TECHNICAL PAPER

ISSN 1047-3289 *J. Air & Waste Manage. Assoc.* 51:1676-1688

Flue Gas Desulfurization: The State of the Art

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ABSTRACT

Coal-fired electricity-generating plants may use SO_2 scrubbers to meet the requirements of Phase II of the Acid Rain SO_2 Reduction Program. Additionally, the use of scrubbers can result in reduction of Hg and other emissions from combustion sources. It is timely, therefore, to examine the current status of SO_2 scrubbing technologies. This paper presents a comprehensive review of the state of the art in flue gas desulfurization (FGD) technologies for coal-fired boilers.

Data on worldwide FGD applications reveal that wet FGD technologies, and specifically wet limestone FGD, have been predominantly selected over other FGD technologies. However, lime spray drying (LSD) is being used at the majority of the plants employing dry FGD technologies. Additional review of the U.S. FGD technology applications that began operation in 1991 through 1995 reveals that FGD processes of choice recently in the United States have been wet limestone FGD, magnesium-enhanced lime (MEL), and LSD. Further, of the wet limestone processes, limestone forced oxidation (LSFO) has been used most often in recent applications.

The SO_2 removal performance of scrubbers has been reviewed. Data reflect that most wet limestone and LSD installations appear to be capable of ~90% SO_2 removal. Advanced, state-of-the-art wet scrubbers can provide SO_2 removal in excess of 95%.

Costs associated with state-of-the-art applications of LSFO, MEL, and LSD technologies have been analyzed with appropriate cost models. Analyses indicate that the capital cost of an LSD system is lower than those of same capacity LSFO and MEL systems, reflective of the relatively less complex hardware used in LSD. Analyses also reflect that, based on total annualized cost and SO_2 removal requirements: (1) plants up to ~250 MW_e in size and firing low- to medium-sulfur coals (i.e., coals with a sulfur content of 2% or lower) may use LSD; and (2) plants larger than 250 MW_e and firing medium- to high-sulfur coals (i.e., coals with a sulfur content of 2% or higher) may use either LSFO or MEL.

IMPLICATIONS

Coal-fired power plants may use SO_2 scrubbers to meet the requirements of Phase II of the Acid Rain SO_2 Reduction Program. Additionally, the use of scrubbers can result in reduction of Hg and other emissions from combustion sources. This paper presents a comprehensive review of the state of the art in FGD technologies for coal-fired boilers.

INTRODUCTION

SO_2 emissions are known to cause detrimental impacts on human health and the environment. The major health concerns associated with exposure to high ambient concentrations of SO_2 include breathing difficulty, respiratory illness, and aggravation of existing cardiovascular disease. In addition to the health impacts, SO_2 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 deterioration of buildings and monuments. While airborne, SO_2 and its particulate matter (PM) derivatives contribute to visibility degradation.

Combustion of sulfur-containing fuels, such as coal and oil, results in SO_2 formation. Electricity-generating plants account for the majority of SO_2 emissions in the United States. The Acid Rain SO_2 Reduction Program, established under Title IV of the Clean Air Act Amendments of 1990, was designed to reduce SO_2 emissions from the power-generating industry. Phase I of this program began on January 1, 1995, and ended on December 31, 1999. In 1997, 423 electricity-generating units, affected under Phase I, emitted 5.4 million tons (4903.2×10^6 kg) of SO_2 compared with the allowable 7.1 million tons (6446.8 ×

10^6 kg).[1] Thus, the SO_2 emissions in 1997 were 23% below the allowable amount. Phase II of the Acid Rain SO_2 Reduction Program began on January 1, 2000. To meet the requirements of this phase, some power plants may use flue gas desulfurization (FGD) technologies. Additional environmental benefits that may result from the use of these technologies are synergistic reductions in Hg emissions, as well as reductions in fine PM concentrations in the atmosphere. It is timely, therefore, to examine the current status of FGD (or SO_2 scrubbing) technologies applicable to electricity-generating plants.

The review of FGD technologies presented in this paper describes these technologies, assesses their applications, and characterizes their performance. Then, the paper presents an analysis of the costs associated with limestone forced oxidation (LSFO), lime spray drying (LSD), and magnesium-enhanced lime (MEL) FGD technology applications. It is expected that this review will be useful to a broad audience, including individuals responsible for developing and implementing SO_2 control strategies at sources, persons involved in developing SO_2 and other regulations, state regulatory authorities implementing SO_2 control programs, and the interested public at large. Persons engaged in research and development efforts aimed at improving cost-effectiveness of FGD technologies may also benefit from this review.

CLASSIFICATION OF FGD TECHNOLOGIES

Various technologies exist to remove SO_2 from flue gas produced by electricity-generating plants. Existing FGD technologies were comprehensively evaluated by the Electric Power Research Institute in their review report.[2] The technologies discussed in this report represent a broad spectrum of maturity. Some can claim tens of thousands of hours of commercial operating experience, while others have been tested only at pilot-scale. A compendium of FGD technology applications is provided in the CoalPower3 database, available from the International Energy Agency's Clean Coal Centre in London.[3]

Conventionally, FGD processes can be classified as once-through or regenerable, depending on how the sorbent is treated after it has sorbed SO_2. In once-through technologies, the spent sorbent is disposed of as a waste or utilized as a byproduct. In regenerable technologies, SO_2 is released from the sorbent during the sorbent's regeneration, and the SO_2 may be further processed to yield H_2SO_4, elemental sulfur, or liquid SO_2. No waste is produced in regenerable technology applications. Both once-through and regenerable technologies can be further classified as either wet or dry. In wet processes, wet slurry waste or byproduct is produced, and the flue gas leaving the absorber is not saturated. The classification of FGD processes is shown in Figure 1.

At present, regenerable FGD technologies are being used only marginally in the United States and abroad, as evident from Table 1.[3] This may be because these processes involve relatively higher costs compared with other FGD processes. For example, capital costs for FGD technology application on a new 300-MW_e plant burning 2.6% sulfur coal were estimated at 170 and 217 $/kW for wet once-through FGD and sodium sulfite regenerable processes, respectively.[2] Considering the relatively marginal application of regenerable FGD processes, this paper focuses only on once-through FGD processes. Accordingly, when wet FGD is mentioned in the remainder of this paper, it is meant to be once-through wet FGD. Similarly, when dry FGD is mentioned, it is meant to be once-through dry FGD.

DESCRIPTION OF ONCE-THROUGH PROCESSES

In once-through technologies, the SO_2 is permanently bound in the sorbent, which must be disposed of as a waste or utilized as a byproduct (e.g., gypsum). This section presents the FGD processes reported in literature[2] and in an International Energy Agency database of commercial applications.[3] For each process, typical SO_2 reduction, advantages, and any constraints are described.

Once-Through Wet FGD Technologies

In these technologies, SO_2-containing flue gas contacts alkaline aqueous slurry in an absorber. The slurry is generally made from either lime [typically 90% or more $Ca(OH)_2$] or limestone (typically 90% or more $CaCO_3$). The most often used absorber application is the countercurrent vertically oriented spray tower. A generic layout of a limestone-based wet FGD process is shown schematically in Figure 2.

In the absorber, SO_2 dissolves in the slurry and initiates the reaction with dissolved alkaline particles. The absorber slurry effluent, containing dissolved SO_2, is held in a reaction tank, which provides the retention time for finely ground lime or limestone particles in the slurry to

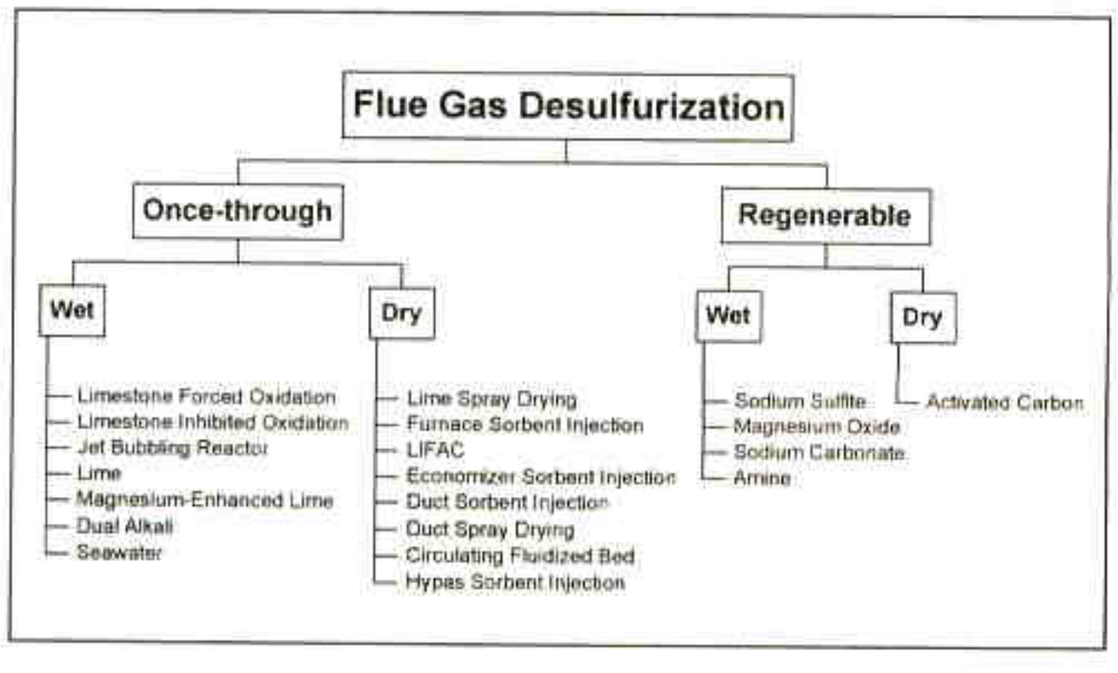

Figure 1. Classification of FGD processes.

Table 1. Generating capacity (MW_e) equipped with FGD technology through 1998.

Technology	United States	Abroad	World
Wet	82,092	114,800	196,892
Dry	14,081	10,654	24,735
Regenerable	2798	2394	5192
Total FGD	98,971	127,848	226,819

dissolve and to complete the reaction with the dissolved SO_2. As a result of this reaction, sulfite/sulfate crystallization occurs in the reaction tank, and alkalinity of the slurry is depleted. Fresh slurry is added to the reaction tank to compensate for this depletion and thereby maintain a *desired level of alkalinity. The slurry is recirculated from* the reaction tank into the absorber. Reaction products from the reaction tank are pumped to the waste-handling equipment, which concentrates the waste. From the waste-handling equipment, the concentrated waste is sent for disposal (ponding or stacking) or, alternatively, processed to produce a salable gypsum (calcium sulfate dihydrate) byproduct. The practical wet FGD processes are described in the following sections.

Limestone Forced Oxidation. Over the years, LSFO, which minimizes scaling problems in the absorber, has become the preferred wet FGD technology process. Gypsum scale typically forms via natural oxidation when the fraction of $CaSO_4$ in the slurry (slurry oxidation level) is greater than 15%. In LSFO, scaling is prevented by forcing oxidation of $CaSO_3$ to $CaSO_4$ by blowing air into the reaction tank (in situ oxidation) or into an additional hold tank (ex situ oxidation).[4] The gypsum thus formed is removed as usual and, as a consequence, the concentration of gypsum in the slurry recycled to the absorber decreases.

The LSFO process can remove in excess of 95% of SO_2. 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. Additional benefits are formation of a stable product, the potential for elimination of landfilling, and smaller dewatering equipment. Further, depending on site-specific conditions, LSFO may produce a salable byproduct 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. The operation of the LSFO process can be improved when organic acids, such as dibasic acid, are added to the limestone slurry. The use of organic acid buffering allows for a smaller absorber and increased sorbent utilization.

Variations in LSFO process design include a cocurrent, downflow absorber with a single level of grid packing. The cocurrent contact of slurry and flue gas allows for a higher flue gas velocity and results in a reduced pressure drop. Additionally, combining the cocurrent absorber tower and reaction tank can reduce space requirements. In this design, limestone slurry is sprayed above the grid and is contacted by the flue gas. Simultaneous forced oxidation and agitation in the reaction tank is accomplished *with a rotating air sparger. This sparger prevents solids* from settling out in the reaction tank and provides nearly complete oxidation of $CaSO_3$ to $CaSO_4$.

Another variation in LSFO design includes contacting flue gas with dilute slurry in a double-loop recycle system. Approximately 25–30% of the SO_2 in the flue gas reacts with the recycle slurry of $CaSO_4$ and $CaCO_3$ in the lower, first stage of the absorber. The flue gas then flows upward to the second stage, where the remaining SO_2 is contacted with dilute slurry of $CaSO_3$, $CaSO_4$, and $CaCO_3$ in the second recycle loop. The $CaSO_3$ reaction product slurry from the second loop drains to the lower first loop of the absorber, where it is oxidized to $CaSO_4$. Minimal addition of fresh $CaCO_3$ to a lower loop helps decrease pH and promote $CaSO_3$ oxidation.

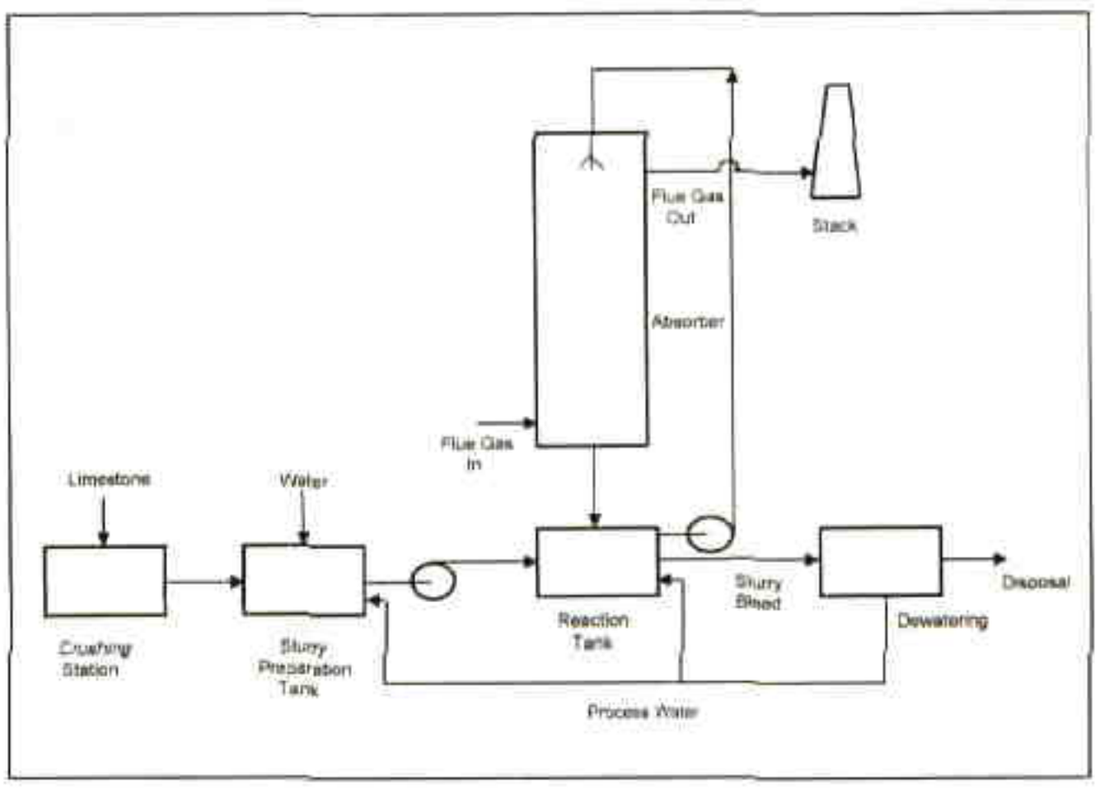

Figure 2. Wet FGD processes.

Limestone Inhibited Oxidation. Another wet limestone process designed to control oxidation in the absorber is limestone inhibited oxidation (LSIO), in which emulsified sodium thiosulfate is added to the limestone slurry feed to prevent the oxidation of $CaSO_3$ to gypsum in the absorber by lowering the slurry oxidation level to less than 15%. 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, and conversion to thiosulfate occurs when sulfur contacts sulfite in the reaction tank. The overall conversion of sulfur to thiosulfate is between 50 and 75%. The amount of thiosulfate (or sulfur) required to achieve inhibited oxidation is a function of system chemistry and

operating conditions. The LSIO chemistry is particularly efficient in applications with high-sulfur coals,[5] because the difficulty in inhibiting the oxidation generally increases with decreasing sulfur content in coal.

In some instances, LSIO may be economically preferred over LSFO when a salable gypsum byproduct is not required. This is because LSIO does not require the use of air compressors, as does LSFO. An additional benefit of using LSIO may be increased limestone solubility, which enhances sorbent utilization. However, in general, solids dewatering is more difficult in LSIO compared with LSFO due to a higher level of sulfites. The waste product, $CaSO_3$, resulting from the LSIO process is landfilled. Note that the LSIO waste has improved dewatering characteristics compared with the waste from natural oxidation operation of a wet FGD absorber. This is because the $CaSO_3$ product from the LSIO tends to form larger crystals, similar to gypsum solids.

Jet Bubbling Reactor. The jet bubbling reactor (JBR) process represents a different approach to gas/liquid contacting for SO_2 removal than does LSFO or LSIO. In JBR, SO_2 absorption, sulfite/bisulfite oxidation, and precipitation of gypsum are accomplished in a single reaction vessel. The contact is achieved by injecting flue gas through gas sparger tubes immersed below the surface of the limestone scrubbing slurry. The so-called "jet bubbling zone" is formed, in which the flue gas vigorously bubbles through the surrounding liquid, thus creating a large gas/liquid interfacial area for SO_2 absorption.[6] In this zone, maintained at a slightly lower pH than that for LSFO (3.5–4.5 compared with 5.5–6.5) to increase reaction rates and prevent sulfite and carbonate scale formation, neutralization and oxidation of bisulfites and formation of gypsum crystals occur. The lower pH allows the JBR to attain essentially 100% utilization of limestone.

The overall chemical reactions in the JBR are similar to those occurring in the LSFO. However, the intermediate reaction compound is a nonscaling bisulfite instead of the scale-producing sulfite. As a result, JBR produces gypsum crystals, which are larger and dewater better than gypsum crystals from LSFO. The total system pressure drop is greater than most conventional spray tower LSFOs. However, the JBR design allows elimination of high-energy-demand slurry spray pumps.

Lime and Magnesium-Enhanced Lime. The lime process uses hydrated calcitic lime slurry in a spray tower, which predominantly is countercurrent flow. Because this slurry is more reactive than limestone slurry, the absorber designed for lime sorbent is generally smaller compared with one designed for limestone slurry. However, lime sorbent is more expensive than limestone sorbent.

The MEL process is a variation of the lime process in that it uses a special type of lime that contains magnesium in addition to its calcitic component. Because of the greater solubility of magnesium salts compared with calcitic sorbents, the scrubbing liquor is significantly more alkaline. Therefore, MEL is able to achieve high SO_2 removal efficiencies in significantly smaller absorber towers than its calcitic lime sorbent counterparts. Additionally, less MEL slurry is needed compared with LSFO for the same level of SO_2 removal. Also, because of the lower liquid recirculation requirement, pumps are smaller, and the scrubber-gas-side pressure drop is lower in an MEL system than in a comparable LSFO system. For these and other reasons, process energy requirements are lower in MEL compared with those needed in LSFO. Furthermore, gypsum produced from the MEL process may be lighter in color than that produced by LSFO. If desired, $Mg(OH)_2$ byproduct can also be produced from the MEL process.[7] $Mg(OH)_2$ is an alkaline reagent, which can be used to reduce SO_3 emissions and also to treat plant liquid effluents prior to discharge.

Dual Alkali. This process utilizes two alkaline materials: a sodium solution for scrubbing and lime for treatment of the scrubbing solution. A sodium sulfite solution is sprayed into an open spray tower to remove SO_2 from the flue gas. Lime is added to the product solution in an external tank to recover the sodium solution and form a $CaSO_3$-rich sludge. Because the absorption step uses a soluble alkali, the dissolution rate of the reagent is not the rate-limiting step as it is in LSFO. Consequently, lower liquid/gas (L/G) ratios are used in the dual alkali process compared with those used in LSFO.

The dual alkali process produces $CaSO_3/CaSO_4$ sludge. This sludge must be disposed of in a lined landfill because of sodium scrubbing solution losses to the product material and the resulting sodium salt concentration in the filter cake. Scrubbing solution losses may be decreased by improved filter cake washing techniques.

In a variation of the dual alkali process, limestone may be added to a slipstream from an open spray tower removing SO_2. Limestone simultaneously recovers sodium sulfite and forms sludge rich in $CaSO_3$. Similarly to the requirements for the lime-based dual alkali process, a lined landfill may be required because of the soluble sodium salts entrained in the solid product. Additionally, these solids must be fixated with lime and fly ash.

The Seawater Process. This process utilizes the natural alkalinity of seawater to neutralize SO_2. The chemistry of the process is similar to that of LSFO, except it does not involve any dissolution or precipitation of solids. Seawater may be available in large amounts at the power plant

as a cooling medium in the condensers. It then can be 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 SO_2.

The absorption of SO_2 takes place in an absorber, where seawater and flue gas are brought into close contact in a countercurrent flow. The scrubber effluent flows to the treatment plant, where it is air-sparged to oxidize absorbed SO_2 into sulfate before discharge.[8] Since sulfate is completely dissolved in seawater, it does not result in any waste product that would require disposal. Sulfate is a natural ingredient in seawater, and typically there is only a slight increase in its concentration in the discharge. This increase is within the variation naturally occurring in seawater. The difference from the background level is normally not detectable within even a short distance from the point of discharge.

Because the utilization of seawater for SO_2 scrubbing introduces a discharge to the ocean, it is necessary to make an assessment based on local conditions. Typically, this assessment includes effluent dilution and dispersion calculations, a description of the effluent, a comparison of effluent data with local quality criteria, a description of the 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.[9]

Once-Through Dry FGD Technologies

In these technologies, SO_2-containing flue gas contacts alkaline (most often lime) sorbent. As a result, dry waste is produced, which is generally easier to dispose of than waste produced from wet FGD processes. The sorbent can be delivered to the flue gas in an aqueous slurry form (LSD) or as a dry powder [furnace sorbent injection (FSI), LIFAC process (LIFAC), economizer sorbent injection (ESI), duct sorbent injection (DSI), duct spray drying (DSD), circulating fluidized bed (CFB), or Hypas sorbent injection (HSI)]. LSD and CFB require dedicated absorber vessels for sorbent to react with SO_2, 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. A schematic of dry FGD processes involving dry powder injection and DSD is shown in Figure 3. In this figure, the flue gas flow for a plant without FGD is shown with the solid line. Sorbent injection locations for alternative dry FGD processes with dry powder injection or DSD are shown schematically with broken lines. These processes are discussed in the following sections.

Lime Spray Drying. This process is most often used by sources that burn low- to medium-sulfur coal. The schematic of LSD is shown in Figure 4. Rotary atomizers or two-fluid nozzles are used to finely disperse lime slurry

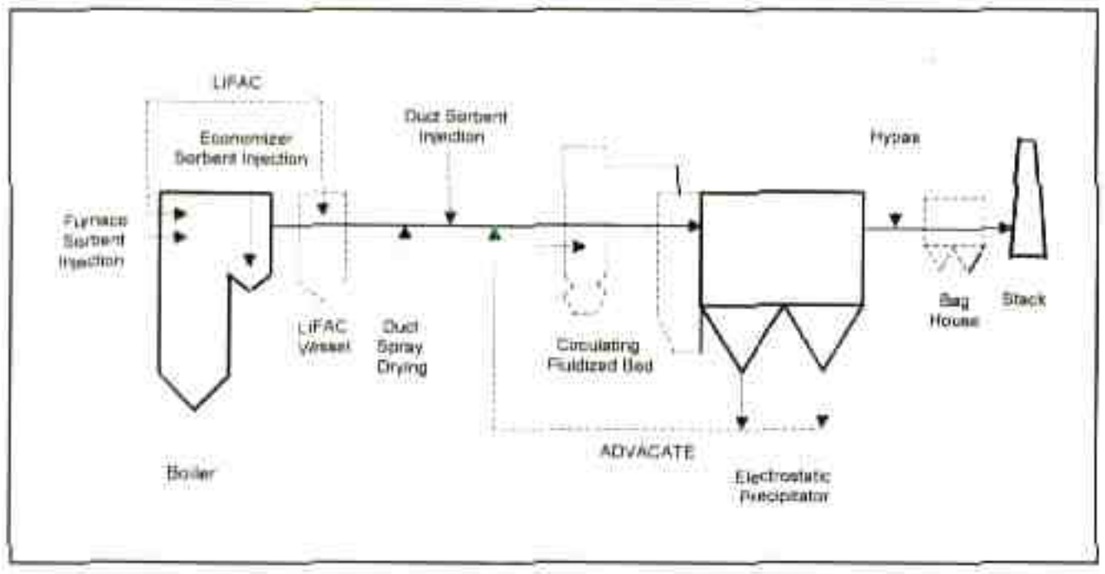

Figure 3. Sorbent injection processes.

into the flue gas. Hot flue gas mixes in a spray dryer vessel with a mist of finely atomized fresh lime slurry. Simultaneous heat and mass transfer between alkali in the finely dispersed lime slurry and SO_2 from the gas phase results in a series of reactions and a drying of reacted products. A close approach to adiabatic saturation (from 10 to 15 °C for coal-derived flue gas) is required to achieve high SO_2 removal. However, complete saturation can impair operation of a spray dryer because of wet solids adhering to vessel walls, to gas flow passages from the vessel, and in the particulate collector.[10] Therefore, the water content of the slurry fed into the spray dryer is carefully controlled to avoid complete saturation of the flue gas.

Studies indicate that most SO_2 capture in the spray dryer occurs when the sorbent is still moist. Therefore, deliquescent additives may be used to increase the duration of time in which the sorbent remains moist. A similar effect is achieved when spray dryers are used on coals with elevated chloride content. However, the addition of deliquescent materials needs to be closely controlled to avoid the wet solids problem noted previously.

Furnace Sorbent Injection. In FSI, dry sorbent is injected directly into the section of the furnace where temperatures are between 950 and 1000 °C. Sorbent particles (most often hydrated lime, sometimes limestone) decompose and become porous solids with high surface area.[11] $CaSO_4$ and any remaining unreacted sorbent leave the

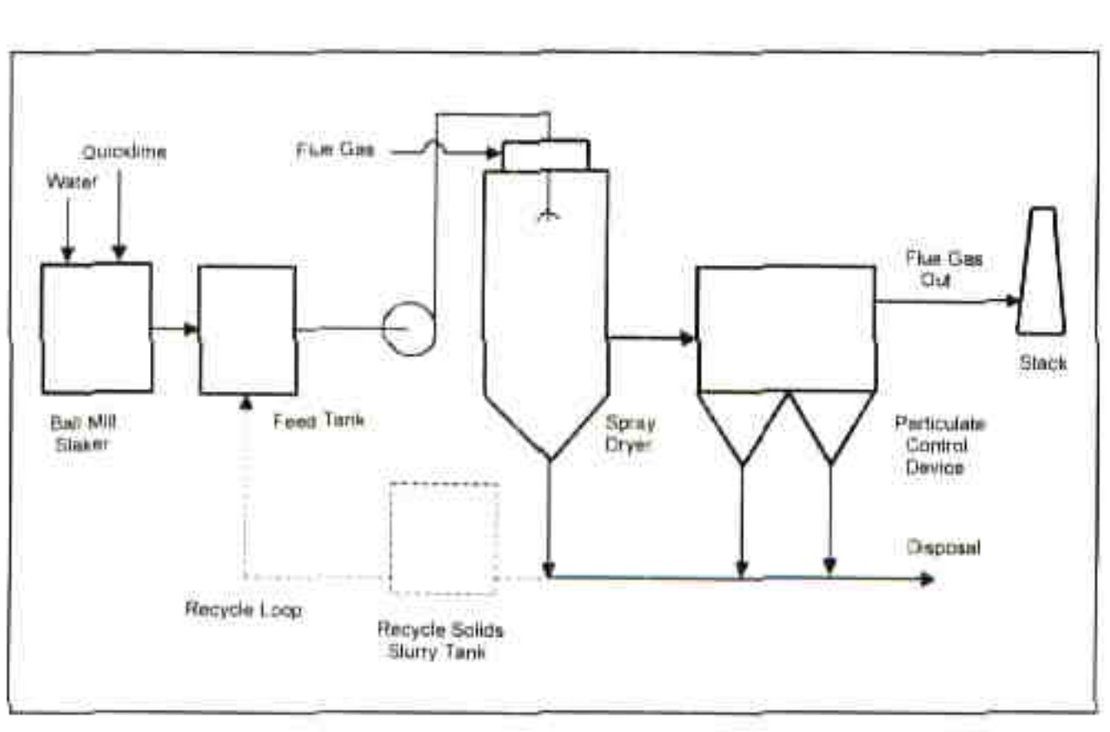

Figure 4. The LSD process.

furnace with the flue gas and are captured as solids in a particulate collection device.

Up to 50% SO_2 removal can be achieved with the FSI process. Critical parameters for the FSI process include injection temperature and residence time in the furnace. Proper injection temperature in the furnace is important so that the CaO formed in the calcination reactions is not dead-burned and sintered. Sufficient residence time in the furnace is also needed to allow reaction of the lime particles with SO_2.

LIFAC. In LIFAC, finely pulverized limestone is injected into the upper part of the furnace, where a portion of the SO_2 is removed in a manner described above for the FSI. The reaction products entrained in the flue gas (along with the fly ash) pass into the activation reactor, where water is sprayed to humidify the flue gas for additional SO_2 removal and particulate conditioning. Dry solid product is captured downstream in the electrostatic precipitator (ESP). The LIFAC process could be considered as FSI with downstream humidification and was developed to improve the SO_2 removal efficiency, improve reagent utilization, and reduce the potential adverse effects on ESP performance that may occur with use of FSI alone.[12]

SO_2 removal in excess of 80% can be achieved with the LIFAC process. Critical parameters for the LIFAC process include temperature of the flue gas at sorbent injection location, residence time in the furnace, temperature of the flue gas entering the activation reactor, droplet size of the water sprayed into the reactor, and residence time in the reactor. Both the flue gas temperature at the reactor inlet and the injected water droplet size affect the water evaporation rate in the reactor. Longer residence time in the reactor is needed for evaporation of the larger water droplets.

Economizer Sorbent Injection. In this process, lime is injected into the convective pass of a coal-fired utility boiler to react with SO_2. The optimum temperature range for SO_2 removal is between 500 and 570 °C. Upon injection, the sorbent reacts directly with SO_2 to form particles of $CaSO_3$. A portion of lime (~10%) decomposes to form reactive CaO particles, which react with SO_2 to form $CaSO_3$ and some $CaSO_4$.[13] Additionally, water may be injected into the ductwork between the air preheater and the existing ESP to increase SO_2 removal efficiency by hydrating any unreacted CaO. The reaction product solids and fly ash are collected in the ESP and sent to an unlined landfill for disposal. With optimum sorbent preparation and proper injection temperature, SO_2 removal efficiencies of up to 80% can be achieved.

Duct Sorbent Injection. This process is intended to control SO_2 directly in the flue gas duct between the air preheater and the particulate control device. In this process, finely dispersed dry sorbent (most often hydrated lime, occasionally sodium bicarbonate) is injected into the flue gas downstream of the boiler's air preheater. Water may be injected into the flue gas upstream of the sorbent to enhance the SO_2/sorbent reaction.[14] Fly ash, reaction products, and any unreacted sorbent are collected in the particulate control device. Some of the particulate control device's catch is recirculated into the duct to increase sorbent utilization, while the remaining catch is disposed. Approximately 50–60% SO_2 capture may be expected with the DSI using lime, and up to 80% SO_2 capture could be achieved with sodium bicarbonate sorbent.

An advanced version of DSI is the ADVACATE process, in which fresh CaO is hydrated and mixed in one step with recycled solids to form a slurry containing ~30% by weight of solids. This slurry is processed in a vertical ball mill to expose fresh silica surfaces for reaction with hydrated lime to form highly reactive, noncrystalline, calcium silicate slurry. The slurry from the mill is pumped to a large mix tank that provides sufficient residence time for the complete reaction of lime with SiO_2 spheres.[15] Dry recycle solids are mixed in a pug mill with the fresh sorbent/recycle slurry to make a slurry with ~70% by weight of solids. This slurry can be injected into the duct downstream of the air preheaters. SO_2 removal of 90% or more can be achieved with ADVACATE.

Duct Spray Drying. In the DSD process, slaked lime slurry is sprayed directly into the ductwork upstream of the existing ESP. Either a rotary atomizer or a dual-fluid atomizer is used to disperse the sorbent into the flue gas.[16] The SO_2 in the flue gas reacts with the alkaline slurry droplets as they dry, forming $CaSO_3$ and $CaSO_4$. To allow for sufficient drying of the slurry droplets, the existing ductwork must be capable of providing at least a 1-sec, but preferably a 2-sec, residence time and must not contain any flow obstructions. The water entering with the lime reagent humidifies the flue gas for better SO_2 removal and ESP conditioning. The reaction products and fly ash are captured downstream in the ESP. The solids collected from the ESP are transported to an unlined landfill for disposal.

Circulating Fluidized Bed. Dry sorbent (hydrated lime) is contacted with humidified flue gas in a CFB downstream of the air preheater. The bed provides a long contact time between the sorbent and flue gas because sorbent passes through the bed several times. CFB is characterized by good SO_2 mass transfer conditions from the gas to the solid phase,[17] which are 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 unreacted alkali. Entrained reaction products are carried by flue gas 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 disposed. Because of a higher PM concentration downstream of the fluidized bed, improvements to the existing ESP may be needed to maintain the required particulate emission levels.

Hypas Sorbent Injection. In the HSI process, the fly ash is first removed from the flue gas by the existing particulate control system. Next, water is injected to cool and humidify the gas. A dry mixture of lime and recycled solids is then injected into the humidified flue gas for reaction with SO_2. The reaction byproducts and remaining fly ash are collected in an added pulse jet fabric filter. A portion of the used reagent collected in the fabric filter is reinjected with fresh sorbent to improve SO_2 removal and overall sorbent utilization.

TECHNOLOGY APPLICATIONS

FGD technology applications were reviewed to identify the technologies that are predominantly being used at power plants. This review was conducted using the data available in ref 3. It should be noted that, as of December 2000, new data became available on the extent of MEL application in the United States,[18] which indicate that MEL has been applied on 15,723 MW_e of capacity in the United States. While this is noted, the review of FGD technology applications was based solely on the information available in ref 3 to maintain consistency in the applications-related data for the numerous technologies considered in this work. Findings of this review are described in the following sections.

Table 1 statistics describe the installation of FGD systems at coal-fired electric power plants through 1998. FGD systems were installed to control SO_2 emissions for 226,819 MW_e of generating capacity worldwide. When capacity is mentioned in this paper, gross or maximum capacity is meant. Of this capacity, 86.8% utilizes wet FGD technologies, 10.9% dry FGD technologies, and the remaining 2.3% regenerable FGD technologies. A similar pattern of FGD technology application can be seen in the United States. Through 1998, almost 100,000 MW_e of capacity in the United States was equipped with FGD technology. Of this capacity, 82.9% utilizes wet FGD technologies, 14.2% dry FGD technologies, and the remaining 2.9% regenerable FGD technologies.

Of the U.S. electricity-generating capacity equipped with wet FGD technologies, 68.9% uses limestone processes. Also, 80.4% of the U.S. generating capacity equipped with dry FGD technologies uses LSD. Limestone wet FGD technology usage also dominates the overseas applications. Limestone processes are used for 93.2% of the overseas electric-generating capacity equipped with wet FGD technology, which make up 89.8% of the total FGD applications abroad. Also, 64.8% of the overseas generating capacity equipped with dry FGD technology uses LSD.

Table 2 summarizes the extent of use through 1998 for FGD processes discussed previously. As can be seen from Table 2, the extent of application of once-through FGD processes varies greatly. Each of the LSFO, wet lime FGD, MEL, and LSD processes has been applied, worldwide, on more than 5000 MW_e generating capacity. In addition, more than 39,000 MW_e of worldwide generating capacity uses natural oxidation-based and LSIO limestone wet FGD systems. Clearly, these processes fall in the category of fully commercial. Three other processes in Table 2 (ESI, DSD, and HSI) may be classified as near-commercial to recognize the fact that they have been demonstrated.[2] However, no data on existing commercial application for any of these three processes could be found. The remaining processes in Table 2 may be classified as those with limited commercial experience. This is an intermediate category for processes that have gained some level of commercial application. This level can vary from,

Table 2. The extent of application of once-through FGD processes (through 1998).

Process	Category	U.S. Applications, MW_e	Foreign Applications, MW_e
LSFO	O/W[a]	20,190	103,827
Other limestone[b]	O/W	36,247	3112
JBR	O/W	123	2012
Wet lime FGD	O/W	14,237	4338
MEL	O/W	8464[c]	50
Dual alkali	O/W	1648	0
Seawater	O/W	75	1050
LSD	O/D[d]	11,315	6904
FSI	O/D	286	2108
LIFAC	O/D	60	978
ESI	O/D	N/C[e]	N/C
DSI	O/D	2400	1125
DSD	O/D	N/C	N/C
CFB	O/D	80	517
HSI	O/D	N/C	N/C

[a]O/W = once-through/wet; [b]Includes LSIO and natural oxidation wet limestone applications; [c]15,723 MW_e of MEL application in the United States were reported by a source other than ref 3 (ref 18). However, as stated before, this review of FGD technology applications is based solely on the data available in ref 3. The decision to elect a single source was dictated by the necessity to maintain consistency in application-related data for the technologies considered in this paper; [d]O/D = once-through/dry; [e]N/C = near commercial.

for example, 60 MW_e in the United States for LIFAC to as much as 2400 MW_e in the United States for DSI.

Recent FGD technology selections made by the U.S. electricity-generating industry can be further understood by examining the pertinent data in ref 3. Between 1991 and 1995, 19,154 MW_e of U.S. electric-generating capacity was retrofitted with FGD technologies. Of this capacity, 75, 17.5, and 7.5% were equipped with LSFO, MEL, and LSD, respectively. Based on these data, FGD processes of choice recently in the United States have been wet limestone FGD, MEL, and LSD (no wet lime FGD applications). Further, of the wet limestone processes, LSFO has been used most often in recent applications.

TECHNOLOGY PERFORMANCE

As discussed before, wet limestone processes (i.e., LSFO, LSIO, JBR, and natural oxidation) and LSD represent the most widely applied FGD technologies. Further, MEL has been used in recent FGD applications. As such, it is useful to assess the SO_2 removal performance potential of these technologies. For this purpose, the design SO_2 removal efficiencies associated with applications of these technologies reported in the CoalPower3 database[1] were examined. These data reflect that wet limestone systems have been designed for high SO_2 removals of up to 98%. However, most wet limestone systems appear to be designed for 90% SO_2 removal. Even though the median design efficiency for all units with wet limestone processes in the CoalPower3 database is 90%, it is worth noting that state-of-the-art wet scrubbers are capable of routinely achieving SO_2 removal efficiencies of greater than 95%.[19] The high-velocity LSFO process, with state-of-the-art design options, is reportedly capable of removing more than 99.6% of SO_2 under test conditions.[20]

Spray dryers often achieve greater than 90% SO_2 removal on coals with 1–2% sulfur.[21] CoalPower3 data indicate that, while the median design efficiency for all units using LSD is 90%, all spray dryers installed during 1991–1995 have a design SO_2 removal efficiency between 90 and 95%. While the median design efficiency for all units with MEL processes in the CoalPower3 database is 90%, it is worth noting that state-of-the-art MEL scrubbers are capable of achieving 98% SO_2 removal while operating at lower L/G ratios than LSFO systems designed to remove SO_2 with the same efficiency. Pilot-scale testing of MEL at an L/G ratio of 45 and inlet SO_2 concentration of 2300 ppmv demonstrated 98% removal efficiency, with an average removal of 97%.[22]

It is useful to examine the improvement in performance of wet limestone and LSD processes over the period of their application. Figure 5 shows ranges and medians of design SO_2 removal efficiency for the pertinent populations of wet limestone FGD and LSD installations in each of the last three decades. A steady improvement in design SO_2 removal efficiency is evident for these processes. This improvement is likely due, in part, to more stringent SO_2 control requirements. However, the trends do reflect that the SO_2 removal efficiencies for the processes considered have improved with time. Note that the lime spray drying process did not become commercial until the early 1980s; therefore, no efficiency could be characterized for the 1970s for this process.

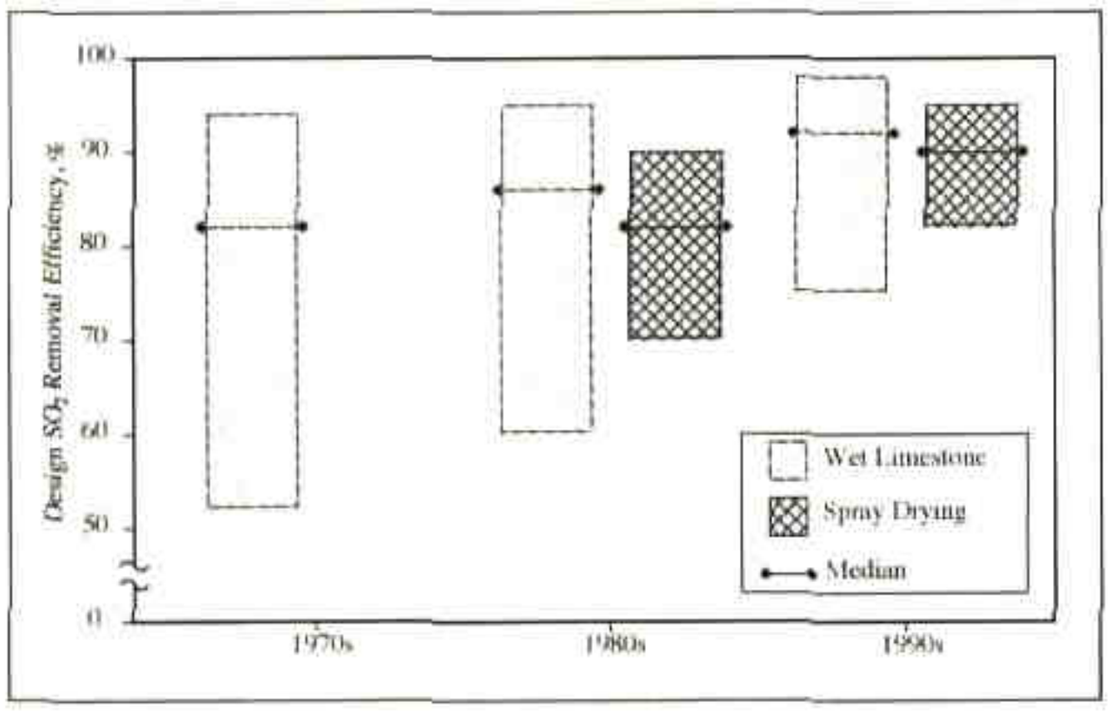

Figure 5. Improvement in design efficiency of FGD technologies.

COSTS OF FGD TECHNOLOGIES

As discussed before, LSFO, LSD, and MEL have been the processes of choice in recent U.S. applications. Therefore, in this work, cost models were developed for state-of-the-art applications of these processes. In the ensuing paragraphs, descriptions and results are provided for the cost models developed in this work. Additional details on these models can be found in ref 23.

Costing Methodology and Economic Assumptions

Pollution control technology costs can generally be categorized as capital, operating and maintenance (O&M), and total annualized costs. Capital cost includes all costs incurred to construct a facility and get it ready to perform its function. O&M costs can be further subdivided into fixed and variable components. The fixed O&M cost 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, energy cost, and cost of any other consumables. Total annualized cost includes the costs associated with capital recovery and annual O&M charges.

Following the EPRI Technical Assessment Guide methodology,[24] in this work, the capital cost of an FGD technology is determined as total capital requirement (TCR). TCR includes the costs associated with installed equipment, general facilities, engineering fees, contingencies, prime contractor's fee, allowance for funds during construction,

inventory cost, and preproduction costs. The TCR calculation methodology is shown in Table 3. Moreover, Table 4 presents the assumptions used in estimating the capital and O&M costs of LSFO, LSD, and MEL technologies. The correlations for various elements of O&M cost applicable to LSFO, LSD, or MEL can be seen elsewhere.[23]

An important consideration in FGD technology applications is the potential ability of a plant to sell SO_2 allowances. Under the Acid Rain SO_2 Reduction Program,[1] trading of SO_2 allowances is permitted. Thus, for example, power-generating plants may elect to comply with their emission limit requirements by installing FGD technology or purchasing SO_2 emission allowances. These SO_2 emission allowances would become available if a plant installs FGD technology to remove more SO_2 than required. By selling these emission allowances, the plant may offset part of the costs associated with FGD technology application. However, while emissions trading may be an important consideration affecting the selection of FGD technology for a plant, this potential for decreasing costs of an application has not been analyzed in this paper. The economics associated with emissions trading would, in general, be plant-specific and depend on market conditions. Therefore, the effects of emissions trading are considered to be beyond the scope of this work.

The LSFO and LSD Cost Models

The Air Pollution Prevention and Control Division of the U.S. Environmental Protection Agency's (EPA's) Office of Research and Development has recently published the Coal Utility Environmental Cost (CUECost) Workbook User's Manual,[25] which can provide budgetary cost estimates with ±30% accuracy for LSFO and LSD applications, based on user-defined design and economic criteria. The algorithms in CUECost provided the starting point for the LSFO and LSD cost models developed in this work. First, sensitivity analyses were conducted with CUECost LSFO and LSD algorithms to identify variables that have a minor impact on cost (i.e., a deviation of less than 5% over selected baselines). These sensitivity analyses revealed that for both LSFO and LSD applications, the majority of cost impacts could be captured through considering capacity, heat rate, coal sulfur content, and coal heating value. The details of sensitivity analyses are given elsewhere.[23]

Next, variables other than capacity, heat rate, coal sulfur content, and coal heating value were fixed at typical values in the corresponding CUECost algorithms to arrive at simplified LSFO and LSD cost models. The resulting simplified LSFO and LSD cost models were then validated using published data.[25-27] Validation results shown in Tables 5 and 6 reflect that, on average, the simplified LSFO and LSD cost models predict the published costs within 10.5 and 15.6%, respectively. The results also reflect that simplified LSFO and LSD cost models are capable of providing budgetary cost estimates within ±30% accuracy.

Table 3. Capital cost calculation methodology used in this work.

Cost Component	Symbol/Calculation
Installed equipment capital cost	BM
Facilities + engineering and home office + process contingency	$A = A_1 + A_2 + A_3$
Project contingency	B
Fee	C
Total plant cost	$TPC = BM \times (1 + A) \times (1 + B) \times (1 + C)$
Financial factor[a]	$D = F_{TCE} + F_{AFDC}$
Total plant investment	$TPI = TPC \times (1 + D)$
Preproduction cost[b] + inventory capital	E
Total capital requirement	$TCR = TPI + E$

[a]F_{TCE} and F_{AFDC} account for total cash expended and allowance for funds during construction, respectively; [b]Preproduction cost incorporates one-twelfth of the projected annual O&M expenses and 2% of the TPI estimate.

Table 4. Economic assumptions used in estimating FGD technology costs.

Parameter	Value(s) or Choice		
	LSFO	LSD	MEL
Cost Basis	1998 Constant Dollars		
Capital Cost-Related:			
General facilities (%)	5	5	5
Engineering and home office (%)	10	10	10
Process (%)	5	5	5
Project (%)	15	15	15
Prime contractor's fee (%)	3	3	3
F_{AFDC} (%)	7.6	7.6	7.6
F_{TCE} (%)	1.0	1.0	1.0
Retrofit difficulty	medium	medium	medium
O&M-Related:			
Operating labor rate ($/hr)	30	30	30
Cost of steam ($/lb) [$/kg]	0.0035 [0.0077]	0.0035 [0.0077]	0.0035 [0.0077]
Energy cost (mills/kWh)	25	25	25
Reagent cost ($/ton) [$/kg]	15 [0.017]	50 [0.055]	50 [0.055]
Reagent inventory (days)	30	30	30
Reagent purity (%)	95.3	90.0	94.0 (lime)
LSFO waste ponding cost ($/ton) [$/kg]	30 [0.033]	NA[a]	NA
LSFO gypsum stacking cost ($/ton) [$/kg]	6 [0.007]	NA	NA
Gypsum byproduct credit ($/ton) [$/kg]	2 [0.002]	NA	2 [0.002]
LSD waste disposal cost ($/ton) [$/kg]	NA	12 [0.013]	NA

[a]NA = not applicable.

Table 5. Model validation summary for LSFO FGD (1994 dollars)

Plant/Unit(s)	Unit Capacity, MW_e	Coal S, wt %	Absorbers	Model Cost, $/kW	Reported Cost,[a] $/kW	Deviation,[b] %
Petersburg/1	239	3.5	1	400	317	+26.2
Cumberland/1	1300	4.0	3	164	200	-18.0
Conemaugh/1&2	1700	2.8	5	174	195	-10.8
Ghent/4	511	3.5	3	213	215	-0.1
Bailly/7&8	600	4.5	1	189	180	+5.0
Milliken/1&2	316	3.2	1	368	348	+5.7
Navajo/1	750	0.75	2	226	236	-4.2

[a]Reported costs are from ref 25; [b]Deviation, % = (model cost − reported cost)/reported cost × 100

The simplified LSFO and LSD cost models were then further adjusted with cost-effective design choices to arrive at the cost models for respective state-of-the-art applications. These design choices, developed from information available on commercial applications, are shown in Table 7. It is recognized, however, that alternate design decisions may be made in the interest of reducing site-specific costs.

The MEL Cost Model

In the MEL process, sorbent (magnesium-enhanced slurry) is prepared in a manner similar 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, an MEL absorber is smaller than a corresponding LSFO absorber. Further, it was assumed that a state-of-the-art MEL system's waste-handling equipment would include the ability to produce gypsum byproduct and would operate in a fashion similar to that in LSFO. Considering these characteristics of MEL, for costing purposes, this process was considered to be a combination of LSFO and LSD. Accordingly, the simplified LSFO and LSD models described previously were appropriately combined to derive an MEL cost model. This model was developed for medium difficulty retrofits. The model was then further adjusted with the cost-effective design choices shown in Table 7 to arrive at a cost model for state-of-the-art MEL applications.

Table 6. Model validation summary for LSD FGD (1994 dollars)

Plant/Unit(s)	Unit Capacity, MW_e	Coal S, wt %	Absorbers	Model Cost, $/kW	Reported Cost,[a] $/kW	Deviation,[b] %
H.L. Spurlock/2	508	3.6	4	222	189	17.5
Wyodak/1	362	0.8	3	203	172	18.0
North Valmy/2	267	0.5	3	205	231	-11.3

[a]Reported costs are from ref 27; [b]Deviation, % = (model cost − reported cost)/reported cost × 100

FGD Technology Costs

Coal sulfur content and plant size are important considerations in FGD technology applications. Therefore, it is useful to examine the effects of these variables on the costs of state-of-the-art LSFO, LSD, and MEL applications. Both capital cost (in $/kW) and total annualized cost (in mills/kWh) were examined for each technology. Note that in calculation of total annualized costs, a capital charge factor of 0.104[30] and a capacity factor of 90% were used.

The effect of coal sulfur content on capital cost of state-of-the-art LSFO, LSD, and MEL technology applications is shown in Figure 6 for a 250-MW_e plant operating with a heat rate of 10,500 Btu/kWh and firing a coal with heating value of 11,900 Btu/lb (2.77×10^7 J/kg). As seen in this figure, both LSFO and MEL capital costs are higher than the capital cost for LSD across the range of coal sulfur content. Also, MEL capital cost is lower than that for LSFO across the range of coal sulfur content. These results are consistent with the fact that, in general, the amount of hardware used decreases from LSFO to MEL to LSD.

The corresponding predictions of total annualized cost are shown in Figure 7. As seen in this figure, lower cost is predicted for LSD compared with that for LSFO for up to ~2.5% coal sulfur. Also, lower cost is predicted for LSD compared with that for MEL for up to ~1.5% coal sulfur. Moreover, total annualized costs of MEL and LSFO are within ~10% of each other across the range of coal sulfur content considered. Based on predictions of total annualized costs, a plant may install either the LSD or MEL system up to ~2.5% sulfur. Beyond this sulfur content, the plant may install either MEL or LSFO.

LSD is most often installed on plants burning low- to medium-sulfur coals, while LSFO and MEL can be used by plants firing coals with a wider range of sulfur content. Therefore, coal sulfur content of 2% was selected as a common basis to examine the effect of plant size on costs of state-of-the-art LSFO, LSD, and MEL technology applications. The model predictions for capital cost and total annualized cost for plants ranging from 100 to 2000 MW_e are shown in Figures 8 and 9, respectively. Note that the discontinuities in the cost curves shown in Figures 8 and 9 result from

Table 7. Cost-effective design choices made to arrive at state-of-the-art cost models.

Parameter	Value(s) or Choice		
	LSFO	LSD	MEL
Maximum absorber size (MW_e)	900[a]	275[b]	714[c]
Material of construction	RLCS[d] or alloy	RLCS	RLCS or alloy
DBA[e] addition	Yes	N/A[f]	N/A
L/G[g] (gal/1000 ft^3)	70	N/A	40
SO_2 removal (%)	95	90	98
Byproduct/waste disposal	*Wallboard or* gypsum stacking	*Waste disposal is* the only choice	*Wallboard* production is the only choice

[a]Based on ref 28; [b]Based on ref 29; [c]Based on ref 18; [d]RLCS = rubber-lined carbon steel; [e]DBA = dibasic acid; [f]N/A = not applicable; [g]L/G = liquid-to-gas ratio

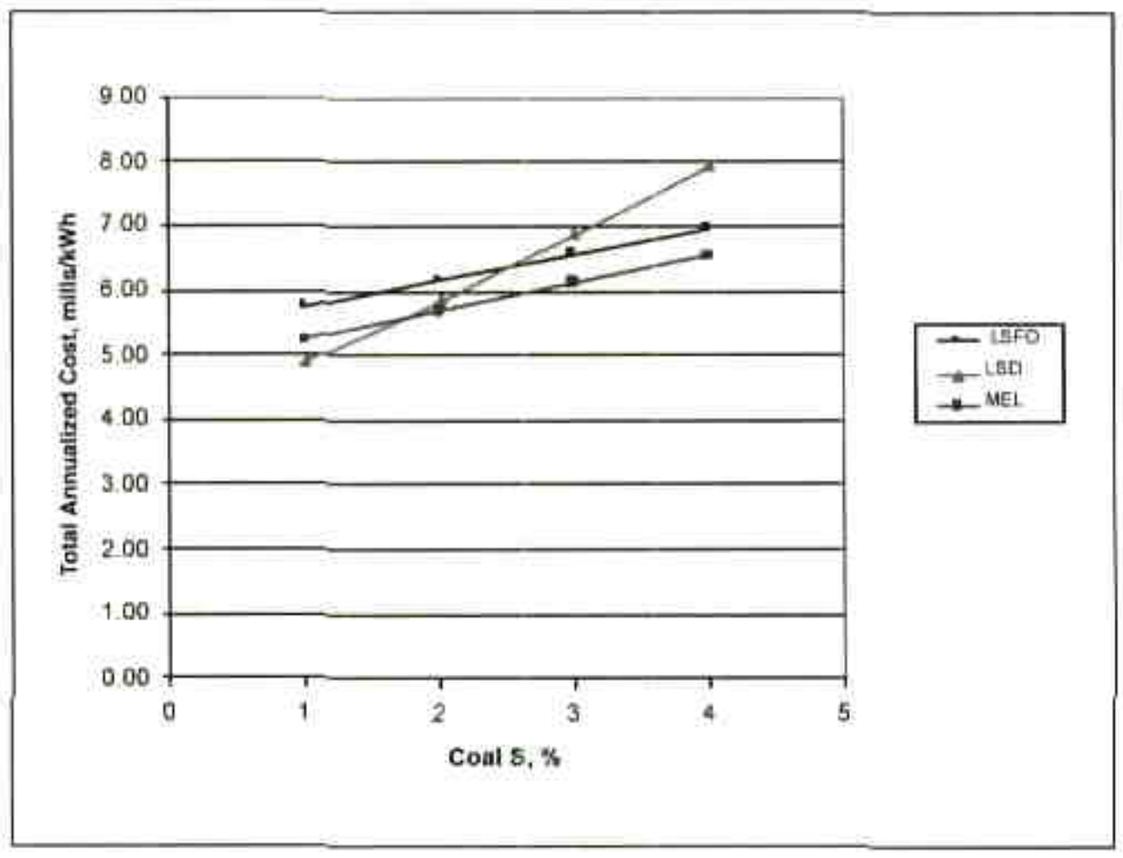

Figure 7. The effect of coal sulfur content on total annualized cost (250-MW_e plant).

limiting the maximum absorber size to correspond to 900, 275, and 714 MW_e for LSFO, LSD, and MEL, respectively.

As seen in Figure 8, the predictions of capital cost are lower for LSD compared with other technologies across the plant capacity range, consistent with the increased complexity of LSFO and MEL hardware. Further, predictions of capital cost are generally higher for LSFO compared with other technologies. The total annualized cost predictions shown in Figure 9 reflect that plants up to ~250 MW_e in size may elect to use LSD for SO_2 control. Plants larger than this may elect to use either of the wet FGD technologies.

Further understanding of technology selections based on total annualized cost and SO_2 reduction requirements may be gained by considering the results of Figures 7 and 9 together. Figure 7 shows that as coal sulfur content increases above ~1.5% and then above ~2.5%, LSD starts to become more expensive than MEL and LSFO, respectively. Also, the total annualized cost curves for LSFO and MEL are within 10% of each other across the range of sulfur content considered. Based on these results and the behavior of total annualized cost curves with changes in plant size shown in Figure 9, the following general observations may be made: (1) plants up to ~250 MW_e in size and firing low- to medium-sulfur coals (i.e., coals with a sulfur content of 2% or lower) may use LSD; and (2) plants larger than 250 MW_e and firing medium- to high-sulfur coals (i.e., coals with a sulfur content of 2% or higher) may use either LSFO or MEL.

It is recognized that the aforementioned general observations are constrained by the assumptions of the respective cost models used and factors (e.g., availability of sorbent and water, cost of sorbent, market for gypsum byproduct, and SO_2 allowance market considerations). Site-specific conditions may deviate from cost model assumptions and be affected by the previously mentioned factors, thereby resulting in costs different from those presented in this work. However, the observations made

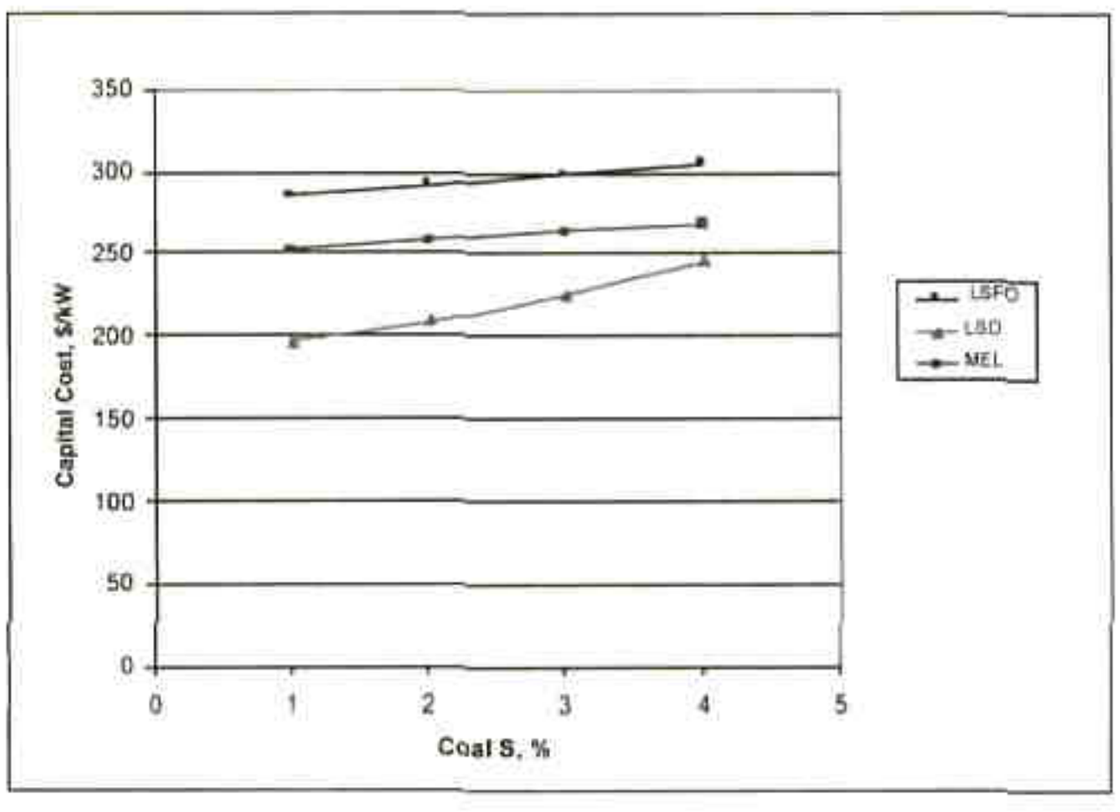

Figure 6. The effect of coal sulfur content on capital cost (250-MW_e plant).

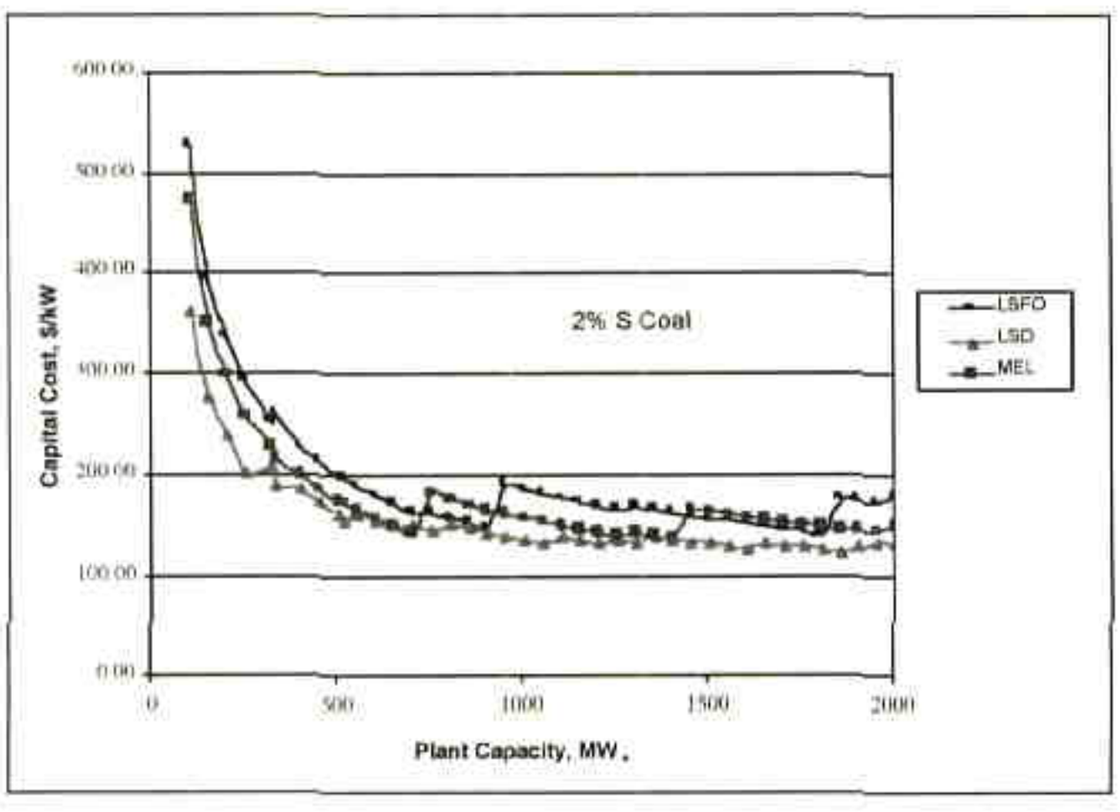

Figure 8. The effect of plant capacity on capital cost of LSFO, MEL, and LSD.

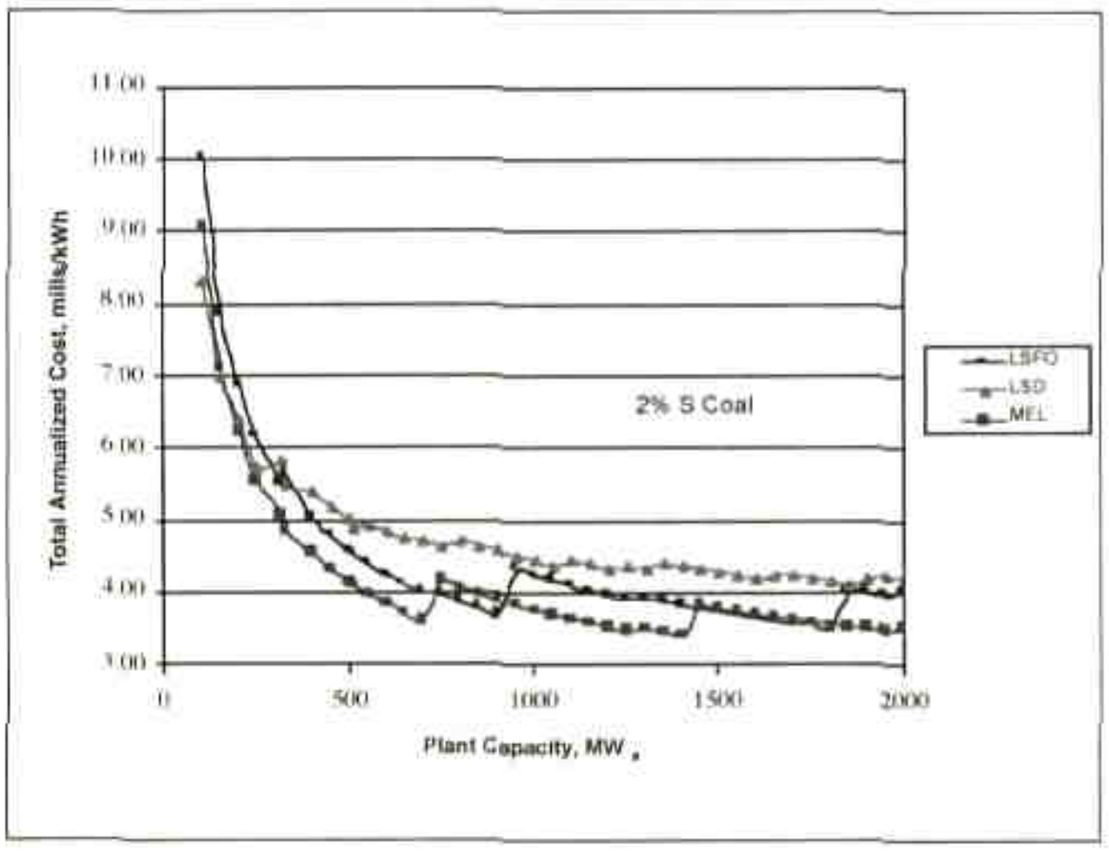

Figure 9. The effect of plant capacity on total annualized cost of LSFO, MEL, and LSD.

in the preceding discussion indicate what may be expected for medium-difficulty retrofits of state-of-the-art LSD, LSFO, and MEL applications.

SUMMARY

This paper presents a comprehensive review of the state of the art in FGD technologies for coal-fired boilers. The review describes the practical FGD processes, assesses their use, determines which of these processes dominate the FGD applications, characterizes the SO_2 reduction performance of wet limestone and LSD processes, and analyzes the costs associated with LSFO, LSD, and MEL FGD technology applications. The review of the pattern of past FGD installations in the United States and abroad reveals that wet FGD technologies, and specifically wet limestone FGD, have been predominantly selected over other FGD technologies. However, LSD is being used at the majority of the plants employing dry FGD technologies. Additional review of the U.S. FGD technology applications that began operation in 1991–1995 reveals that FGD processes of choice recently in the United States have been wet limestone FGD, MEL, and LSD. Further, of the wet limestone processes, LSFO has been used most often in recent applications.

As discussed previously, wet limestone processes (i.e., LSFO, LSIO, JBR, and natural oxidation) and LSD represent the most widely applied FGD technologies. As such, it is useful to assess the SO_2 removal performance potential of these technologies. A review of data reflects that most wet limestone systems appear to be designed for 90% SO_2 removal; however, the state-of-the-art wet scrubbers are capable of routinely achieving SO_2 removal efficiencies of more than 95%. The data also reflect that, while the median design efficiency for all units using LSD is 90%, all spray dryers installed during 1991–1995 have a design SO_2 removal efficiency between 90 and 95%. Finally, the data reveal that SO_2 removal efficiencies for wet limestone and LSD applications have improved with time.

Costs associated with state-of-the-art applications of LSFO, MEL, and LSD technologies have been analyzed with appropriate cost models. Analyses indicate that the capital cost of an LSD system is always lower than those of same-size LSFO and MEL systems, reflective of the relatively less complex hardware used in LSD. Analyses also reflect that, based on total annualized cost and SO_2 removal requirements, plants up to ~250 MW_e in size and firing low- to medium-sulfur coals (i.e., coals with a sulfur content of 2% or lower) may use LSD; and plants larger than 250 MW_e and firing medium- to high-sulfur coals (i.e., coals with a sulfur content of 2% or higher) may use either LSFO or MEL.

It is recognized that the previously mentioned general observations are constrained by the assumptions of the respective cost models used and factors (e.g., availability of sorbent and water, cost of sorbent, market for gypsum byproduct, and SO_2 allowance market considerations). Site-specific conditions may deviate from cost model assumptions and be affected by the aforementioned factors, thereby resulting in costs different from those presented in this work. However, the observations made in the preceding discussion indicate what may be expected for medium-difficulty retrofits of state-of-the-art LSD, LSFO, and MEL applications.

ACKNOWLEDGMENTS/DISCLAIMER

The research described in this article has been reviewed by the Air Pollution Prevention and Control Division, EPA, and approved for publication. The contents of this article should not be construed to represent Agency policy, nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

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