal impact on

cost, based on the sensitivity analysis. Absorber inlet temperature was fixed in the model at 295 °F, resulting from 300 °F air heater outlet temperature used as the default in CUECost. The moisture fraction was assumed to be 6.0 percent H₂O at the absorber inlet.

The cost of the spray pumps for the absorbers was estimated based on the absorber outlet flue gas flow rate and the number of pumps (*N_p*) required. The *N_p* required was based on the required slurry flow rate per absorber and a maximum single pump capacity of 43,000 gpm (CUECost default). The required slurry flow rate was determined by L/G, dependent on whether the design incorporated DBA additive. The gas flow rate was determined at 127 °F and at 9 in. H₂O gauge (CUECost default). Moisture content was estimated at 14 percent H₂O (CUECost calculated). CUECost estimated air addition at 2 moles oxygen for each mole of sulfite to be oxidized (CUECost default). For a typical SO₂ concentration, this air addition is less than 1 percent by volume of the total flow and has not been included.

The above assumptions allowed estimation of BM_R cost, depending on the absorber construction material used and on the presence of DBA addition in the system as follows:

$$BM_R = BARE\,MODULE_R + ABSORB\,ER \cdot N_a + PUMP \cdot N_p \quad (6-7)$$

where ABSORB_{ER} is the absorber cost equal to:

$$ABSORB\,ER\,1 = 173978 \cdot \left(\frac{ACFM}{1000} \right)^{0.5575} \cdot N_a \quad (6-8)$$

or to:

$$\text{ABSORBER } 2 = 230064 \cdot \left(\frac{\text{ACFM}}{1000} \right)^{0.5638} \cdot N_a \quad (6-9)$$

for the RLCS or alloy material of construction, respectively (N_a is number of absorbers).

The cost of pumps, PUMPS, was expressed as:

$$\text{PUMPS} = 910.85 \cdot \left(\frac{F_{GPM}}{N_p} \right)^{0.5954} \cdot N_p \quad (6-10)$$

where GPM is slurry flow rate (gpm) and N_p is the number of pumps. The slurry flow rate varied, depending on whether dibasic acid additive was selected.

Auxiliary cost for the SO₂ Removal Area (BARE MODULE_R) was calculated as follows:

$$\begin{aligned} \text{BARE MODULE}_R = & 0.8701 \cdot \left(\frac{FR_{SO_2}}{1000} \right)^3 - 188.2 \\ & \cdot \left(\frac{FR_{SO_2}}{1000} \right)^2 + 34809 \cdot \left(\frac{FR_{SO_2}}{1000} \right) + 1905302 \end{aligned} \quad (6-11)$$

Flue Gas Handling Area

The flue gas handling system cost (ductwork and ID fans) was based on CUECost, a polynomial on flue gas flow rate entering the absorbers, exiting absorbers, and number of absorbers. If a design included reheat, a term was added for the required temperature increase. The cost of ID fans was estimated using a power law based on the inlet gas flow

rate per fan multiplied by the number of fans required.

The cost of the BM_G was based on the number of absorbers, flow entering the absorbers, and flow exiting the absorbers. Pressure of the gas exiting the absorbers was fixed at 4 in. H₂O gauge. The temperature of the gas exiting the absorbers was fixed at 127 °F, the CUECost default wet bulb temperature. Flue gas moisture content was approximated at 14 percent H₂O at the absorber outlet and through the remainder of the FGD system (CUECost default).

The cost of fans was estimated by a power law based on the number of fans required and the flue gas flow rate. Fans were assumed to be installed in groups of 2, 4, or 8 with a maximum individual fan capacity of 1,600,000 cfm. The number of fans was based on conditional tests of the smallest number option (2, 4, or 8) resulting in an individual fan capacity of less than 1,600,000 cfm. Inlet pressure for sizing fans was fixed at 12 in. H₂O vacuum. Temperature at the fan inlet was fixed at 295 °F (CUECost method).

If the design incorporated reheat, BM_G cost was adjusted according to the design temperature increase. The reheat temperature rise was fixed at 25 °F (CUECost default).

By assuming the above design criteria, BM_G cost was estimated as follows:

$$\text{BM}_G = \text{BARE MODULE}_G + \text{ID FANS} \quad (6-12)$$

The auxiliary cost of the Flue Gas Handling Area (BARE MODULE_G) was calculated as:

$$\begin{aligned} \text{BARE MODULE}_G = & -0.1195 \cdot \left(\frac{\text{ACFM}}{1000} \right)^2 \\ & + 777.76 \cdot \left(\frac{\text{ACFM}}{1000} \right) + 238203 + 0.000012 \\ & \cdot \left(\frac{\text{ACFM}}{1000} \right)^3 - 0.1651 \cdot \left(\frac{\text{ACFM}}{1000} \right)^2 + 1288.82 \\ & \cdot \left(\frac{\text{ACFM}}{1000} \right) + 559693 - 0.2009 \cdot \left(\frac{\text{ACFM1}}{1000 \cdot N_a} \right)^2 \\ & + 1266.4 \cdot \left(\frac{\text{ACFM1}}{1000 \cdot N_a} \right) + 420141 \end{aligned} \quad (6-13)$$

where ACFM1 is flue gas flow rate out of the absorber.

The cost of fans (ID FANS) was calculated as:

$$\text{ID FANS} = 91.24 \cdot \left(\frac{\text{ACFM}}{N_f} \right)^{0.6842} \cdot N_f \quad (6-14)$$

where N_f is the number of fans.

Waste/By-product Handling Area

The BM_W cost (dewatering, disposal/storage, and washing) - a second order polynomial on SO_2 mass flow rate for gypsum stacking was used based on CUECost. Moreover, a third order polynomial on SO_2 mass flow rate was used for landfill disposal or wallboard gypsum production. The cost of thickener was estimated as a linear function of the waste solids removal rate. The waste amount was estimated from a mass balance.

The BM_W cost was fixed by the disposal option chosen and by the amount of sludge to be disposed of. The amount of sludge was based on inlet SO_2 flow rate, SO_2 removal

efficiency (fixed at 95 percent), and $CaCO_3$ in the limestone. All SO_2 removed was assumed to be oxidized to form calcium sulfate dihydrate (gypsum). The BM_W cost was estimated as follows:

$$BM_W = \text{BARE MODULE}_W + \text{THICKENER} \quad (6-15)$$

- For the Waste/By-product Handling System with gypsum stacking (BARE MODULE_{W1}):

$$\begin{aligned} BM_{W1} = & -4.0567 \cdot \left(\frac{FR_{SO_2}}{1000} \right)^2 + 1788 \\ & \cdot \left(\frac{FR_{SO_2}}{1000} \right) + 80700 \end{aligned} \quad (6-16)$$

- For the Waste/By-product Handling System with landfill (BARE MODULE_{W2}):

$$\begin{aligned} BM_{W2} = & 0.325 \cdot \left(\frac{FR_{SO_2}}{1000} \right)^3 - 168.77 \cdot \left(\frac{FR_{SO_2}}{1000} \right)^2 \\ & + 29091 \cdot \left(\frac{FR_{SO_2}}{1000} \right) + 773243 \end{aligned} \quad (6-17)$$

- For the Waste/By-product Handling System with wallboard gypsum production (BARE MODULE_{W3}):

$$BM_{W3} = BM_{W2} \cdot 1.25 \quad (6-18)$$

The cost of thickener (THICKENER) was estimated as:

$$\begin{aligned} \text{THICKENER} = & 9018.7 \cdot FR_{SO_2} \cdot 0.95 \\ & \cdot \frac{172}{64 \cdot 2000} + 114562 \end{aligned} \quad (6-19)$$

Support Equipment Area

The BM_E cost (electrical, water, and air) - a third order polynomial based on net generating capacity provided by the user.

The cost of the chimney was estimated with a power law on total flue gas flow exiting each absorber, based on CUECost. Separate power laws were used depending on whether reheat was included in the design.

The BM_E cost was a function of chimney cost. Chimney cost was estimated with a power law based on flow rate per absorber. Temperature at the chimney inlet was selected in the model at 127 °F, while the pressure was selected at 4 in. H₂O gauge. The BM_E cost was estimated as follows:

$$BM_E = \text{BARE MODULE}_E + \text{CHIMNEY} \quad (6-20)$$

For a BM_E with reheat, the cost of chimney (CHIMNEY 1) was estimated as:

$$\text{CHIMNEY 1} = 40208 \bullet ACFM1^{0.3339} \quad (6-21)$$

For a BM_E without reheat, the cost of chimney (CHIMNEY 2) was estimated as:

$$\text{CHIMNEY 2} = 23370 \bullet ACFM1^{0.3908} \quad (6-22)$$

The auxiliary cost for Support Equipment Area was estimated as:

$$\begin{aligned} \text{BARE MODULE}_E &= 0.0003 \bullet MW_e^3 - 1.0677 \\ &\bullet MW_e^2 + 1993.8 \bullet MW_e + 1177674 \end{aligned} \quad (6-23)$$

Total Capital Requirement

Once the BM cost had been determined, it was possible to calculate TCR. The general TCR determination procedure is illustrated in Table 6-3. Following the EPRI TAG methodology, installed BM cost was multiplied by appropriate contingency

factors, resulting in an estimate of Total Plant Cost (TPC). The financial factor D, which includes the effects of inflation on the cost of capital and relates to the time required to construct the FGD equipment, was applied to the TPC to estimate Total Plant Investment (TPI). Finally, TCR is the sum of TPI, inventory cost, and pre-production cost. Pre-production cost incorporated one-twelfth of the projected annual O&M expense and 2 percent of the TPI estimate. Detailed calculations are described below.

Following the EPRI TAG approach, 5 percent for general facilities, 10 percent for engineering and home office, 5 percent for process contingency, and 15 percent for project contingency were applied. This model also included a Prime Contractor's Fee of 3 percent, which is the CUECost algorithm default. This cost was added to arrive at the Total Plant Cost (TPC). Using these CUECost defaults and adding yielded TPC for the model:

$$\begin{aligned} TPC &= BM \bullet \left(1 + \frac{A_1}{100} + \frac{A_2}{100} + \frac{A_3}{100} \right) \\ &\bullet \left(1 + \frac{B}{100} \right) \bullet \left(1 + \frac{C}{100} \right) \end{aligned} \quad (6-24)$$

TPC could then be adjusted for financial factors dependent on the time required to complete the project. Allowance for Funds During Construction Factor (F_{AFDC}) and Total Cash Expended Factor (F_{TCE}) are used to adjust TPC. F_{AFDC} accounts for interest during construction and F_{TCE} allows for de-escalation of cost. CUECost includes time requirements for various size FGD installations.

Table 6-3. TCR Calculation Method

Cost Component	Symbol / Calculation
Capital Cost	$BM = BM_F + BM_R + BM_G + BM_W + BM_E$
Facilities + Engineering & HO ^a + Process Contingencies	$A = A_1 + A_2 + A_3$
Project Contingency	B
Fee	C
Total Plant Cost (TPC)	$TPC = BM * (1 + A) * (1 + B) * (1 + C)$
Financial Factor	$F_{TCE} + F_{AFDC} = D$
Total Plant Investment	$TPI = TPC * D$
Pre-production Cost + Inventory Capital	E
Total Capital Requirement	$TCR = TPI + E$

^a HO = Home Office

Applying the F_{TCE} and F_{AFDC} appropriate to the unit size results in Total Plant Investment (TPI):

$$TPI = TPC \cdot (F_{TCE} + F_{AFDC}) \quad (6-25)$$

In regulatory cost determinations, it is usually preferable to assume constant dollars; e.g. no inflation. Such analysis should yield a F_{TCE} equivalent to 1 (no inflation), and an F_{AFDC} dependent on cost of capital without inflation. Applying an F_{AFDC} rate of 7.6 percent and zero inflation results in factors listed in Table 6-4. Constant dollar factors listed in Table 6-4 are used in the subsequent model development.

The Total Capital Requirement (TCR) was determined by adding pre-production cost and inventory capital to TPI. CUECost estimates pre-production cost as a sum of 2 percent of TPI plus one-twelfth of projected annual fixed O&M cost plus one-twelfth of projected annual variable O&M cost adjusted for the capacity factor, as follows:

$$TCR = 1.02 \cdot TPI + \frac{Fixed\ O\ \&\ M}{12} + \frac{Variable\ O\ \&\ M}{CF \cdot 12} + INVENTORY \quad (6-26)$$

where inventory capital (INVENTORY) is the cost of reagent required to meet the bulk storage requirement. A 60-day limestone inventory was incorporated (limestone cost of \$15/ton was used). CF is plant capacity factor. CF is defined as a ratio of the average output to the rated output of a plant on an annual basis.

Finally, a correction was made to the TCR to account for the cumulative effect of variables with minor cost impact (Table 6-2), which were determined based on the sensitivity analyses. The CUECost-determined TCR for baseline conditions shown in Table 6-1 and for minor effect variables fixed as shown in Table 6-2 was equal to \$205/kW. However, when minor effect variables were set to maximize their combined effect on cost, the resulting value of TCR was \$226/kW. Therefore, TCR was multiplied by the adjustment factor of 1.1024 (226/205) to yield the Adjusted TCR.

Table 6-4. Financial Factors for FGD Construction, Constant Dollars

Unit Capacity	Years to complete	AFDC Factor	TCE Factor
$MW_e < 160$	1	0.0000	1.0000
$160 \leq MW_e < 400$	2	0.0380	1.0000
$400 \leq MW_e < 725$	3	0.0779	1.0000
$725 \leq MW_e < 1300$	4	0.1199	1.0000
$1300 \leq MW_e < 2000$	5	0.1640	1.0000
$MW_e = 2000$	6	0.2104	1.0000

Operation and Maintenance Cost

The O&M cost was calculated next. The O&M cost includes fixed and variable components. Fixed O&M cost incorporates:

- operating labor
- maintenance labor and materials
- administration and support labor

Variable O&M cost is composed of:

- reagent
- dibasic acid
- disposal(by-product credit given)
- steam
- electrical energy

Fixed O&M cost components were estimated as follows. Operating labor (OL) was estimated by the equation below, using a power law on the unit's capacity and estimating the number of workers needed in combination with an operating labor rate (\$30/hr):

$$OL = 41.69041 \cdot MW_e^{-0.322307} \cdot \frac{MW_e \cdot 30 \cdot 40 \cdot 52}{100} \quad (6-27)$$

Maintenance labor and materials (ML&M) cost was determined as a percentage (3 percent) of BM cost. Administration and support (A&S) labor was estimated as a fraction of maintenance labor and materials plus operating labor, as given by the equation:

$$A \& S = 0.3 \cdot (0.4 \cdot ML \& M + OL) \quad (6-28)$$

Variable O&M cost components were estimated as a sum of limestone, DBA, disposal, steam, and electrical energy costs. Cost of limestone (unit price of limestone at \$15/ton) was:

$$C_{CaCO_3} = \frac{FR_L}{2000} \cdot 8760 \cdot CF \cdot 15 \quad (6-29)$$

where CF is capacity factor.

Cost of dibasic acid (unit price of dibasic acid at \$430/ton):

$$C_{DBA} = FR_{SO_2} \cdot \frac{0.95}{2000} \cdot \frac{20}{2000} \cdot 8760 \cdot CF \cdot 430 \quad (6-30)$$

Cost of disposal if the gypsum stacking method is selected (\$6/ton):

$$C_{DS} = 6 \cdot 8760 \cdot CF \cdot FR_{SO_2} \cdot 0.95 \cdot \frac{172}{64 \cdot 2000} \quad (6-31)$$

Cost of disposal if landfilling is selected (\$30/ton):

$$C_{DL} = 30 \cdot 8760 \cdot CF \cdot FR_{SO_2} \cdot 0.95 \cdot \frac{172}{64 \cdot 2000} \quad (6-32)$$

Cost of disposal was set to zero if wallboard production was selected. In addition, for this case a by-product credit (\$2/ton) was given as described below:

$$CREDIT = FR_{SO_2} \cdot 0.95 \cdot \frac{172}{64 \cdot 2000} \cdot 8760 \cdot CF \cdot 2 \quad (6-33)$$

Cost of steam (price of steam estimated at \$3.50/1000 lb):

$$STEAM = \frac{TER}{855.14 \cdot 1000} \cdot 8760 \cdot CF \cdot 3.5 \quad (6-34)$$

Cost of electrical energy (power consumption for LSFO estimated at 2.0 percent) was estimated using the default CUECost power price of 25 mills/kWh:

$$POWER = 0.02 \cdot \frac{(1000 \cdot MW_e \cdot 0.8231)}{1000} \cdot 8760 \cdot CF \cdot 25 \quad (6-35)$$

As an annual expense, the components of variable O&M cost were adjusted for the capacity factor of the unit(s).

Validation

Capital cost predictions of the simplified LSFO cost model were validated against reported capital cost for eight recent retrofit LSFO systems.

LSFO cost estimates derived by the simplified model described above were validated against reported costs (CUECost manual) for eight Phase I plants with retrofit scrubbers. These eight plants included one LSFO retrofit, Gibson, and seven LSFO retrofits of various configurations. Since the simplified cost model incorporates generalizations applied to the CUECost algorithm, it was necessary to validate this model against these recent retrofits.

Model estimates of TCR and published costs are presented in Table 6-5 and are further illustrated in Figure 6-2. These results reflect that the simplified LSFO cost model on the average predicts the published capital cost within 10.5 percent.

In the validation study, a heat rate of 10,500 Btu/kWh and a coal heating value of 11,900 Btu/lb were used for all plants. All of the Phase I units in Table 6-5 were designed without reheat. Absorber materials of construction and the disposal mode for each unit are shown in Table 6-5. The simplified model cost was de-escalated to 1994 dollars to maintain consistency with reported costs.

Table 6-5. Model Validation Summary for LSFO FGD (1994 Dollars)

Plant	Absorber Material/ Disposal	Unit Capacity, MWe	Absorbers	Coal Wt%S	Model, \$/kW	Reported, \$/kW	Deviation, ^a percent
Petersburg	Alloy/landfill	239	1	3.5	400	317	+26.2
Cumberland	RLCS/stacking	1300	3	4.0	164	200	-18.0
Conemaugh	RLCS/wallboard	1700	5	2.8	174	195	-10.8
Ghent	Alloy/stacking	511	3	3.5	213	215	-0.1
Bailly	RLCS/wallboard	600	1	4.5	189	180	+5.0
Milliken	RLCS/wallboard	316	1	3.2	368	348	+5.7
Navajo	Alloy/landfill	750	2	0.75	226	236	-4.2

^aDeviation=(Model-Reported)/Reported•100%

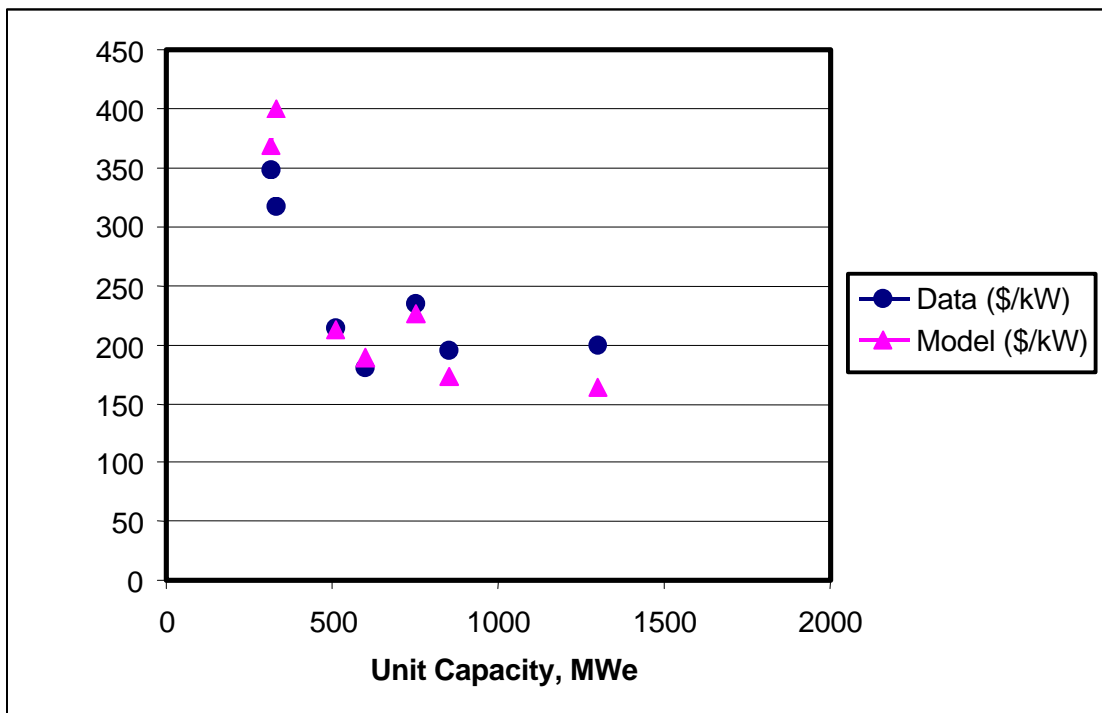


Figure 6-2. Comparison of model predictions with cost data for LSFO.

Recently, IPM model predictions⁶³ of TCR's were published for 2-4 percent sulfur coals. The comparison of the simplified CUECost model prediction to the IPM model for 2, 3, and 4 percent sulfur coals is given in Figure 6-3. As can be seen in Figure 6-3, model predictions of TCR are not very sensitive to coal sulfur content for the range of 2 to 4 percent.

State-of-the-art Model

The algorithms developed thus far incorporated a variety of adjustments to CUECost algorithms to eliminate variables that did not have significant impact on cost. At this point, however, it is helpful to specify a "state-of-the-art" LSFO system by which to estimate the cost of possible future retrofits. It is recognized that alternate design decisions may be made in the interest of reducing cost based on site specific conditions or other engineering features resulting in cost savings not reflected otherwise.

Therefore, the simplified LSFO cost model was further adjusted with cost-effective design decisions to arrive at the LSFO part of the SUSCM (*LSFO SUSCM*). This latter model is expected to provide the budgetary cost estimates for future LSFO applications. The assumptions made in arriving at the LSFO SUSCM are described below.

1. Absorbers serving flue gas from units up to 900 MW_e in capacity are used in the LSFO SUSCM designs. This is consistent with the recently reported information for Units 1 and 2 of Tampa Electric's Big Bend Station. At this station, both units were retrofitted with a single 60-ft diameter 890-MW_e module^{44,64,65}.
2. The "state-of-the-art" scrubber is constructed of rubber-lined carbon steel

or alloy material. Scrubber cost was assumed to be the average of rubber-lined carbon steel and alloy materials.

3. The "state-of-the-art" scrubber uses dibasic acid addition, resulting in modest capital savings and significant O&M savings.
4. The "state-of-the-art" scrubber uses gypsum stacking or wallboard production as the waste disposal method. Waste disposal bare module cost was assumed to be the average of the cost for the two disposal methods.
5. Sorbent inventory of 30 days.
6. The cost of chimney was assumed to be the average of chimney cost with and without reheat.

"State-of-the-art" decisions are shown in Table 6-6.

Combining the equations developed before with these "state-of-the-art" design decisions yields a LSFO SUSCM-derived estimate of TCR for a "state-of-the-art" FGD unit. TCR predictions using LSFO SUSCM are shown in Figure 6-4. These predictions are based on units with a heat rate of 10,500 Btu/kWh and a capacity factor of 90 percent. The results reflect that the capital cost is not sensitive to coal sulfur content. However, as expected, capital cost does reflect an economy-of-scale. It is worth noting that the discontinuities in capital cost curves reflect the addition of an absorber as unit capacity changes from less than 900 MW_e to greater than 900 MW_e and from less than 1800 MW_e to greater than 1800 MW_e. This is because of assumption 1 described above.

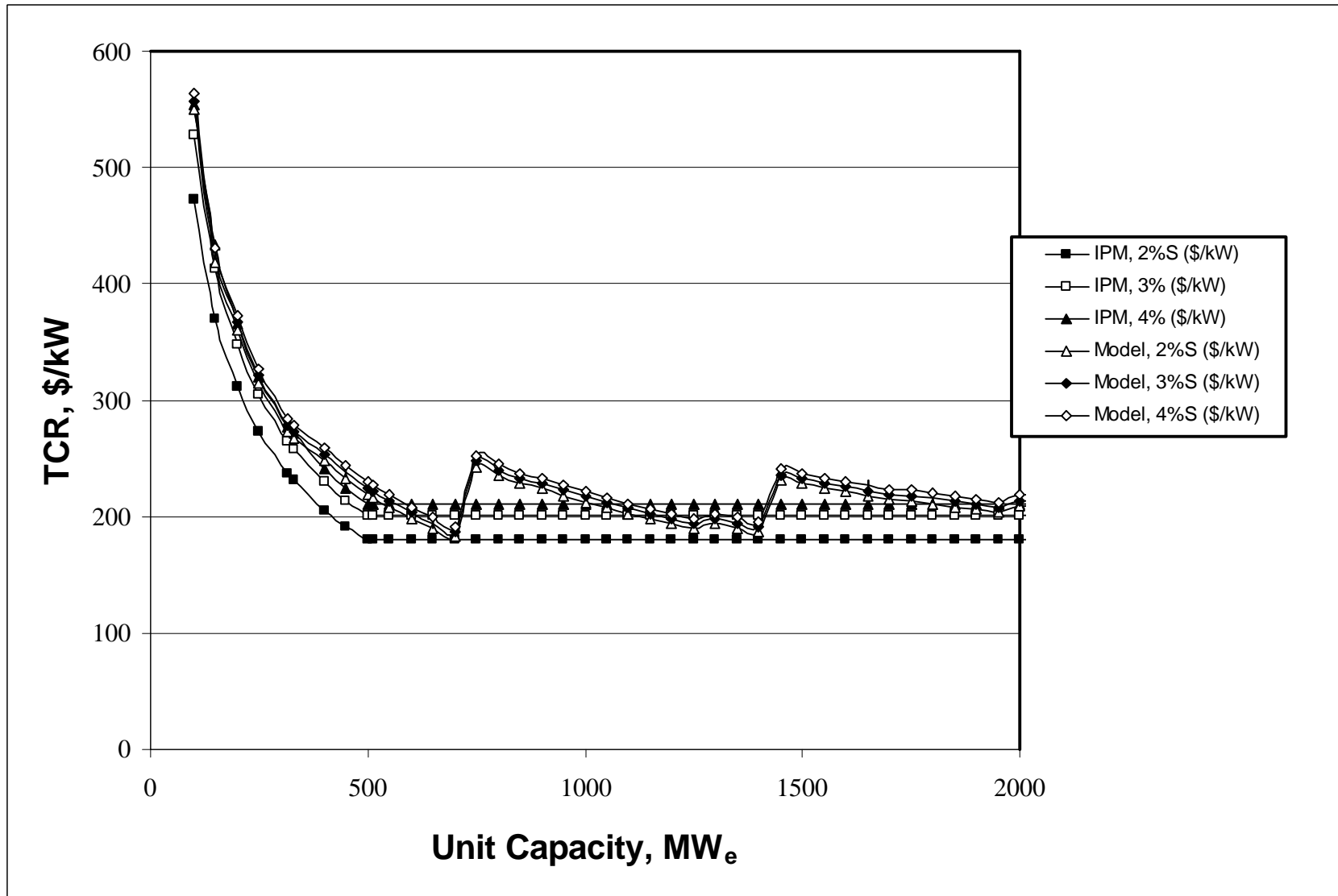


Figure 6-3. Comparison of LSFO cost model to IPM model predictions for 2 to 4 percent sulfur coal.

Table 6-6. “State-of-the-art” LSFO Design Decisions

Parameter	Units	Value
Single Absorber Size	MW _e	900
Absorber Diameter	ft	60
DBA Addition ^a	-----	Yes
L/G	gal/1000 ft ³	70
O ₂ in Stack	%	8
Material of Construction	-----	Average of RLCS and alloy
SO ₂ Removal	%	95
Flue Gas Temperature from Absorber	°F	300
Flue Gas Velocity into Absorber	ft/s	14
Inventory for Limestone	days	30
Limestone Purity (CaCO ₃)	%	95.3
Waste Disposal	-----	Average of wallboard or gypsum stacking
Power Requirement	%	2
Flue Gas Reheat ^a	----	Average of Yes and No

^aYes/No decision only; no addition rate considerations

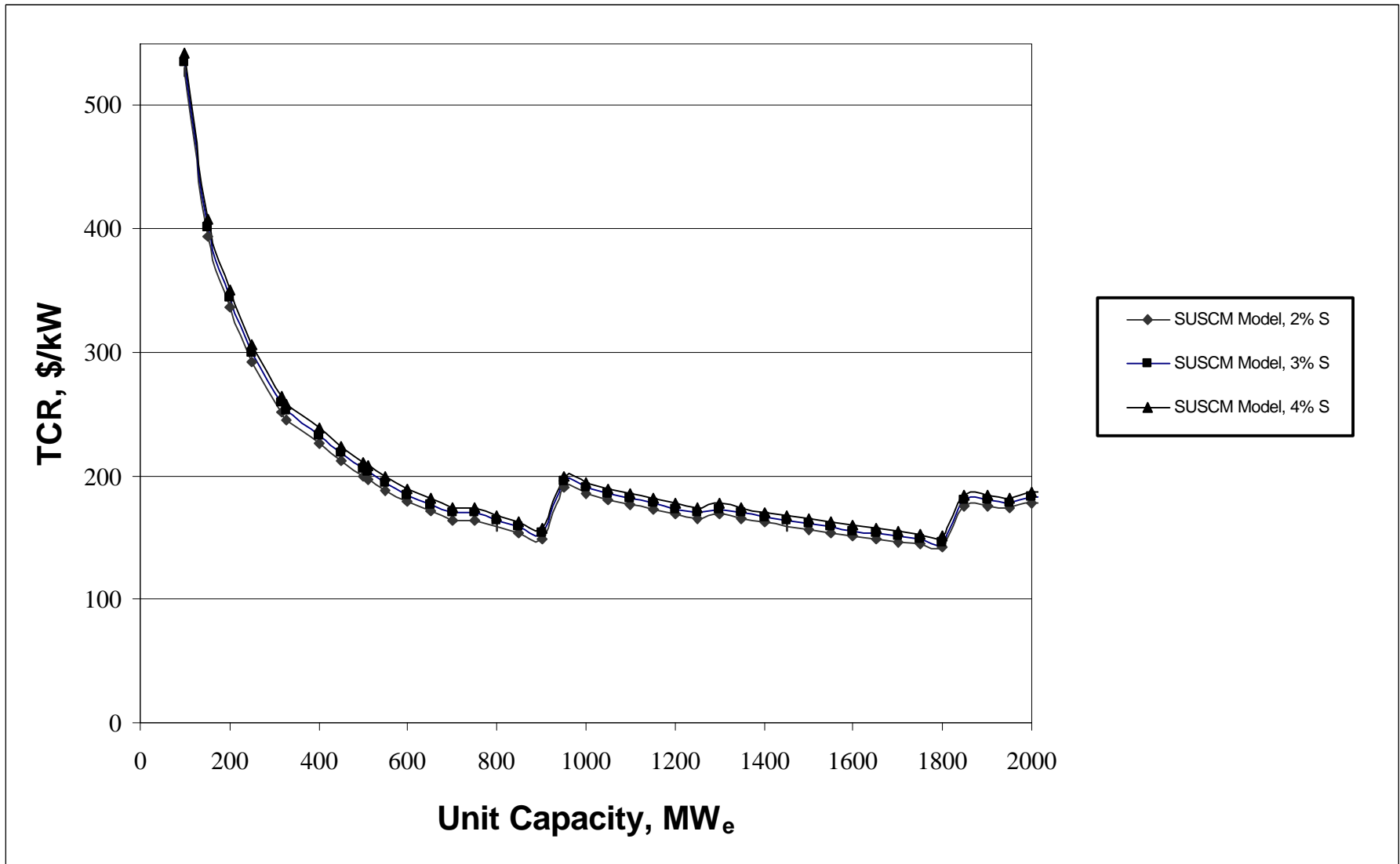


Figure 6-4. TCR predictions for 2 to 4 percent sulfur coal by LSFO SUSCM.

For comparison, the published average cost of 24 Phase I units⁶⁶ was \$249/kW (1995 dollars) or \$241/kW when de-escalated to 1994 dollars. Significant cost reductions may be realized by employing “state-of-the-art” design. For example, the LSFO SUSCM predicts a LSFO TCR of \$211/kW for a 500 MW_e system with 4 percent sulfur coal. For the same conditions, the simplified LSFO model predicted a TCR of \$229/kW.

Setting the LSFO SUSCM parameters to values representative of conditions at Big Bend Station resulted in a predicted TCR of \$153/kW (with the TPC of \$107 million). Further, giving the credit for the effect of high velocity in the absorber, TCR decreases to \$145/kW.

As described earlier, fixed O&M was a function of the installed BM cost and the unit capacity (MW_e). The LSFO SUSCM prediction of fixed O&M for a unit with a heat rate of 10,500 Btu/kWh is shown in

Figure 6-5. The fixed O&M cost is based on capital cost and, therefore, reflects the same trends as capital cost. The LSFO SUSCM prediction for Big Bend Station’s fixed O&M is \$6/kW-year.

As can be seen in Figures 6-4 and 6-5, LSFO SUSCM predictions of TCR and of fixed O&M are not very sensitive to coal sulfur content in the range of 2 to 4 percent.

Variable O&M is a function of the sulfur input and power requirements, adjusted for capacity factor. The LSFO SUSCM prediction of variable O&M for a unit with a 10,500 Btu/kWh heat rate and 90 percent capacity factor is shown in Figure 6-6. Variable O&M costs on a mills/kWh basis are constant across the unit capacity range and increase with fuel sulfur content. The LSFO SUSCM prediction for Big Bend Station’s variable O&M is 1.37 mills/kWh.

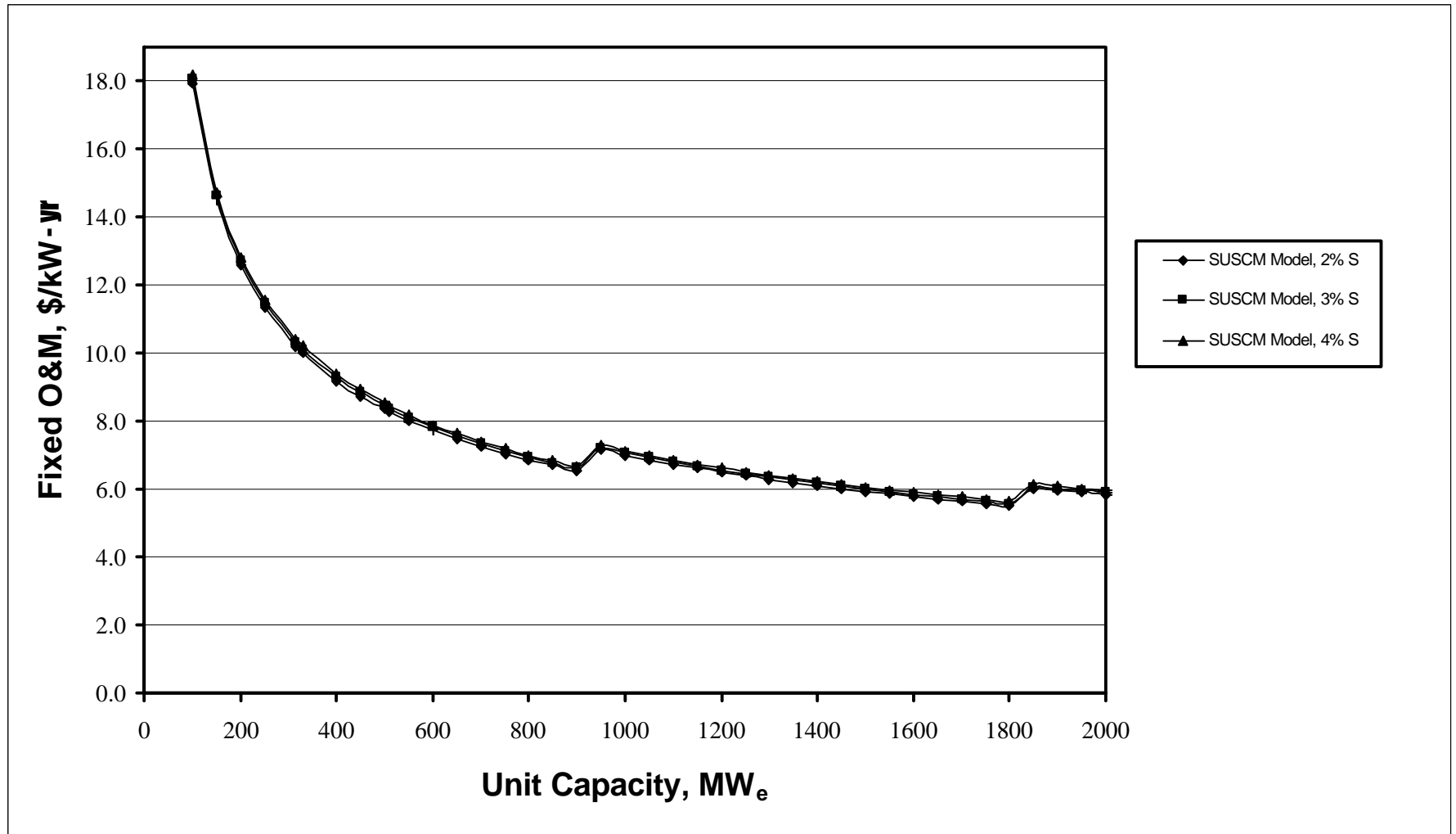


Figure 6-5. Fixed O&M predictions for 2 to 4 percent sulfur coal by LSFO SUSCM.

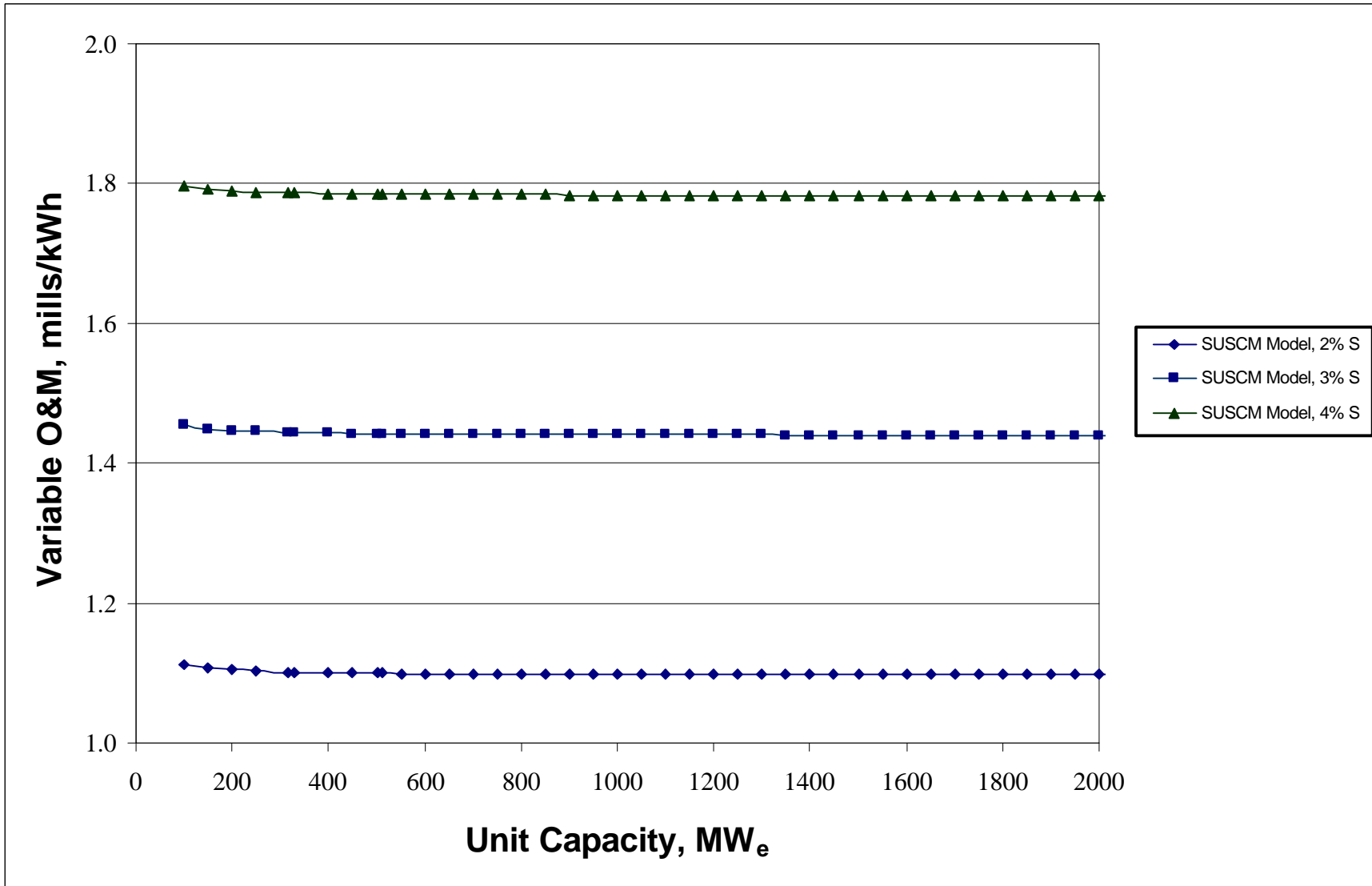


Figure 6-6. Variable O&M predictions for 2 to 4 percent sulfur coal by LSFO SUSCM.

Lime Spray Drying

Sensitivity Analysis

Sensitivity analyses were performed to determine variables that have relatively minor impacts on FGD cost. The objective of these analyses was to build an order of magnitude cost estimate model using commonly available parameters that significantly affect cost.

For the sensitivity analyses, it was necessary to identify a baseline LSD system as a point of reference. A 500-MW_e unit with a 10,500 Btu/kWh heat rate burning 1.5 percent sulfur coal was selected as the baseline unit.

The primary design elements fixed in this baseline LSD system were the spray dryer absorber construction materials and stack construction. RLCS was selected as the construction material for the baseline unit. The baseline LSD system uses two absorbers per CUECost methodology (maximum absorber size 300 MW_e). Other variables were fixed at CUECost default values for the baseline LSD, including 90 percent SO₂ removal efficiency. Thus defined, the baseline LSD has an annual operating cost of 10.02 mills/kWh.

Results of the sensitivity analyses are summarized in Table 6-7. Values for the variables were selected to span realistic ranges. High and low values of variables were selected and the corresponding cost was then determined for each single variable perturbation. Next, the differences in cost predictions were calculated between baseline and high, as well as low, values for each perturbed variable.

Based on the sensitivity analyses, it appears that the majority of cost impacts can be

accounted for with capacity, heat rate, coal sulfur content, and coal heating value.

By fixing variables that have minor impacts on the cost, the methodology can be reduced to a function of just a few variables. The variables that have minor impacts on the cost, as predicted by CUECost sensitivity analyses, as well as the respective fixed values, are shown in Table 6-8. These fixed values are based on the baseline CUECost case.

Capital Cost

Similarly to LSFO, installed capital cost (BM cost) for LSD is calculated for each of five major equipment areas. The estimation methods used for the five major equipment areas are described below.

Reagent Feed Area

Reagent Feed Area cost (including receiving, storing, and slaking) - addition of a linear component based on the design lime addition rate (lb/h) and a power law component based on fresh lime slurry feed rate (gpm). The lime addition rate was determined by the uncontrolled SO₂ emission rate and the coal sulfur content. Fresh lime slurry feed rate was calculated for the lime addition rate at 30 percent solids, 1.3 specific gravity, and 90 percent lime purity.

The Reagent Feed Area cost (BM_F) was estimated based on the lime feed rate. Lime purity has been fixed at 90 percent CaO, which was used as the default composition in CUECost. The cost estimate was then calculated for the coal sulfur content, which

Table 6-7. Sensitivity Analysis of LSD Annual Operating Cost (baseline value of 10.02 mills/kWh)

Variable, units	Baseline	Variable's High Value	Variable's Low Value	Cost for High Value of Variable, mills/kWh	Cost for Low Value of Variable, mills/kWh	High Value Difference, ^a %	Low Value Difference, ^b %
Capacity, MW _e	500	2000	100	4.76	18.77	52.5	87.3
Heat Rate, Btu/kWh	10,500	11,000	8,000	10.29	8.58	2.7	-14.4
Coal Sulfur Content, %	1.5	2.00	1.00	11.16	8.86	11.4	-11.6
Coal Heating Value, Btu/lb	11,922	14,000	10,500	9.14	10.78	-8.8	7.6
Air Heater Outlet, °F	300	360	280	10.10	9.99	0.8	-0.3
SO ₂ Removal, %	90	95	85	10.16	9.88	1.4	-1.4
Adiabatic Saturation Temp, °F	127	145	110	10.04	10.00	0.2	-0.2
Approach to Saturation, °F	20	50	10	10.05	10.01	0.3	-0.1
Recycle Slurry Solids, %	35	50	10	10.01	10.10	-0.1	0.8
# of Absorbers	2	3	N/T ^c	10.57	N/T	5.5	N/T
Absorber Material	RLCS	Alloy	N/T	10.69	N/T	6.7	N/T

^a(Cost for High Value of Variable – 10.02) / 10.02 • 100%

^b(Cost for Low Value of Variable – 10.02) / 10.02 • 100%

^cN/T = not tested

Table 6-8. Representative Values for LSD Variables with Minor Cost Impacts

Variable	Units	Value	Comments
Coal Heating Value	Btu/lb	11,900	Baseline 11,922
Lime Purity	% CaO	90.0	
SO ₂ Control Efficiency	%	90.0	
Ambient Pressure	in. Hg	29.4	
Air Heater Outlet Temperature	°F	300	
Moisture in the Flue Gas	%	6.0	Before control device
		14.0	After control device
Approach to Saturation	°F	20	
Adiabatic Saturation Temperature	°F	127	
Recycle Slurry Solids	%	35	

determined the stoichiometric ratio (1.75 taken for 3.43 percent S coal), and by the maximum feed rate to the FGD system. As described earlier, the heating value was fixed at 11,900 Btu/lb in this model. The SO₂ flow rate can be estimated based on the coal sulfur content, unit capacity [MW_e], and heat rate [Btu/kWh] as follows:

$$FR_{SO_2} = \frac{Wt\%S \cdot 1000}{HHV} \cdot \left[\frac{64}{32} \right] \cdot MW_e \cdot HR \quad (6-36)$$

where Wt%S is coal sulfur content (wt%) MW_e is LSD size, HR is plant heat rate (Btu/kWh), and HHV is coal heating value (Btu/lb).

Once the SO₂ flow rate is known, the Reagent Feed Area cost (BM_F) may be estimated as follows:

$$BM_F = \left(170023 \cdot \frac{FR_L}{1000} + 3764611 \right) + (72338 \cdot GPM^{0.3195}) \quad (6-37)$$

where FR_L is the reagent feed rate:

$$FR_L = FR_{SO_2} \cdot 1.75 \cdot \frac{56}{64} + FR_{SO_2} \cdot 1.75 \cdot \frac{56}{64} \cdot \frac{1-0.9}{0.9} \quad (6-38)$$

and GPM is slurry flow rate:

$$F_{GPM} = \frac{FR_L \cdot \frac{74}{56} + FR_L \cdot \frac{74}{56} \cdot \frac{1-0.3}{0.3}}{8.34 \cdot \left(\frac{1+0.3}{60} \right)} \quad (6-39)$$

SO₂ Removal Area

SO₂ Removal Area cost (including spray dryers, tanks, and pumps) - third order polynomial based on coal sulfur content.

- Cost of spray dryers - second order polynomial based on actual gas flow rate entering each absorber [cfm] multiplied by the number of absorbers. Absorber size was limited

in CUECost to treat a maximum of 300 MW_e; larger units require multiple equal size absorbers.

The SO₂ Removal Area cost (BM_R) required estimation of flue gas flows and selection of absorber materials. Gas flow was calculated in a manner similar to that used for LSFO calculations to yield the flow as shown below:

$$ACFM = \frac{1000}{10^6} \cdot \frac{9780}{60} \cdot \frac{(460 + 295)}{528} \cdot \frac{100}{(100 - 6)} \cdot MW_e \cdot HR \cdot \left(\frac{0.04}{P} + \frac{0.209}{P} \cdot \frac{(P - 0.04)}{(0.209 - P)} \right) \quad (6-40)$$

The pressure at the absorber inlet was fixed at 12 in. H₂O vacuum (the CUECost default). Ambient pressure was fixed at the CUECost default of 29.4 in. Hg. Oxygen at 9.0 percent was assumed throughout the LSD. The moisture fraction was assumed to be 6 percent at the spray dryer inlet. The above assumptions allowed for the estimation of the SO₂ Removal Area cost (BM_R), as shown below.

$$BM_R = BARE\ MODULE_R + SPRAY\ DRYERS \quad (6-41)$$

- For the SO₂ Removal System with RLCS construction, the cost of spray dryers (SPRAY DRYERS1) was calculated as:

$$SPRAY\ DRYERS1 = \left(-3.57 \cdot \left(\frac{ACFM}{N_a \cdot 1000} \right)^2 + 9246 \cdot \left(\frac{ACFM}{N_a \cdot 1000} \right) + 791896 \right) \cdot N_a \quad (6-42)$$

- For the SO₂ Removal System with alloy construction, the cost of spray dryers (SPRAY DRYERS2) was calculated as:

$$SPRAY\ DRYERS2 = \left(-4.85 \cdot \left(\frac{ACFM}{N_a \cdot 1000} \right)^2 + 12538 \cdot \left(\frac{ACFM}{N_a \cdot 1000} \right) + 1080990 \right) \cdot N_a \quad (6-43)$$

where N_a is the number of absorbers.

Auxiliary cost (BARE MODULE_R) was calculated as:

$$BARE\ MODULE_R = \left(581877809 \cdot Wt\%S^3 - 3653117 \cdot Wt\%S^2 + 693335 \cdot Wt\%S + 214198 \right) \cdot N_a + 677421 \cdot Wt\%S^{-0.0966} \quad (6-44)$$

Flue Gas Handling Area

Flue Gas Handling Area cost (including ductwork and fans) - linear addition of power laws based on the actual flue gas flow rate entering the absorber, exiting the absorber, exiting the particulate control device, and exiting the ID fans.

- Cost of ID fans - power law based on the flue gas flow rate [cfm] handled by each fan multiplied by the number of fans required. The number of fans required was determined by the total gas flow rate and the maximum gas flow rate per fan (1,600,000 cfm).

The Flue Gas Handling Area cost (BM_G) was estimated based on flue gas flow rates at multiple locations: entering the absorber, exiting the absorber, exiting the particulate control device, and exiting the ID fans. The

flue gas exiting the absorber was assumed to be at 17 in. H₂O vacuum and 147 °F, consistent with a 20 °F approach to saturation. Flue gas exiting the particulate control device was assumed to be at 23 in. H₂O vacuum and 147 °F. Flue gas exiting the fans was assumed to be at 1 in. H₂O gauge positive pressure and at 152 °F. The CUECost model adjusts flue gas flow rates to account for water evaporation and acid gas removal. For flue gas flow estimating purposes, all flue gas flows after the absorber inlet had a water content of 14 percent.

The Flue Gas Handling Area cost included the cost of ID fans. It was estimated using the flue gas flow rate exiting the particulate control device and the number of fans required. CUECost determines the number of fans through a series of logical comparisons based on maximum individual fan capacity at the specified pressure change across the fan. The pressure differential across the fans was fixed at 24 in. H₂O.

Based on the assumptions presented above, the Flue Gas Handling Area cost (BM_G) was estimated as follows:

$$BM_G = \text{BARE MODULE}_G + \text{ID FANS} \quad (6-45)$$

The area's auxiliary cost (BARE MODULE_G) was estimated by the following equation:

$$\begin{aligned} \text{BARE MODULE}_G = & \left(1721.8 \cdot \left(\frac{\text{ACFM}}{1000} \right)^{0.683} + 1326.2 \cdot \left(\frac{\text{ACFM1}}{1000} \right)^{0.7131} \right) \cdot N_a \\ & + \left(15338 \cdot \left(\frac{\text{ACFM}}{1000} \right)^{0.5} + 47680 \cdot \left(\frac{\text{ACFM1}}{1000} \right)^{0.5576} \right) \\ & + \left(4840.4 \cdot \left(\frac{\text{ACFM2}}{1000} \right)^{0.5} + 2695.9 \cdot \left(\frac{\text{ACFM3}}{1000} \right)^{0.5} \right) \quad (6-46) \end{aligned}$$

where ACFM1, ACFM2, and ACFM3 are flue gas flow rates at the exit from the absorber, particulate control device, and ID fans, respectively. N_a is the number of absorbers.

The cost of ID fans (ID FANS) was calculated as:

$$\text{ID FANS} = 91.24 \cdot \left(\frac{\text{ACFM2}}{N_f} \right)^{0.6842} \cdot N_f \quad (6-47)$$

where N_f is the number of fans.

Waste/By-product Handling Area

Waste/By-product Handling Area cost (including disposal and storage) - second order polynomial based on coal sulfur content (Wt%S).

Waste Handling Area cost (BM_W) was estimated as a function of coal sulfur content. Waste included fly ash and was presumed to be sent to a landfill. BM_W was estimated as follows:

$$BM_W = 2051841884 \cdot \text{Wt\%S}^2 - 1443163 \cdot \text{Wt\%S} + 1026479 \quad (6-48)$$

Support Equipment Area

Support Equipment Area cost (including electrical, water, and air) - second order polynomial based on the unit capacity (MW_e).

- Cost of chimney - power law based on the flue gas flow rate (ACFM3) exiting the ID fans.

The Support Equipment Area cost (BM_E) included the chimney without reheat. The chimney cost (CHIMNEY) was based on the

flue gas flow rate and was estimated as follows:

$$CHIMNEY = 23370 \cdot ACFM^3^{0.3908} \quad (6-49)$$

Support Equipment Area cost (BM_E) was calculated as:

$$BM_E = -1.211 \cdot MW_e^2 + 2704.2 \cdot MW_e + 1354716.2 + CHIMNEY \quad (6-50)$$

Adding the BM cost components for the five major areas yields an estimate for installed capital cost.

Total Capital Requirement

Once the BM cost had been determined, it was possible to calculate LSD TCR. Total Plant Cost (TPC) was calculated in the same manner as explained before for LSFO in equation (6-24).

Next, TPC was adjusted for financial factors dependent on the time required to complete the project.

As explained before for LSFO, the adjustment results in Total Plant Investment (TPI) as described before in equation (6-25).

Since it is usually preferable to assume constant dollars in regulatory applications, a constant dollar analysis was done as explained before in the LSFO section.

Current dollar factors were used for validation, assuming that the published cost for TCR was in current dollars. Constant dollar factors were used in the subsequent model development.

TCR was determined by adding pre-production cost and inventory capital to TPI.

CUECost estimates pre-production cost at 2 percent of TPI plus one-twelfth of the projected annual O&M (fixed plus variable adjusted for capacity factor) cost. Similar to considerations for LSFO, a 60 day lime inventory was incorporated in the model. The default cost of lime used here was \$50/ton. Substituting the default factors in TPI and the default cost of lime yields TCR as described before by equation (6-26).

The CUECost-determined TCR for baseline conditions shown in Table 6-8 and for minor effect variables fixed as shown in Table 6-9 was equal to \$159/kW. However, when the minor effect variables were set to yield the highest cost, the resulting value of TCR was \$165/kW. Therefore, TCR was multiplied by the adjustment factor of 1.038 (165/159) to yield the Adjusted TCR.

Operation and Maintenance Cost

The O&M cost was calculated next. The O&M cost includes fixed and variable components. The fixed O&M cost incorporates:

- operating labor
- maintenance labor and materials
- administration and support labor

The variable O&M cost is composed of:

- reagent
- disposal
- fresh water
- energy

Fixed O&M cost components were estimated as follows. Operating labor (OL) was estimated by the equation below, using a power law on unit capacity and estimating number of workers needed in combination with an operating labor rate (\$30/hr):

$$OL = 18.25 - 2.278 \cdot MW_e \cdot \frac{\ln(MW_e) \cdot 30 \cdot 40 \cdot 52}{100} \quad (6-51)$$

The maintenance labor and materials (ML&M) cost was determined as a percentage (2 percent) of BM cost. Administration and support (A&S) labor was estimated as a fraction of maintenance labor and materials and of operating labor, as given by the equation:

$$A \& S = 0.3 \cdot (0.4 \cdot ML \& M + OL) \quad (6-52)$$

Variable O&M cost components were estimated as a sum of lime, disposal, fresh water, and energy costs. The cost of lime (unit price of lime at \$65/ton) was:

$$C_{CaO} = \frac{FR_L}{2000} \cdot 8760 \cdot CF \cdot 65 \quad (6-53)$$

where CF is capacity factor.

The cost of disposal (\$30/ton) is:

$$C_{DL} = \frac{8760}{2000} \cdot CF \cdot 30 \cdot \left(FR_{SO_2} \cdot \frac{129}{64} + MW_e \cdot 1000 \cdot 0.1 \cdot \frac{HR}{HHV} \right) \quad (6-54)$$

The cost of energy (energy consumption for LSD estimated at 0.7 percent) was estimated using the default CUECost energy price of 25 mills/kWh):

$$POWER = 0.007 \cdot \frac{(1000 \cdot MW_e)}{1000} \cdot 8760 \cdot CF \cdot 25 \quad (6-55)$$

As an annual expense, the components of variable O&M cost were adjusted for the capacity factor of the unit.

Validation

The 1995 EIA-767 browser database²⁸ on LSD systems installed in the 1980's has been used for validation. Six LSD systems were found in this database with adequate data to perform validation. However, costs provided for Stanton 1, East Bend 2, and Craig 3 units appeared unreasonably low for a FGD system of this type and were not considered during validation.

Due to the vintage of these LSD system costs, it was presumed in modeling that they were built with RLCS absorbers. Since spray dryers typically operate between 20 and 30 °F above the dewpoint, no reheat was assumed in these designs. Table 6-9 presents validation data for the LSD model. The results of validation are also shown in

Table 6-9. Validation of LSD Model

Plant/Unit	Unit Capacity, MW _e	Coal S, wt %	Number of absorbers	Reported Cost, \$/kW	Model Cost, \$/kW	Deviation, ^a %
H.L. Spurlock/2	508	3.6	4	189	222	17.5
Wyodak/1	362	0.8	3	172	203	18.0
North Valmy/2	267	0.5	3	231	205	-11.3

^aDeviation = (Model - Reported) / Reported • 100%

Figure 6-7. These results reflect that the simplified LSD cost model on average predicts the published capital cost within 15.6 percent.

State-of-the-art Model

The algorithms developed thus far incorporated a variety of adjustments to CUECost algorithms to eliminate variables that did not have a significant impact on cost. At this point, however, it is helpful to specify a "state-of-the-art" LSD system by which to measure the cost of possible future retrofits. It is recognized that alternate design decisions may be made in the interest of reducing cost based on site specific conditions, or other engineering advances, resulting in cost savings not reflected otherwise.

The model (LSD SUSCM) assumes use of the minimum number of absorbers possible based on the maximum size constraint of 275 MW_e.⁶⁷ The "state-of-the-art" LSD used in the LSD SUSCM incorporates a RLCS absorber construction, and a 30 day reagent

inventory. "State-of-the-art" LSD design decisions are shown in Table 6-10.

Combining the equations developed before with these "state-of-the-art" design decisions yields a LSD SUSCM-derived estimate of the TCR for a "state-of-the-art" FGD unit. TCR predictions using LSD SUSCM are shown in Figure 6-8.

As described earlier in this chapter, fixed O&M cost is a function of the installed BM cost and the unit capacity (MW_e). The LSD SUSCM prediction of fixed O&M cost for a unit with a heat rate of 10,500 Btu/kWh is shown in Figure 6-9.

The LSD SUSCM prediction of variable O&M cost for a unit with a 10,500 Btu/kWh heat rate and 90 percent capacity factor is shown in Figure 6-10. Variable O&M costs on a mills/kWh basis are constant across the unit capacity range and increase with fuel sulfur content.

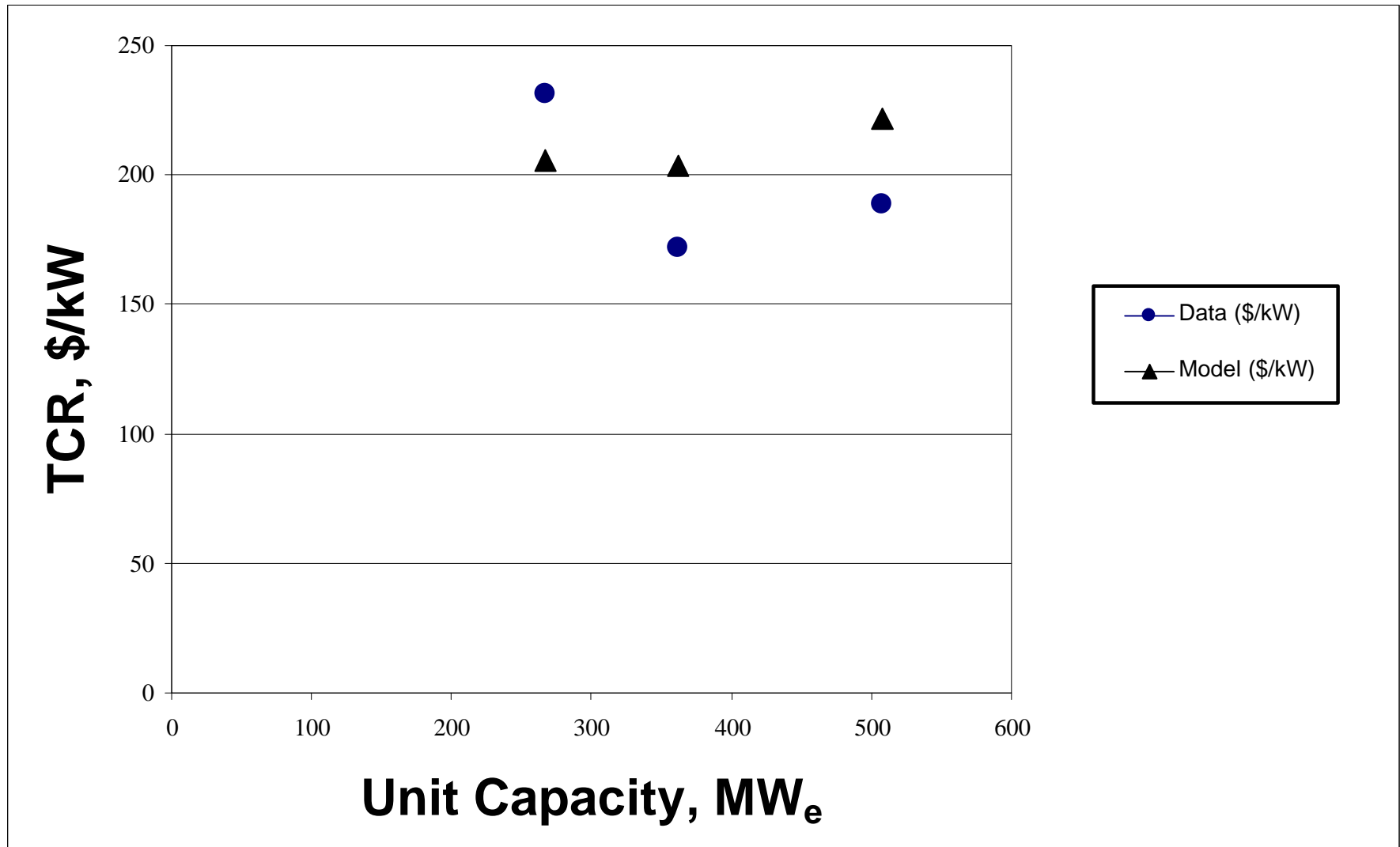


Figure 6-7. Validation of LSD cost model.

Table 6-10. “State-of-the-art” LSD Design Decisions

Parameter	Units	Value
Single Absorber Size	MW _e	275
O ₂ in Stack	%	8
Material of Construction	-----	RLCS ^a
SO ₂ Removal	%	90
Stoichiometry		1.4 for 2% S Coal
Flue Gas Temperature	°F	300
Lime Inventory	days	30
Lime Purity	%	94
Lime Cost	\$/ton	50
Waste Disposal Cost	\$/ton	12

^aRLCS = Rubber-lined Carbon Steel

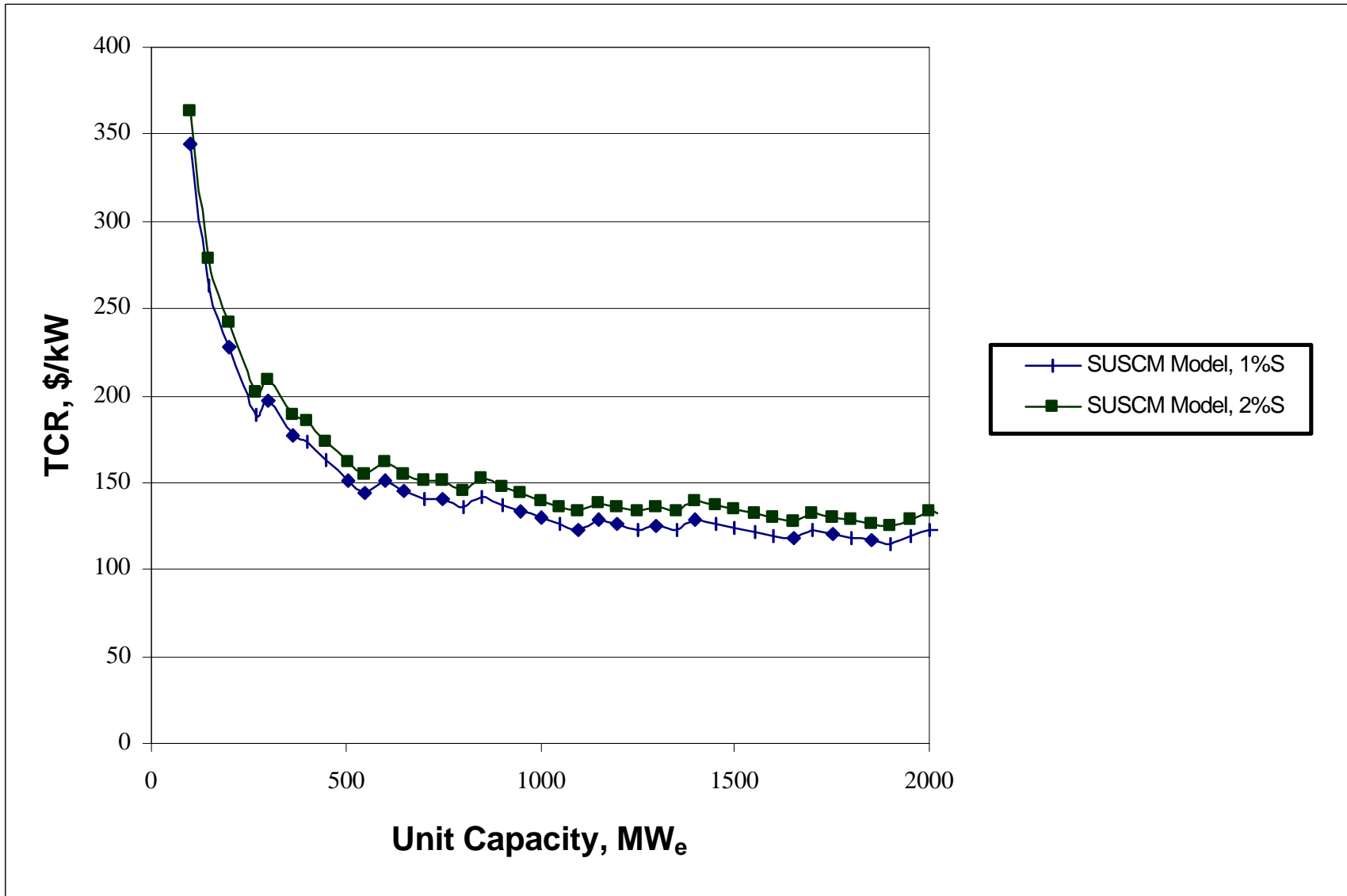


Figure 6-8. LSD TCR predictions by LSD SUSCM.

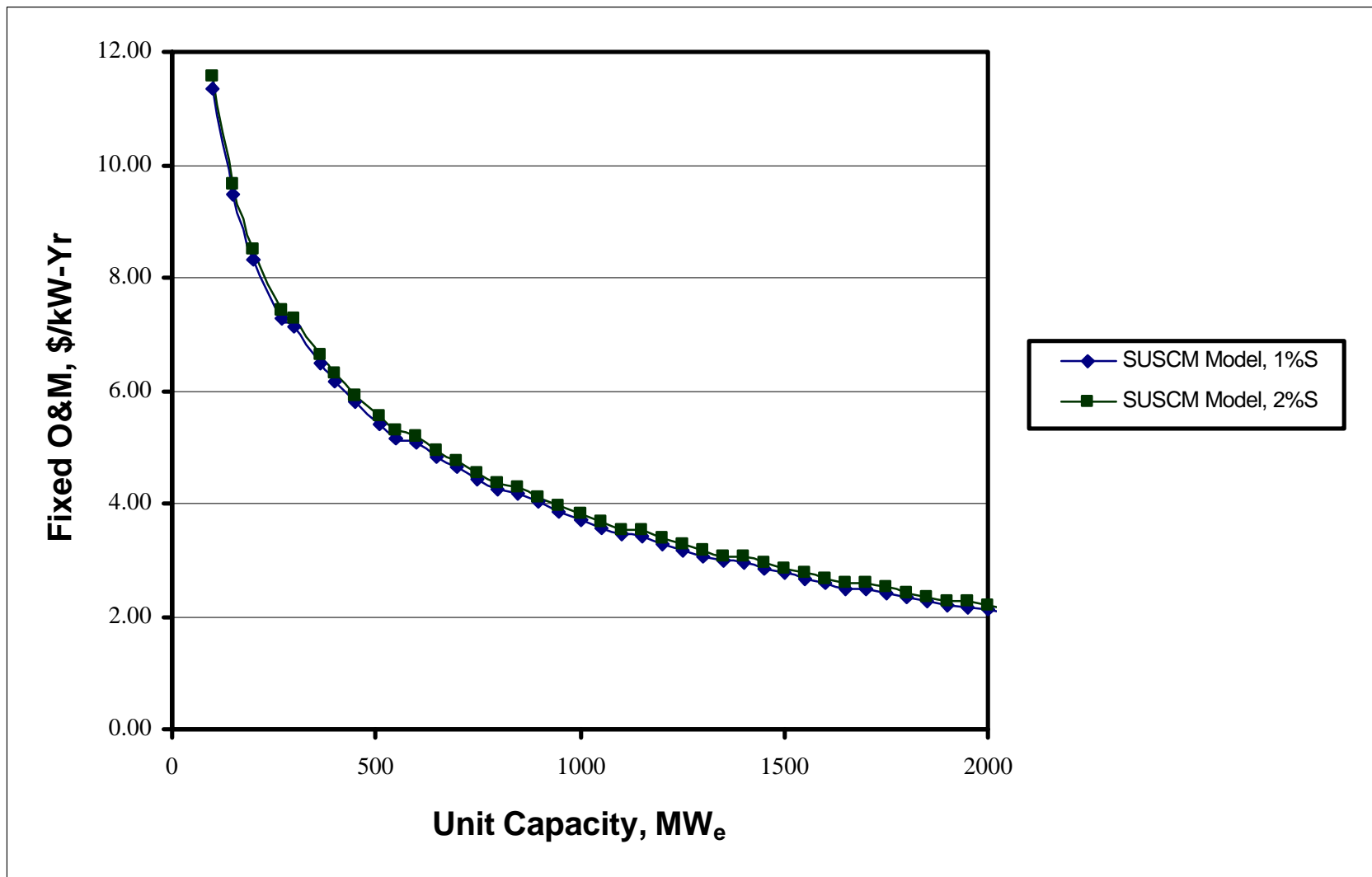


Figure 6-9. LSD fixed O&M predictions by LSD SUSCM.

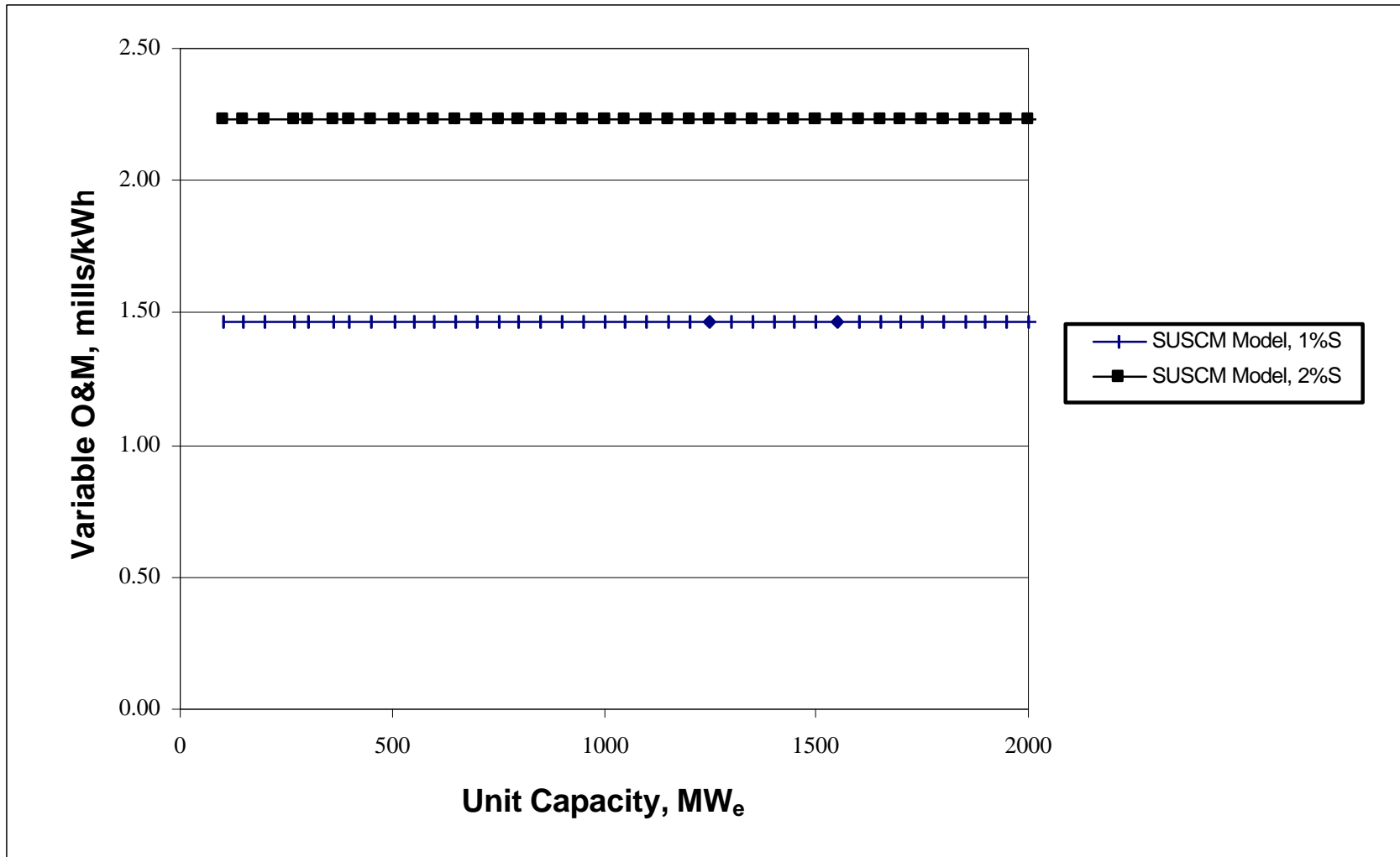


Figure 6-10. LSD variable O&M predictions by LSD SUSCM.

Magnesium-enhanced Lime

General Approach

The approach taken was to estimate the Magnesium-enhanced Lime (MEL) system cost, both capital and O&M, based on the estimation methods previously described for LSFO and LSD. As described earlier, for costing purposes, MEL can be considered to be a combination of LSFO and LSD. The MEL cost was based on a retrofit presenting a medium difficulty. The derived algorithm was then further simplified by making state-of-the-art design decisions to build a cost model. TCR was estimated in the same manner as previously described for LSFO and LSD.

Capital Cost

The BM was calculated for each of five major equipment areas, as described before for LSFO (Reagent Feed, SO₂ Removal, Flue Gas Handling, Waste Handling, and Support Equipment). Each major equipment area may have extraordinary items estimated apart from the rest of the equipment system. The estimation methods used for the five major equipment areas were as described below.

The Reagent Feed, SO₂ Removal, and Waste Handling Area cost estimates were explicitly determined by the SO₂ feed rate to the FGD system. This estimate is determined in CUECost by the coal sulfur content and coal use rate with no provision for sulfur retention in the ash. The higher heating value (HHV) of the coal was fixed at 11,900 Btu/lb. SO₂ feed rate to the FGD system was estimated as given before in equation (6-1).

Adding the BM cost components from the five major systems yields an estimate for the MEL installed capital cost.

Reagent Feed Area

The Reagent Feed Area (BM_F) cost (including receiving, storage, and slaking of magnesium enhanced lime) was estimated using the same methodology as the one used before for the LSD reagent feed area. The reagent feed ratio remained constant with respect to coal sulfur content.

The BM_F was estimated based on lime feed rate. Lime purity has been fixed in this model at 94 percent CaO. Lime addition rate was fixed in this model at a 1.00 reagent feed ratio. These parameters allowed the BM_F cost to be estimated as follows:

$$BM_F = 170023 \cdot \frac{FR_L}{1000} + 3764611 + 72338 \cdot F_{GPM}^{0.3195} \quad (6-56)$$

where FR_L is reagent feed rate (lb/hr) and F_{GPM} is slurry flow rate (gpm).

SO₂ Removal Area

The SO₂ Removal Area (BM_R) cost (including absorber and spray pumps) of the MEL system is expected to require nominally the same size and number of tanks as the LSFO. This system's cost was estimated as a third order polynomial on SO₂ rate to the scrubber. The cost components were calculated as follows:

- Cost of absorber - Estimated at 90 percent of the cost of LSFO absorbers to approximate the reduction in height and elimination of spray headers for the MEL system. The cost estimate was based on a power law with the absorber inlet flow rate to each absorber multiplied by the number of absorbers. Separate power laws were used depending on the absorber construction materials. Maximum

absorber size was limited to 275 MW_e; larger units require multiple, equal size absorbers.

- Cost of spray pumps - the same methodology as previously employed to estimate LSFO spray pump cost was applied to MEL (a power law applied to the slurry flow rate per absorber per pump multiplied by the number of pumps). The slurry flow rate (gpm) was calculated based on the gas flow rate per absorber at the exhaust temperature, but at 1 in. H₂O less than the inlet pressure. L/G was fixed at 40 consistent with the open tower design and 95 percent SO₂ removal.

The BM_R cost estimation required calculation of the flue gas flow through the FGD system. Tank cost was estimated on the same basis as the one used for LSFO. Absorber cost was estimated based on inlet flue gas flow rate and construction materials. Spray pump cost was estimated based on gas flow rates exhausting the absorber.

The flue gas flow rate was calculated in the same manner as previously explained for LSFO and LSD. Pressure at the absorber inlet was fixed at 7 in. H₂O gauge, the CUECost default. Ambient pressure was fixed at the CUECost default of 29.4 in. Hg. Temperature of the flue gas entering the absorber may vary significantly for different units but is expected to have minimal impact on TCR, based on the sensitivity analysis for the LSFO. Absorber inlet temperature was fixed in the model at 295 °F, resulting from the 300 °F air heater outlet temperature used as the default in CUECost. Oxygen at 9.0 percent was assumed at the absorber inlet.

The moisture was assumed to be 6.0 percent at the absorber inlet.

The cost of spray pumps for the absorbers was estimated based on the absorber outlet flow rate and the number of pumps required. The number of pumps (n_p) required was based on the required slurry flow rate per absorber, and a maximum pump capacity (43,000 gpm, the same as for LSFO). The required slurry flow rate was determined by the L/G, estimated at 40 for 95 percent SO₂ removal in an open spray tower. The gas flow rate was determined at 127 °F and at 6 in. H₂O gauge. Moisture content was estimated at 14 percent.

These approximations allowed estimation of the BM_R cost depending on the material of construction for the absorber as follows:

$$BM_R = \text{BARE MODULE}_R + \text{ABSORBER} + \text{PUMPS} \quad (6-57)$$

- For SO₂ Removal Area with alloy absorber construction:

$$\text{ABSORBER}_2 = 230064 \cdot 0.9 \cdot \left(\frac{\text{ACFM}}{1000} \right)^{0.5638} \cdot N_a \quad (6-58)$$

where ACFM is flue gas flow at the absorber inlet in cfm and N_a is the number of absorbers.

- For SO₂ Removal Area with RLCS absorber construction:

$$\text{ABSORBER}_1 = 173978 \cdot 0.9 \cdot \left(\frac{\text{ACFM}}{1000} \right)^{0.5575} \cdot N_a \quad (6-59)$$

The cost of pumps (PUMPS) was calculated as follows:

$$PUMPS = 910.85 \cdot \left(\frac{F_{GPM}}{N_p} \right)^{0.5954} \cdot N_p \quad (6-60)$$

where F_{GPM} is slurry flow rate in gpm and N_p is the number of pumps.

The area auxiliary cost was estimated as follows:

$$BARE\ MODULE_R = 0.825 \cdot \left(\begin{array}{l} 0.8701 \cdot \left(\frac{FR_{SO_2}}{1000} \right)^3 - 188.2 \cdot \left(\frac{FR_{SO_2}}{1000} \right)^2 \\ + 34809 \cdot \left(\frac{FR_{SO_2}}{1000} \right) + 1905302 \end{array} \right) \quad (6-61)$$

Flue Gas Handling Area

The Flue Gas Handling Area (BM_G) cost (including ID fans) – MEL was assumed to have the same flue gas handling requirements as LSFO. Therefore, cost was estimated with the same methodology (a polynomial on gas flow rate entering absorbers, exiting absorbers, and the number of absorbers).

The BM_G cost was based on the number of absorbers, flow entering absorbers (ACFM), and flow exiting absorbers (ACFM1). Pressure of the gas exiting the absorbers was fixed at 4 in. H_2O gauge. The temperature of the gas exiting the absorbers was fixed at 127 °F, the CUECost default wet bulb temperature. Flue gas moisture content was approximated at 14 percent at the absorber outlet and through the remainder of the FGD system.

The cost of the fans was estimated by a power law based on the number of fans required and the flue gas flow rate. Fans were assumed to be installed in groups of 2, 4, or 8 with a maximum fan capacity of 1,600,000 cfm. The number of fans was based on conditional tests of the smallest number option (2, 4, or 8) resulting in an individual fan capacity of less than 1,600,000 cfm. Inlet pressure for sizing fans was fixed in the model at 12 in. H_2O vacuum. Temperature at the fan inlet was fixed in the model at 295 °F.

By fixing these design criteria, the BM_G cost was estimated as follows:

$$BM_G = BARE\ MODULE_G + ID\ FANS \quad (6-62)$$

where area auxiliary cost ($BARE\ MODULE_G$) was:

$$BARE\ MODULE_G = -0.1195 \cdot \left(\frac{ACFM}{1000} \right)^2 + 777.76 \cdot \left(\frac{ACFM}{1000} \right) + 238203 - 0.2009 \cdot \left(\frac{ACFM1}{1000 \cdot N_a} \right)^2 + 1266.4 \cdot \left(\frac{ACFM1}{1000 \cdot N_a} \right) + 420141 + 0.000012 \cdot \left(\frac{ACFM}{1000} \right)^3 - 0.1651 \cdot \left(\frac{ACFM}{1000} \right)^2 + 1288.82 \cdot \left(\frac{ACFM}{1000} \right) + 559693 \quad (6-63)$$

and cost of fans (FANS) was:

$$FANS = 91.24 \cdot \left(\frac{ACFM}{N_f} \right)^{0.6842} \cdot N_f \quad (6-64)$$

where N_f is the number of fans.

Waste/By-product Handling Area

The Waste/By-product Handling Area (BM_w) cost (including thickener and stabilization equipment) for the MEL waste handling area was based on LSFO landfill disposal cost (a third order polynomial on the SO_2 mass flow rate). If forced oxidation is employed, system cost would be equivalent to LSFO gypsum stacking or wallboard by-product options, also estimated as polynomials based on SO_2 mass flow rate. The cost of equipment components in this model was calculated as follows:

- Cost of thickener - estimated with the same method as the one used for LSFO thickener. Thickener was estimated as a linear function of waste solids removal rate.
- Cost of stabilization equipment included a lime bin, ash bin, and small pugmill to the waste handling system in addition to components used in the LSFO algorithm. This additional cost was included because, for natural oxidation, waste must be mixed with lime and fly ash prior to landfilling. Equipment cost estimates for this additional equipment were based on a fraction of Waste Handling Area cost, including the thickener.

Waste/By-product Handling Area cost (BM_w) was fixed by the disposal option chosen and by the amount of sludge to be disposed of. For MEL under natural oxidation, landfill disposal is the method used by most installations. This procedure requires similar equipment as LSFO for landfill disposal but is sized differently to account for the more difficult dewatering characteristics of the MEL waste. The LSFO Waste Handling Area, excluding the thickener, was presumed to be dominated by filter cost. This model

assumed 20 percent higher cost based on the SO_2 flow rate compared to LSFO system.

The thickener cost was estimated for LSFO as a linear function of dry waste disposal rates. This is consistent with basing cost on the surface system of the thickener. MEL wastes from a natural oxidation process require significantly more surface system per pound of waste than gypsum wastes due to slower settling rates. Magnesium salts are expected to remain in solution and do not affect settling rates. The amount of particulate waste was based on inlet SO_2 flow rate, removal efficiency (fixed at 95 percent), a reagent feed ratio of 1.05 based on CaO, and an estimated 5 percent inerts in the lime. For waste handling cost estimation purposes, all SO_2 removed was assumed to precipitate as calcium sulfite hemihydrate.

In addition to the waste handling equipment estimated by CUECost for LSFO, lime and flyash bins and a pugmill are required. The total cost of this equipment was estimated at 10 percent of the waste handling system cost, including the thickener.

BM_w cost was estimated as follows:

$$BM_w = BARE\ MODULE_w + THICKENER + D \& P \quad (6-65)$$

where:

$$BARE\ MODULE_w = \left(0.325 \cdot \left(\frac{FR_{SO_2}}{1000} \right)^3 - 168.77 \cdot \left(\frac{FR_{SO_2}}{1000} \right)^2 \right) \cdot 1.25 + 29091 \cdot \left(\frac{FR_{SO_2}}{1000} \right) + 773243 \quad (6-66)$$

and

$$THICKENER = 9018.7 \cdot FR_{SO_2} \cdot 0.95 \cdot \frac{172}{64 \cdot 2000} + 114562 \quad (6-67)$$

Bin and pugmill cost (D&P) was 10 percent of Waste Handling Area.

Support Equipment Area

Support Equipment Area (BM_E) cost, including the chimney, was estimated with a third order polynomial. The cost of the chimney was estimated based on total gas flow exiting each absorber.

Support Equipment Area cost (BM_E) was estimated as follows:

$$BM_E = BARE\ MODULE + CHIMNEY \quad (6-68)$$

where:

$$BARE\ MODULE = 0.825 \cdot \left(\begin{array}{l} 0.0003 \cdot MW_e^3 - 1.0667 \cdot MW_e^2 \\ + 1993.8 \cdot MW_e + 1177674 \end{array} \right) \quad (6-69)$$

The chimney cost was estimated with a power law based on flow rate per absorber in the same manner as for LSFO. Temperature at the chimney inlet was fixed in the model at 127 °F, while the pressure was fixed at 4 in. H₂O gauge:

$$CHIMNEY = 23370 \cdot ACFM1^{0.3908} \quad (6-70)$$

Total Capital Requirement

Once the BM cost was determined, it was possible to calculate TCR. First, Total Plant Cost (TPC) was estimated in the same manner as previously described for LSFO and LSD in equation (6-24).

Next, TPC was adjusted for financial factors depending on the time required to complete the project. Applying the TCE and FDC factors appropriate to the unit size (as explained previously) results in Total Plant Investment (TPI) as shown before.

In regulatory applications, it is usually preferable to assume constant dollars; e.g., no inflation. Therefore, constant dollars were used in the subsequent model development.

Finally, the Total Capital Requirement (TCR) was determined in the manner described earlier in this chapter for LSFO. The cost of lime of \$50/ton was used.⁶⁸ This lime typically contains 5 percent MgO. Substituting the default factors in TPI and the default cost of lime yielded a TCR prediction.

Operation and Maintenance Cost

O&M cost was calculated next. O&M cost includes fixed and variable components. Fixed O&M cost incorporates:

- operating labor
- maintenance labor and materials
- administration and support labor

Variable O&M cost is composed of:

- reagent
- disposal (by-product credit given)
- energy

Fixed O&M cost components were estimated as follows. Operating labor (OL) cost was estimated by the equation below, using a power law on unit capacity and estimating the number of workers needed in combination with an operating labor rate (\$30/hr):

$$OL = 41.69041 \cdot MW_e^{-0.322307} \frac{MW_e}{100 \cdot 30 \cdot 40 \cdot 52} \quad (6-71)$$

Maintenance labor and materials (ML&M) cost was determined as a percentage (3 percent) of BM cost. Administration and support (A&S) labor was estimated from maintenance labor and materials and operating labor as given by the equation below:

$$A \& S = 0.3 \cdot (0.4 \cdot ML \& M + OL) \quad (6-72)$$

The variable O&M cost component was estimated as the sum of lime, disposal, and energy costs. The cost of lime (unit price of lime at \$50/ton) was:

$$C_{CaO} = \frac{FR_L}{2000} \cdot 8760 \cdot CF \cdot 50 \quad (6-73)$$

where CF is the capacity factor.

The cost of disposal if gypsum stacking method is selected (\$6/ton) was:

$$C_{DS} = 6 \cdot 8760 \cdot CF \cdot FR_{SO_2} \cdot 0.95 \cdot \frac{129}{64 \cdot 2000} \quad (6-74)$$

The cost of disposal for landfill (\$30/ton) was:

$$C_{DL} = 30 \cdot 8760 \cdot CF \cdot FR_{SO_2} \cdot 0.95 \cdot \frac{129}{64 \cdot 2000} \quad (6-75)$$

The cost of energy (energy consumption for MEL estimated at 1.05 percent) was estimated using the default CUECost energy price of 25 mills/kWh):

$$POWER = 0.0105 \cdot \frac{(1000 \cdot MW_e \cdot 0.8231)}{1000} \cdot 8760 \cdot CF \cdot 25 \quad (6-76)$$

As an annual expense, the components of the variable O&M cost were adjusted for the capacity factor of the unit(s).

State-of-the-art Model

At this point, it is helpful to specify a "state-of-the-art" MEL system by which to measure the cost of possible future retrofits. Alternate design decisions may be made in the interest of reducing cost based on site specific conditions or other engineering advances resulting in cost savings not reflected in this model.

MEL SUSCM will assume use of the minimum number of absorbers possible, based on the maximum size constraint of 275 MW_e. The "state-of-the-art" MEL scrubber used in this model incorporates RLCS or alloy absorber construction and salable gypsum. "State-of-the-art" MEL design decisions are shown in Table 6-11.

Combining the equations developed earlier with these "state-of-the-art" design decisions yields a model description of a "state-of-the-art" MEL FGD system.

MEL SUSCM TCR predictions for MEL are shown in Figure 6-11 for 2, 3, and 4 percent S coals. These predictions are based on units with a heat rate of 10,500 Btu/kWh and a capacity factor of 90 percent. MEL SUSCM predictions reflect that capital cost is not sensitive to coal sulfur content.

Table 6-11. “State-of-the-art” MEL Design Decisions

Parameter	Units	Value
Single Absorber Size	MW _e	275
O ₂ in Stack	%	8
Material of Construction	-----	Average of RLSC and alloy
SO ₂ Removal	%	98
L/G	gal/1000 ft ³	40
Inventory for Lime	days	30
Lime Purity (CaO)	%	94
Sorbent Cost	\$/ton	50
Waste Disposal	-----	wallboard
Power Requirements	%	1.05
MEL/LSFO Capital Cost Ratio	-----	0.80-0.85
ID Fans Cost	\$	² / ₃ of LSFO ID Fans Cost

The fixed O&M cost prediction is shown in Figure 6-12. These costs are based on capital cost and, therefore, reflect the same trends as capital costs.

Variable O&M cost predictions by MEL SUSCM are shown in Figure 6-13. Variable O&M cost on a mills/kWh basis is constant across the unit capacity range and increases with fuel sulfur content.

Summary of FGD Cost

The comparison of capital and O&M costs for three technologies considered here is shown in Table 6-12. Ranges of costs are given in

1998 constant dollars for a 100 to 1000 MW_e unit. As can be seen in Table 6-12, capital cost for LSFO used on a small unit (100 MW_e) is considerably higher than capital cost of MEL used on the same size unit. For a large unit (1000 MW_e), capital cost is comparable for LSFO and for MEL.

Fixed O&M cost is similar for LSFO and MEL over the entire unit size range considered. However, variable O&M cost is lower for LSFO than for MEL, largely due to the difference in the sorbent cost (\$15/ton for LSFO versus \$50/ton for MEL).

Table 6-12. Cost in 1998 Constant Dollars for Selected FGD Technologies

Technology	Capacity Range ^a MW _e	Capital Cost, \$/kW	Fixed O&M, \$/kW-Yr	Variable O&M, mills/kWh
LSFO ^b	100 - 1000	542 – 195	18 – 7	1.80 – 1.78
LSD ^c	100 - 1000	363 – 140	12 - 4	2.24 – 2.24
MEL ^d	100 - 1000	384 – 238	16 – 8	2.02 – 2.01

^a Unit has a heat rate of 10,500 Btu/kWh and a capacity factor of 90 percent.

^b 4.0 percent sulfur coal application, SO₂ removal of 95 percent.

^c 2.0 percent sulfur coal application, SO₂ removal of 90 percent.

^d 4.0 percent sulfur coal application, SO₂ removal of 96 percent.

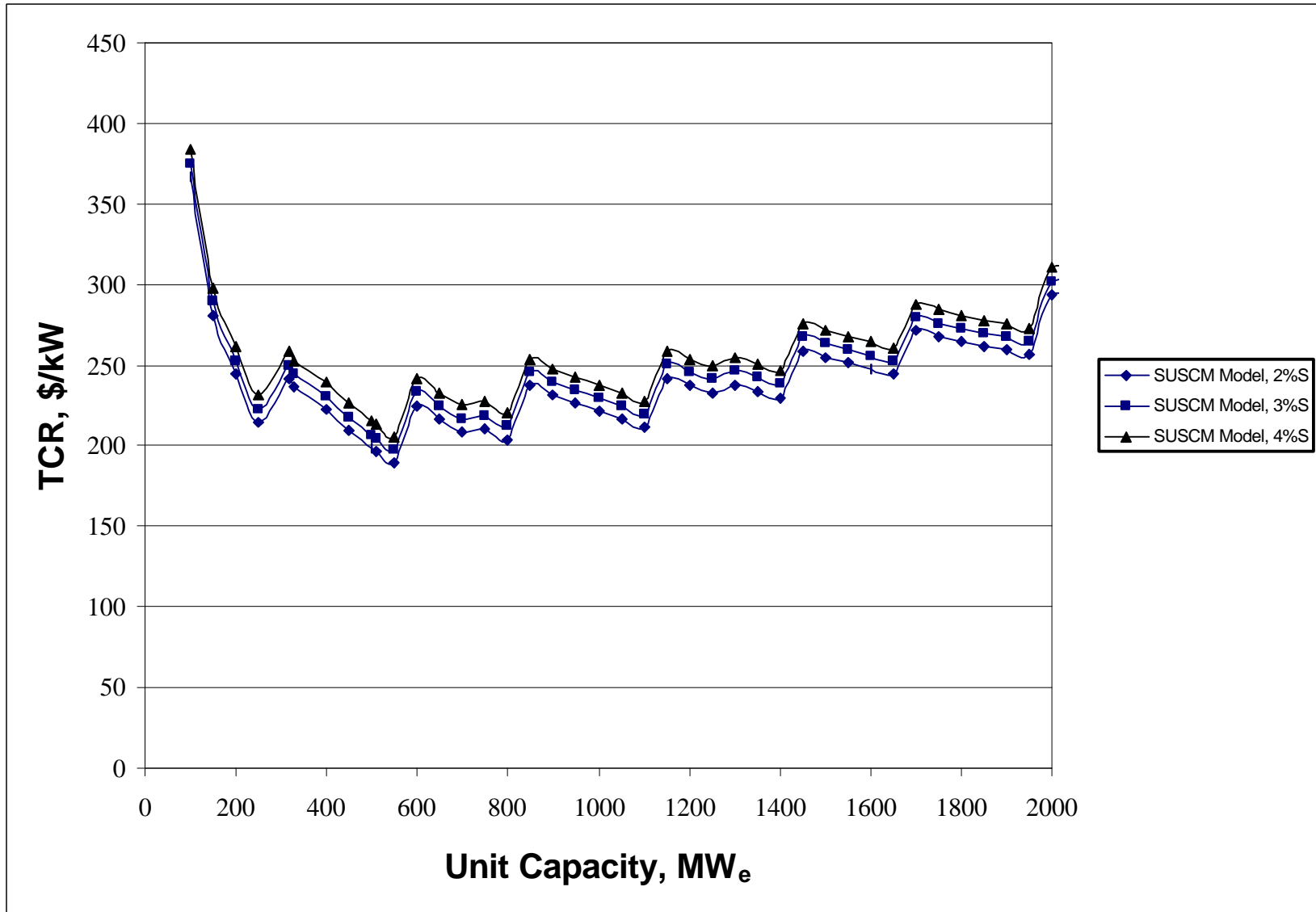


Figure 6-11. MEL TCR predictions by MEL SUSCM.

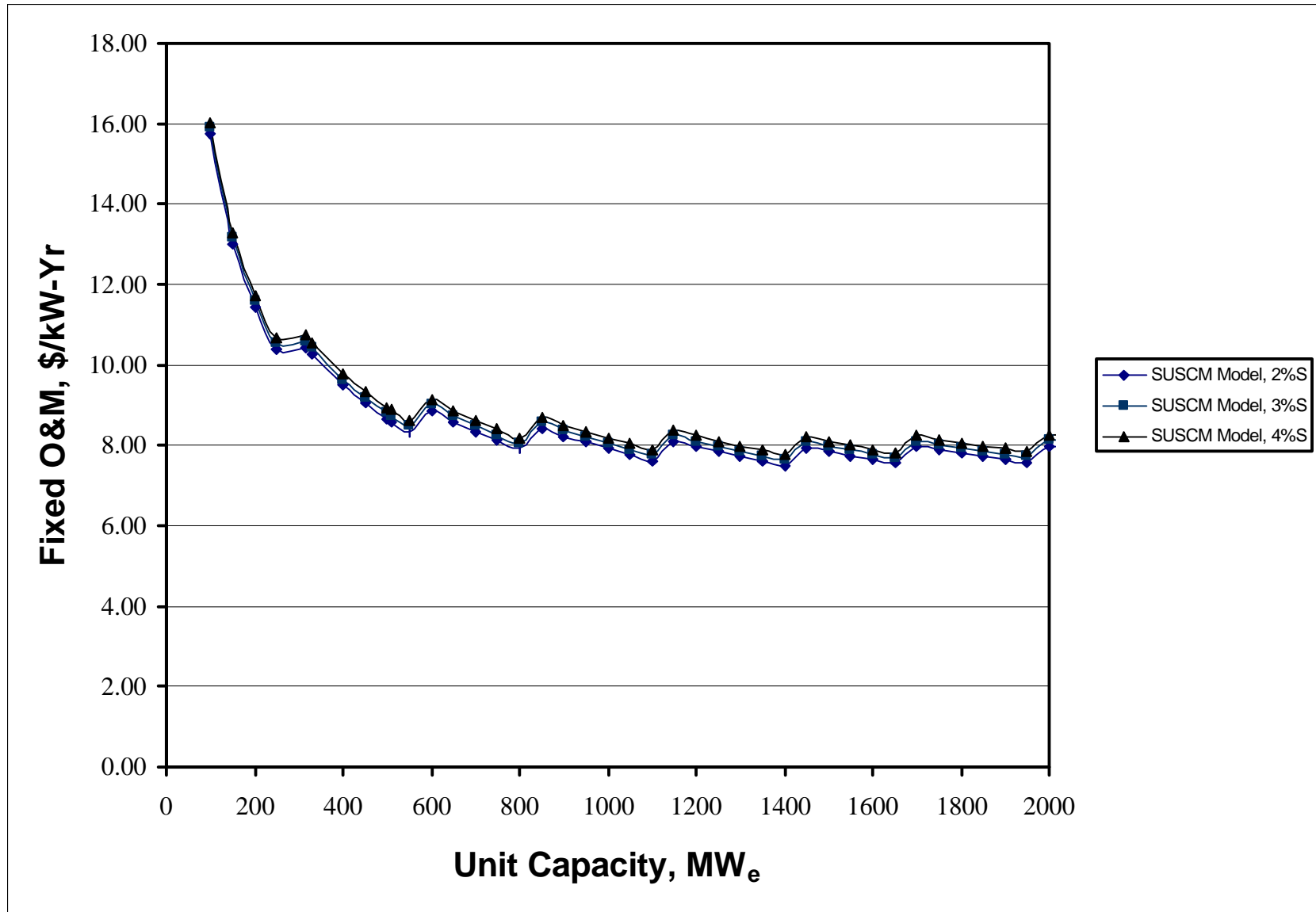


Figure 6-12. MEL fixed O&M predictions by MEL SUSCM.

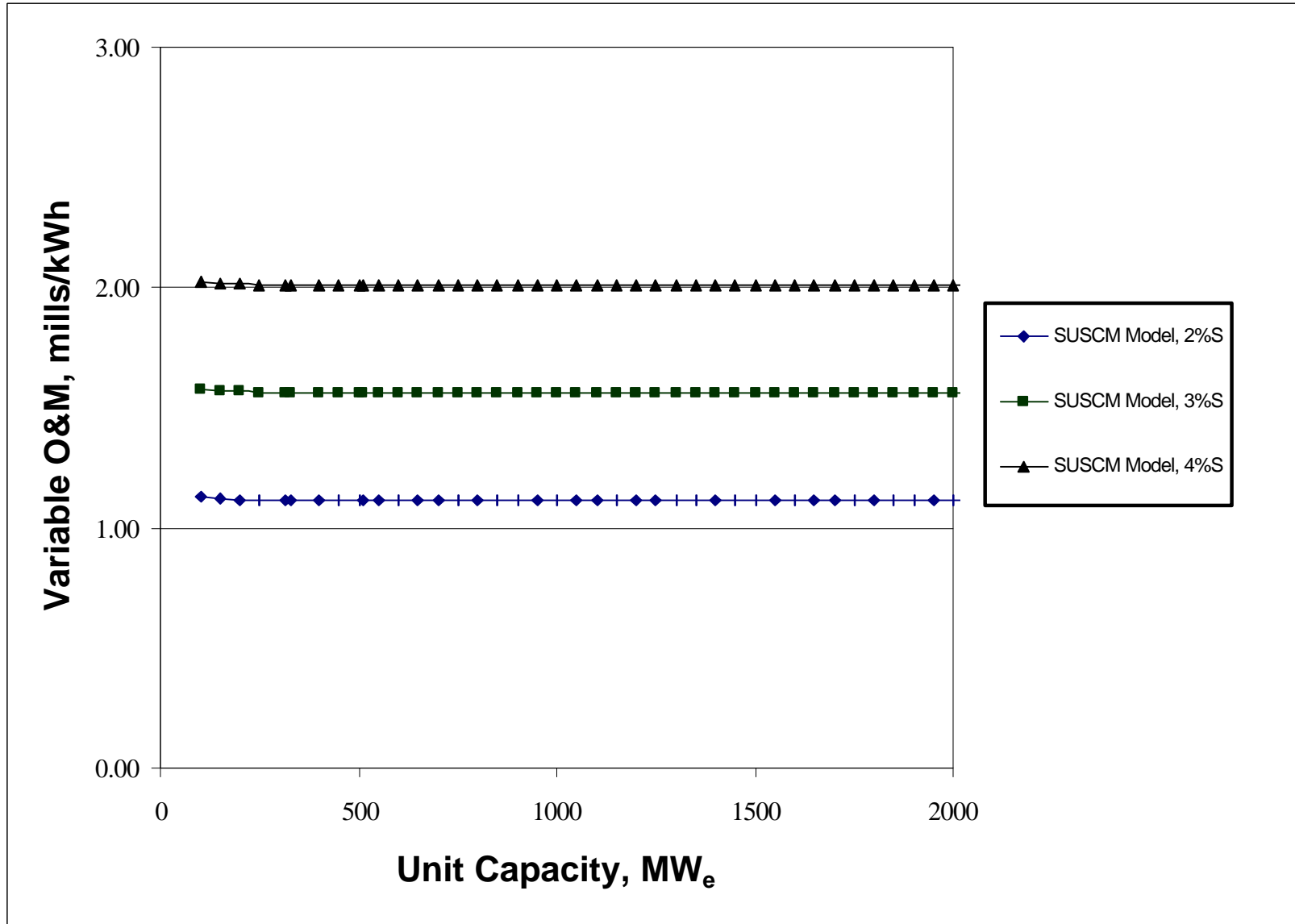


Figure 6-13. MEL variable O&M predictions by MEL SUSCM.

CHAPTER 7

ADDITIONAL BENEFITS

Introduction

The removal of mercury from flue gas by existing FGD processes could be viewed as an added benefit of controlling SO₂ emissions. Mercury emissions from coal-fired power generation sources are reported to be almost 33 percent of the total anthropogenic emissions in the U.S.⁶⁹ In coal-fired power generation, mercury is volatilized and converted to mercury vapor (Hg⁰) in the high temperature regions of combustion devices. Hg⁰ is transformed into oxidized mercury (Hg⁺⁺) as the flue gas cools. Therefore, the species predominantly present in flue gas include species of elemental Hg⁰ and Hg⁺⁺. It follows that control of both of these mercury species is necessary to achieve total mercury emission control.

At present, the control of mercury emissions from coal-fired boilers is not commercially practiced in the U.S. The combination of low mercury concentration and large flue gas volumes increases the difficulty and cost of controlling mercury emissions from coal-fired utility boilers compared to controlling mercury emissions from municipal waste

combustors.⁷⁰ However, numerous studies have been conducted that reported some level of mercury emission control by the existing FGD processes. The capability of existing FGD processes to remove mercury from coal-fired flue gas is affected by the mercury species present. Because of mercury's being the object of particularly strong concern due to its harmful effects on human health, the ability of the existing FGD processes to remove mercury from flue gas is discussed in more detail in the following sections of this chapter.

Another added benefit of controlling SO₂ emissions is the effect that decreased emissions of SO₂ have on the formation of fine particulate aerosols. July 1997 revisions to the National Ambient Air Quality Standards (NAAQS) place emphasis on particulate matter less than 2.5 μm in aerodynamic diameter (PM_{2.5}).³⁶ These aerosols are formed in the atmosphere in the presence of SO₂ and other gases. Therefore, an increased scrubber SO₂ removal efficiency, leading to lower SO₂ emissions, may decrease the amount of PM_{2.5}. Source emissions characterization is required to understand the fate of aerosol precursors (such as SO₂) in the particle formation process in the atmosphere.⁷¹ While PM_{2.5} can be produced directly by a variety of sources, it can also be produced by atmospheric reactions in the presence of SO₂, NO_x, and VOCs emitted from stationary sources.⁷² SO₂ is a precursor for sulfuric acid and sulfate secondary PM_{2.5} particles. Sulfate accounts for approximately 47 percent of PM_{2.5} in the eastern United States.⁷³ One strategy to control PM_{2.5} emissions from stationary coal-burning sources is to upgrade the existing particulate control device. The alternative route may be to control PM_{2.5} precursors, most notably SO₂. In this latter

case, modern, state-of-the-art SO₂ scrubbers, designed primarily for high efficiency control of SO₂, could provide an additional benefit by controlling PM_{2.5} precursors.

Once-through Wet FGD

A wide range of total mercury removal efficiency has been reported for once-through wet FGD applications on bituminous-coal-fired power generation units. Existing conventional wet scrubbers can remove water-soluble Hg⁺⁺ compounds (e.g., mercuric chloride) from flue gas. However, a major part of Hg⁰, which is insoluble in water and the most volatile of the trace metal species, may pass through wet FGD and particulate matter control devices.⁷⁴ Therefore, should the control of mercury emissions be desired beyond the inherent control by once-through wet FGD, Hg⁰ would need to be adsorbed by the sorbent or converted by reagents or catalysts to a soluble form of mercury that could be collected by a wet FGD process.

A mercury measurement program conducted on six full-scale coal-fired boilers equipped with ESP and limestone or lime FGD processes demonstrated an average total mercury removal across the wet FGD system of 54 percent (ranging from 45 to 67 percent).⁷⁵ The ESP inlet and stack flue gas speciation data indicated 80 to 95 percent removal of Hg⁺⁺ across the ESP and wet FGD system combination. This test program showed that mercury was also removed by the fly ash particles (and occasionally bottom ash). The total mercury removal (defined as the difference between the mercury input based on coal firing rate and coal mercury concentration and mercury stack emissions) ranged from 59 to 75 percent and averaged 67 percent. It should be noted that the results of

this study were obtained during routine wet FGD operations and no adjustments were made to maximize mercury removal.

The statistical analysis of results in the above program showed a significant correlation between oxidized mercury removal and scrubber slurry pH, with a higher pH resulting in higher mercury removal. Among coal parameters (all coals included in the program were mid-chlorine coals), the coal oxygen concentration showed a strong negative correlation with oxidized mercury removal. A weaker correlation was identified between nitrogen and ash content of coal and total mercury removal.

Another study on mercury capture by wet FGD revealed that it could be affected both by the scrubber design (open spray versus tray tower) and operational parameters, such as pH and L/G of the absorber.⁷⁶ Mercury emissions from systems equipped with wet FGD decreased with increasing L/G in the range from approximately 30 up to approximately 130 (gal/1,000 ft³). The decrease of mercury emissions was due to the decrease of oxidized mercury emissions. Elemental mercury emissions following the scrubber remained fairly consistent over the tested range of operating conditions and the outlet elemental mercury concentration was approximately the same as the inlet one. Operation of the scrubber with the gas flow distribution tray enhanced mercury removal over the above L/G range. For example, at the L/G of 100 (gal/1,000 ft³) the mercury emissions for a system with tray scrubber were on the average 38 percent lower than these measured for a system with the open spray scrubber. Pilot-scale tests have demonstrated the potential for removing approximately 85 percent of the total mercury emissions using a wet limestone process, with

a scrubber configured as a tray tower and operated at an L/G of approximately 70 (gal/1,000 ft³).⁷⁴

However, some sampling efforts have indicated an apparent re-emission of Hg⁰ at the outlet of wet FGD systems.⁷⁷ The results of triplicate measurements revealed from 7.1 to 38.5 percent increases of Hg⁰ concentration across a wet FGD system operating on flue gas with inlet concentrations of from 2400 to 2900 ppm SO₂.

As discussed before, should the control of mercury emissions by a wet FGD process alone be desired beyond the inherent control by the existing once-through wet FGD, Hg⁰ would need to be converted by reagents or catalysts to a soluble form of mercury that could be collected.

Therefore, bench-scale research and pilot studies are currently underway to more fully understand the oxidation of Hg⁰ upstream of and subsequent to removal in FGD systems.⁷⁸ The study concentrates on determining whether the catalyst remains active for mercury oxidation after an extended exposure to utility flue gas. One of the findings from the bench-scale phase of this study is that HCl may be participating in the oxidation mechanism of elemental mercury.

Another route pursued on a bench scale is to find liquid additives that, once atomized into the flue gas, would be capable of oxidizing Hg⁰.⁷⁹ Commercial solutions of chloric acid and sodium chlorate were capable of transferring 10 percent of Hg⁰ into solution. Additionally, approximately 80 percent of the nitric oxide was removed. Further pilot-scale evaluations continue to examine mercury speciation and to develop control options.

Dry FGD

Similarly to wet FGD performance discussed above, a wide range of 55 to 96 percent reduction in mercury emissions has been shown with spray dryers installed on full-scale, bituminous-coal-fired boilers.⁷⁴ A significantly lower reduction of 6 to 23 percent was reported for some subbituminous-coal-fired boilers. It is thought that the higher mercury removal efficiencies seen on bituminous-coal-fired boilers are related to the higher coal chlorine concentration in these coals, compared to subbituminous coals.⁷⁴ Pilot-scale tests with the spray drying process have demonstrated a 64 percent total mercury emission reduction across the spray dryer with 68 percent of total inlet mercury being oxidized mercury.⁷⁴

Another dry FGD process that is capable of additionally removing mercury is the CFB. Recently presented results of the pilot-scale testing of a CFB process for mercury adsorption⁸⁰ indicated approximately 50 percent of the total mercury removal by hydrated lime alone and up to 80 percent removal with the supplemental injection of iodine-impregnated activated carbon into the CFB.⁸¹ Only total mercury removal has been tested, mercury speciation in the flue gas was not reported, and there were no attempts made to speciate mercury.

The duct injection process may also be used to control mercury emissions. If, in this process, a sorbent appropriate for mercury capture, such as activated carbon or zeolite⁸², is co-injected along with the sorbent for SO₂ capture, then emissions of SO₂ and mercury may be reduced. In this context, research on modified hydrated lime sorbents has been reported.⁸³ However, the duct injection process has been used sparingly and is

considered, at present, to be a niche application.

In summary, the amount of mercury removed in an unmodified FGD system is believed to be a function of mercury speciation. Wet FGD systems may be able to remove approximately half of the total mercury from the flue gas, depending on the coal fired. Similarly, spray dryers have been found to be able to remove between 6 and 96 percent of total mercury, depending on the type of coal fired. Currently, bench- and pilot-scale research is underway to more fully understand mercury speciation and develop enhanced FGD or stand-alone mercury control options.

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