FLUE GAS DESULFURIZATION TECHNOLOGY EVALUATION
Dry Lime vs. Wet Limestone FGD
PROJECT NUMBER 11311-001

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1. INTRODUCTION

Wet flue gas desulfurization (FGD) has its roots before the Second World War. With clean air legislation in 1970, wet FGD became the process of choice because it was the only option that had been developed. The lime spray dryer (LSD) FGD technology soon became available, but its focus was on low-sulfur fuels and low removal requirements. By the late 1980s, another dry (actually moist) technology, the circulating dry scrubber (CDS) was introduced. Over the past 15 years, these dry technologies have been applied to ever-higher sulfur fuels and in ever-higher SO_2 removal applications, with good success. Over this same period of time, a main attraction for wet limestone FGD, the ability to make a commercial product (gypsum wallboard) from its by-product has begun to run into regional market saturation. It has become common that new wet limestone FGD systems are being built to produce gypsum for disposal. This paper investigates the current comparative status of three FGD systems:

- Lime spray dryer FGD (LSD)
- Circulating dry scrubber (CDS, or fluidized bed FGD)
- Wet limestone FGD with forced oxidation to produce gypsum (LSFO)

As long as there have been choices in FGD, the properties of the coal to be burned at the power plant have had a strong influence on the selection. This study compares FGD selection for three coals:

- Low-Sulfur Appalachian Coal (2.00 lb/MBtu)
- Medium-Sulfur Appalachian Coal (3.00 lb/MBtu)
- Low-Sulfur Powder River Basin Coal (1.44 lb/MBtu)

Much study work has been done based on the common 500 MW power plant size. In previous studies for National Lime Association, S&L has used this basis as well. However, current state of the art for dry systems has produced a largest absorber module of 400 MW. For a 500 MW plant, this places the dry technologies at a cost disadvantage that does not exist for developers planning a plant 400 MW or smaller. For this reason, the study compares the choices at two sizes:

- 500 MW
- 400 MW

Finally, for each case, a sensitivity analysis was performed with respect to the by-products:

- Commercial use of by-product
- Disposal of by-product

No study of this type would be complete without providing information on capital cost of the technologies. However, this report comes at a time of unprecedented turmoil in the cost arena. FGD costs have been severely impacted by several influences:

- General inflation
- Structural steel marketplace
- Structural nickel alloy marketplace
- Industrial plastic resin marketplace
- FGD supplier marketplace

As a result, FGD prices have seen a minimum of 25% inflation in the past year. Some recent contracts have been signed at prices over 300% higher than the market of 5 years ago. The costs have been prepared on a consistent, uniform basis and show a level that some buyers achieved in mid-2006. Sargent & Lundy cautions the reader that the costs provided herein are not indicative of any cost you may actually achieve. However, we believe the costs are valid for comparative purposes. These costs should *not* be used for any of these purposes:

- Planning the cost of a FGD project
- Budget requests or allocations
- Solicitation of pollution control bonds

In today's marketplace, it is impossible to determine capital cost of an FGD system until the contract is signed with the supplier.

This report is organized as follows. First, a generic description of each FGD system is presented, which serves as a backdrop for the comparison of the specific designs that follow. For the wet FGD systems, limestone forced oxidation (LSFO) technology is used as the basis of comparison with the dry systems. Next, an overview of the advantages and disadvantages of dry vs. wet limestone FGD systems is provided. This is followed by a description of the three types of fuel, as well as the operating parameters of the 400 and 500 MW units for which the dry and wet limestone FGD systems are evaluated. Major equipment and operating characteristics of the dry FGD systems and LSFO technology are presented next, followed by a discussion of key operating cost drivers (e.g., reagent, energy consumption) and the higher costs associated with retrofit vs. new units. Finally, detailed capital, operating and levelized costs are presented for the 500 MW and 400 MW systems.

2. FLUE GAS DESULFURIZATION DESCRIPTION

2.1 DRY FGD SYSTEMS

The term "Dry FGD" covers a range of technologies that include adsorption of sulfur dioxide (SO_2) on dry particles, sorption of SO_2 on dry particles in a humid environment, and sorption of SO_2 on moist particles in the process of drying. This evaluation includes spray dryer FGD (moist particles) and circulating dry scrubber (humid environment). The circulating dry scrubber (CDS) is also commonly known as a circulating fluidized bed (CFB) scrubber as well as by trade names, such as Turbosorp®.

The spray dryer FGD technology was developed beginning in 1977, to serve users of low-sulfur coal, for whom the new regulations meant they had to scrub, but only to 70% SO_2 removal. As experience was gained, the technology was shown to be capable of much higher performance. Gradually, it has been adopted by users of coal with higher and higher sulfur content, to where it is frequently considered for medium sulfur coal, depending upon the severity of the emission requirement. Maximum absorber size has grown from the early 150 MW units to 400 MW today.

Likewise, the CDS technology was developed beginning in 1984, to serve small boilers and incinerators, where wet limestone FGD would be too capital cost intensive. As with the spray dryers, experience showed the CDS technology to be capable of much higher performance than was originally thought. Gradually, it has been adopted by users with higher sulfur coal and maximum absorber size has grown to 330 MW (sold), 440 MW (bid).

Dry FGD systems are typically located after the air preheater, but before the particulate collector. The by-product is collected in either a baghouse or electrostatic precipitator (ESP). However, to achieve sulfur dioxide (SO_2) reduction above 80% with good reagent utilization, dry scrubbers are generally followed by a baghouse. The lime spray dryer (LSD) FGD and the circulating dry scrubber (CDS) FGD technologies are described in more detail below.

2.1.1 Lime Spray Dryer FGD

Although the FGD process can be done with either (granular) hydrated lime, $Ca(OH)_2$, or with (pebble) quicklime, CaO, the quantities are large, so the more economical quicklime is generally used. Preparation of the lime slurry reagent involves slaking the quicklime, either in a conventional lime slaker with a high efficiency grit removal and lime recovery system or in a ball mill slaker system. While ball mill slakers typically require higher capital than retention/detention type slakers, they produce a more uniform lime slurry that minimizes operational issues associated with insoluble grit that is inherent in all quicklime products and results in increased reagent consumption. For the spray dryer FGD system, the ball mill slaker is assumed for this report. Slaked lime is stored in an agitated tank for use. The slurry reagent is fed to the absorber to replenish lime consumed in the reaction; the feed rate is typically controlled based on the removal efficiency required.

The spray dryer is a large, empty vessel where the flue gas is directed at high velocity toward the lime slurry atomizers. The atomizer is a device that uses high energy to divide the slurry into extremely fine droplets and hurl them into the high-velocity gas stream. The result is three nearly simultaneous (though incomplete) activities: Sulfur dioxide is absorbed into the droplets; the sulfur dioxide reacts with the lime to form a mixture of calcium sulfite and calcium sulfate, and the droplet dries. The resulting particles are carried out of the absorber in the flue gas stream and collected in the baghouse. Depending upon the supplier, spray dryer FGD systems may be supplied with either of two atomization systems. Rotary atomizers use a high-speed (~10,000 rpm) wheel (like a large, vertical-shaft brake drum). Ceramic-lined nozzles are cut into the hub of the wheel, such that when slurry is poured inside the wheel, the speed of the wheel imparts the necessary energy to the liquid spraying out of the hub nozzles. On the other hand, dual-fluid nozzles consist of concentric tubes carrying slurry and compressed air. When the slurry reaches the nozzle at the end of the assembly, the compressed air imparts the energy necessary to break the slurry into very fine droplets. Both systems have substantial experience in the industry. In the process of evaporating the droplets, the flue gas is cooled from the absorber inlet temperature of approximately 300°F to an outlet temperature of 160°F to 180°F. The actual outlet temperature is chosen based on the optimum reagent utilization, the flue gas saturation temperature, and the required SO_2 removal efficiency.

To summarize, the droplets absorb SO_2 from the gas and the SO_2 reacts with the lime in the slurry. Desulfurized flue gas, along with reaction products ($CaSO_3$, $CaSO_4$), excess lime ($Ca(OH)_2$), and fly ash passes out of the dry scrubber to the baghouse. The primary reaction in the spray dryer FGD is as follows:

$$SO_2 + CaO + 1/2\ H_2O \Rightarrow CaSO_3 \bullet 1/2\ H_2O$$

A part of the $CaSO_3$ reacts with oxygen in the flue gas to form calcium sulfate ($CaSO_4$):

$$CaSO_3 + ½O_2 + 2H_2O \Rightarrow CaSO_4 \bullet 2H_2O$$

The first generation of LSD systems was designed to achieve 70% SO_2 reduction efficiencies. This was done primarily to comply with the New Source Performance Standards for low-sulfur coals. However, further experience with Powder River Basin (PRB) coal has prompted equipment suppliers to guarantee SO_2 reduction efficiencies of up to 95% or as low as 0.06 lb/MBtu. The figure below represents the maximum expected SO_2 reduction for a LSD system with baghouse as it relates to the sulfur content in the coal. It represents a general guideline for application of LSD technology, and is derived from Sargent & Lundy's in-house database on the technology performance, as obtained from various suppliers of FGD systems.

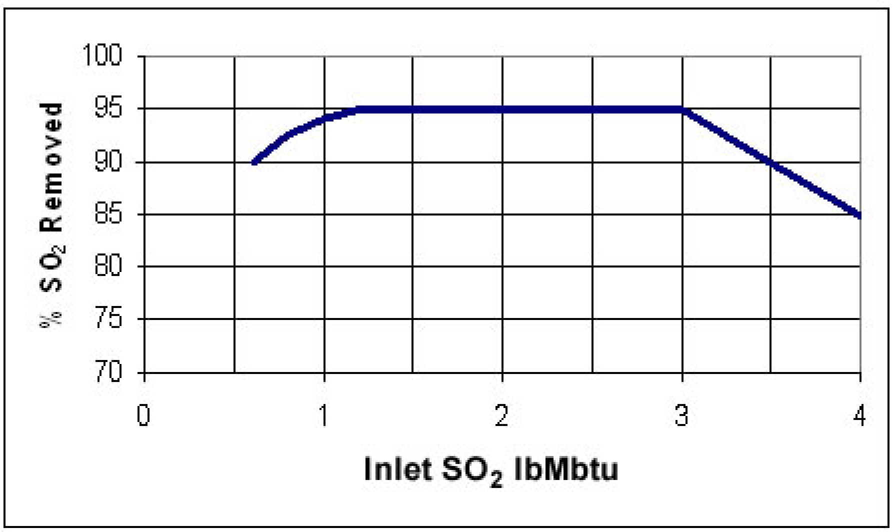

FIG 2.1-1 RELATION BETWEEN INLET SO_2 AND SO_2 REMOVAL EFFICIENCY FOR LSD

2.1.2 Circulating Dry Scrubber

In a CDS system, flue gas is treated by injecting a mixture of dry hydrated lime or lime slurry and recycled by-product (a mixture of reaction products ($CaSO_3$, $CaSO_4$), excess lime ($Ca(OH)_2$), and fly ash). Preparation of hydrated lime requires the use of a lime hydrator (an atmospheric hydrator is analyzed in this report). Although hydrated lime can be purchased, converting commercial quicklime into the hydrated lime on-site typically offers a lower cost alternative, particularly for larger units or multiple unit systems, and thus is the option presented in this report. The reagent is fed to the absorber to replenish hydrated lime consumed in the reaction. Like LSD systems, the feed rate is typically controlled based on the SO_2 removal efficiency required.

The CDS technology is similar to other wet and dry FGD technologies in that solids are continuously recycled to the absorber to achieve high utilization of the reagent. However, the CDS technology has a distinctive feature in that reactive material also recirculates within the absorber to achieve a high retention time, and for this reason is called Circulating Dry Scrubber. It is this circulation that makes high removal efficiency possible with the CDS process. The characteristic fluidized bed is established by a bank of venturis that increase the flue gas velocity at the entrance to the reactor.

Water is injected in the scrubber near the outlet of the venturis to humidify the flue gas and promote the absorption of SO_2 from the flue gas and the reaction of SO_2 with the lime. Humidification is controlled to maintain a flue gas temperature of typically 160°F or approximately 30°F above the adiabatic saturation temperature of the flue gas. Hydrated lime absorbs SO_2 from the gas and forms calcium sulfite and calcium sulfate. Similar to spray dryer systems, the desulfurized flue gas, along with reaction products ($CaSO_3$, $CaSO_4$), excess lime ($Ca(OH)_2$), and fly ash passes out of the fluidized bed to the baghouse.

SO_2 absorbed in the droplet reacts with hydrated lime in the circulating fluidized bed to form calcium sulfite ($CaSO_3$) in the following reaction:

$$SO_2 + Ca(OH)_2 \Rightarrow CaSO_3 \bullet 1/2\ H_2O + ½\ H_2O$$

A part of the $CaSO_3 \bullet 1/2\ H_2O$ reacts with oxygen in the flue gas to form calcium sulfate ($CaSO_4$):

$$CaSO_3 \bullet 1/2\ H_2O + ½O_2 \Rightarrow CaSO_4 \bullet 2H_2O$$

For CDS systems, SO_2 removal guarantees of 95-98% are available from the system suppliers identified below. The CDS technology has been demonstrated to achieve these levels of SO_2 reduction efficiency albeit with a significant degradation in reagent utilization, as discussed in Section 5.1, Reagent Cost.

2.1.3 Commercial Status

LSD systems are in operation at many facilities, ranging in size from less than 10 MW to over 600 MW. Historically, multiple modules have been used for plants greater than 300 MW in capacity. However, a recent retrofit project using PRB coal has been awarded which has a set of 2 modules at approximately 400 MW each, scheduled for start-up in 2008. For eastern bituminous coals, some FGD vendors have proposed modules for units sized up to 350 MW. Applications include commercial units with coal sulfur as high as 2.0%, depending on the degree of removal required. LSD systems are available from a number of vendors with varying levels of U.S. utility experience including:

- Alstom Power Environmental (formerly ABB Environmental, Flakt, Rockwell) (rotary)
- Anhydro, ℅ Hamon Research-Cottrell (rotary)
- Fisia Babcock GmbH, ℅ Babcock Power Environmental (dual-fluid nozzle)
  Note: Fisia Babcock is the Niro Atomizer licensee for European applications.
- GEA Niro Atomizer, ℅ Babcock & Wilcox (rotary)
- Wheelabrator Air Pollution Control (Siemens) (dual-fluid nozzle)

CDS technology, by contrast, has a limited number of applications in the utility industry, although this number is growing, particularly in China. Facilities range in size from less than 10 MW to 350 MW. The largest CDS units are in China, with twenty 290 to 350 MW recently installed, and designed for SO_2 capture capability of greater than 90% on up to 3% sulfur coals. The suppliers claim to see no technical obstacle to a single-module CDS absorber up to 700 MW, but commercial considerations limit their offerings to sizes near what they have built.

In the United States, CDS applications are limited to two small units burning low sulfur coal, plus a 2 x 250 MW installation in Puerto Rico. Three additional units (104 MW, 150 MW (equivalent) and 2 x 330 MW) are currently under construction in the US. However, interest in CDS technology is increasing due to the technology's ability to accommodate sulfur peaks with high SO_2 removal. Three process developers offer minor variations on the CDS technology:

- Allied Environmental Solutions, Inc. (formerly Lurgi) "CDS-FGD"
- Austrian Energy and Environment, % Babcock Power Environmental; "Turbosorp® FGD"
- Wulff Deutschland GmbH, % Nooter/Eriksen; "Graf/Wulff."

Each of these suppliers has significant experience in power plant applications, although for reasons not related to the technology they have not been widely applied in the U.S.

2.2 WET LIMESTONE FGD SYSTEMS

Wet limestone FGD systems remove the sulfur dioxide (SO_2) from the flue gas by passing the flue gas through a chamber that exposes the flue gas to a slurry of finely ground limestone. The slurry absorbs the SO_2 from the flue gas and the calcium in the limestone reacts with the SO_2 to form a mixture of calcium sulfite and calcium sulfate.

Limestone for the wet limestone FGD process is usually received as gravel. Preparation of the limestone slurry involves grinding the limestone extremely finely in a horizontal ball mill containing water. The slurry is pumped through banks of spray nozzles to create fine droplets to facilitate intimate and uniform contact with the updraft of flue gas.

After absorbing the sulfur dioxide from the flue gas, the slurry collects in the bottom of the absorber in a reaction tank, where it is aerated to oxidize bisulfite ion to sulfate and avoid formation of calcium sulfite hemihydrate ($CaSO_3 \bullet \frac{1}{2} H_2O$) in favor of producing calcium sulfate dihydrate ($CaSO_4 \bullet 2H_2O$), or gypsum, which precipitates. This is where the term "forced oxidation" originates. Oxidized slurry is then recirculated to the spray headers. A portion of the slurry is withdrawn to remove the precipitated gypsum. Typically, the slurry is dewatered in a two-stage process involving a hydroclone and vacuum filter system to produce a gypsum cake for disposal or sale. Water removed from the slurry is returned to the process. A portion of this water is removed from the system as wastewater to limit accumulation of chloride salts and other undesirable constituents introduced with the coal.

Unlike dry FGD systems, an LSFO system is typically located after the particulate matter removal device (baghouse or electrostatic precipitator). Cleaned gas is discharged directly to the stack with no further treatment.

At one time, wet limestone FGD systems were applied to coals of all sulfur levels, including low sulfur coals (<1%) in the western United States. However, as the lime spray dryer technology became proven, that became the predominant choice for these coals. Today, wet limestone FGD systems are typically installed on medium- to high-sulfur fuels (>2%), where the relatively low cost of reagent can pay back the higher initial capital cost.

For wet limestone forced oxidation FGD systems, SO_2 removal guarantees of up to 99% (without additives) are available from the system suppliers and have been demonstrated in commercial applications, though there is a practical outlet limitation at 0.04 lb. SO_2/MBtu, which represents a lower percentage removal for the lowest sulfur coals.

For more detail on wet FGD systems, refer to the Wet FGD Technology Evaluation (2007) prepared for NLA by Sargent and Lundy [to be posted on NLA's website Summer 2007]. Here, only those aspects that affect comparison with dry FGD will be noted.

Wet limestone FGD is the predominant technology for large-scale utility applications in most parts of the world. Absorber size ranges from less than 100 MW to more than 1,000 MW, with 400 to 500 MW absorbers being common in every supplier's experience. Nearly 20 suppliers have provided major systems over the last 25 years, with at least eight of those currently doing credible business in the United States today:

- Advatech (J/V of URS, Mitsubishi Heavy Industries America)
- Alstom Power Environmental (formerly ABB Environmental, Combustion Engineering, Peabody)
- Babcock & Wilcox
- Chiyoda, % Black & Veatch
- Fisia Babcock GmbH, % Babcock Power Environmental
- Hitachi America, Ltd.
- Marsulex Environmental Technology (formerly GE Environmental Services, Inc.)
- Wheelabrator Air Pollution Control (Siemens)

3. COMPARISON OF DRY AND WET LIMESTONE FGD TECHNOLOGIES

3.1 DRY PROCESS ADVANTAGES

Dry FGD systems have the following advantages when compared to wet limestone forced oxidation (LSFO) FGD technology:

1. The absorber vessel can be constructed of unlined carbon steel, as opposed to lined carbon steel or solid alloy construction for wet limestone FGD. Therefore, for systems using a single absorber (i.e., up to 400 MW), the capital cost is typically lower than for wet limestone FGD.

2. Overall power consumption is lower for dry technologies than for wet limestone FGD systems.

3. By-product produced is in a dry form and can be handled with conventional pneumatic fly ash handling equipment.

4. The by-product is stable for most landfill purposes and in most cases can be disposed of concurrently with fly ash.

5. The dry system is a less complicated process and generally has less equipment than does the wet limestone FGD system. This results in lower O&M labor and maintenance material requirements.

6. Pressure drop across a spray dryer absorber is typically lower than that across wet limestone FGD absorbers.

7. High chloride levels in the fuel improve, rather than hinder, SO_2 removal performance.

8. Sulfur trioxide (SO_3), which condenses to sulfuric acid aerosols in an FGD system, is removed efficiently (greater than 90%) with a dry FGD baghouse. Wet scrubbers have less affinity for acid mist and, according to FGD suppliers, they typically capture between 25% and 50% of sulfuric acid aerosols. Alkali injection upstream of a particulate collection system or even the addition of a wet electrostatic precipitator would be required for a wet

limestone FGD system to achieve the same level of acid mist removal as the inherent removal in a dry FGD system. The emission of sulfuric acid mist, if above a threshold value of 7-10 ppm, may result in a pale blue plume visible after the water vapor plume dissipates.

9. Flue gas following a dry scrubber is not saturated with water (being 30°F to 50°F above the saturation temperature), which reduces or eliminates the visible moisture plume seen from the stack of a wet limestone FGD. Wet limestone scrubbers produce flue gas that is saturated with water, which would require some form of reheater or a gas-gas heat exchanger to increase the flue gas temperature to achieve a dry stack and nearly-invisible plume. Due to the high costs associated with heating the flue gas, all recent wet limestone FGD systems in the United States have used wet stacks. Avoidance of the expensive wet stack liner and/or flue gas reheat is another capital cost advantage to dry FGD systems.

10. There is no liquid waste from a dry FGD system, whereas wet limestone systems typically produce a liquid waste stream, depending on the disposition of the by-product. In some cases, a wastewater treatment plant must be installed to treat the liquid waste prior to discharge. The wastewater treatment plant produces a small volume of solid waste, rich in heavy metals (including oxidized mercury) that must be disposed of in a special landfill. On the other hand, a dry FGD system can provide an outlet for process wastewater from other parts of the power plant. Since the dry FGD system evaporates all its water, the heavy metals, including oxidized mercury, report to the FGD by-product. The by-product tends to trap the trace metals and the concentrations are low enough to qualify the by-product for a standard landfill.

3.2 DRY PROCESS DISADVANTAGES

Dry FGD systems have the following disadvantages when compared to wet limestone FGD technology:

1. The largest absorber module used in the industry is on the order of 400 MW in capacity. For units larger than 400 MW, the need for multiple modules typically causes a dry process to be

more expensive than a single-module wet limestone FGD process. This will also result in a more costly inlet and outlet ductwork and damper configuration.

2. The dry systems use a more expensive reagent (quicklime or hydrated lime) than limestone-based FGD systems (more dollars per mole of calcium), and the reagent must be stored in a steel or concrete silo, rather than a pile.

3. Reagent utilization for a dry system is poorer compared to a wet limestone system, e.g., the reagent stoichiometric ratio for a dry FGD system is higher than the stoichiometric ratio for a wet limestone FGD system (on the same basis) to achieve comparable SO_2 removal (more moles of calcium per mole of sulfur dioxide).

4. Dry FGD produces a by-product that has fewer uses due to its properties, i.e., mixed ash/by-product, permeability, soluble products, such as calcium chloride. Although development continues to seek useful applications of the by-product, most current reuse of the material is for reclamation at mine-mouth plants. Wet limestone FGD, on the other hand, can produce commercial-grade gypsum for use in wallboard, cement or agricultural applications. However, the market for wallboard-grade gypsum is becoming saturated in many locales, which diminishes this opportunity.

5. Combined removal of fly ash and by-product solids in the particulate collection system precludes commercial sale of fly ash unless the unit is designed to remove FGD by-product and fly ash separately. In some cases, dry FGD could be retrofitted after an existing ESP, which would allow the sale of fly ash, but would also require operation of two particulate collection devices.

6. The CDS process is best suited to base-load applications, as high flue gas velocities are required to maintain the fluidized bed. If turn-down is to be accommodated, a gas recirculation system is required to maintain fluidization velocity at reduced loads.

4. DESIGN BASIS OF SYSTEMS ANALYZED IN REPORT

4.1 DESIGN CRITERIA

Table 4.1-1 lists the design basis for the FGD systems analyzed in this report.

TABLE 4.1-1

STUDY DESIGN BASIS

Fuel	Low sulfur Appalachian		Medium sulfur-Appalachian		Low sulfur Powder River Basin	
Fuel analysis, % wt.:						
Moisture	6.0		6.0		30.4	
Ash	9.1		9.1		6.4	
Carbon	72.6		72.6		47.8	
Hydrogen	4.8		4.8		3.4	
Nitrogen	1.4		1.4		0.7	
Sulfur	1.3		2.0		0.6	
Oxygen	4.7		4.7		10.8	
Chlorine	0.1		0.1		0.03	
High heating value, Btu/lb	13,100		13,100		8,335	
SO_2 generation, lb/Mbtu	2.0		3.0		1.44	
Unit size, MW	**400**	**500**	**400**	**500**	**400**	**500**
Heat input to boiler, MBtu/hr	4,000	5,000	4,000	5,000	4,149	5,186
Coal feed rate, tons/hr	153	191	153	191	249	311
Flue gas flow at FGD inlet, Macfm						
• Dry systems	1.43	1.79	1.43	1.79	1.58	1.97
• LSFO	1.36	1.70	1.36	1.70	1.50	1.87
Flue gas temperature at FGD inlet, °F	280					
Flue gas flow at FGD outlet, Macfm						
• Dry systems	1.28	1.60	1.28	1.60	1.40	1.75
• LSFO	1.20	1.50	1.20	1.50	1.31	1.64
Flue gas temperature at FGD outlet, °F						
• Dry systems	160		160		165	
• LSFO	130		130		130	
SO_2 outlet, lb/MBtu						
• Lime spray dryer	0.10		0.15		0.072	
• Circulating dry scrubber	0.04		0.06		0.04	
• LSFO	0.04		0.06		0.04	
SO_2 reduction efficiency, %						
• Lime spray dryer	95		95		95	
• CDS	98		98		97.2	
• LSFO	98		98		97.2	

Note that a significant difference between the lime spray dryer (LSD) and circulating dry scrubber (CDS) is that the LSD is not capable of SO_2 removal efficiency in excess of 95%. More and more permits are requiring higher removal efficiency. The study design basis is for 98% removal with the CDS and LSFO processes, but 95% for the LSD process with the cost analysis including purchase of allowances for the difference between 95 and 98 % SO_2 removal.

4.2 SUBSYSTEMS FOR DRY FGD

The overall designs of the lime spray dryer (LSD) and circulating dry scrubber (CDS) systems evaluated in this report are similar to one another, with the principal differences being:

- Lime slurry preparation: a ball mill slaker to prepare quicklime for the LSD vs. an atmospheric hydrator to prepare hydrated lime for the CDS
- Introduction of moisture: LSD injects a slurry of water and solids vs. CDS, which injects the solids and the humidification water separately
- Type of absorber vessel: vertical open chamber for LSD vs. fluidized bed for CDS
- Recycle of solids from the baghouse to the absorber: a LSD recycles 2 to 3 times the amount of by-product produced vs. a CDS which recycles up to 100 times the by-product rate.
- Particulate control: baghouse alone for LSD vs. baghouse preceded by a pre-collector for CDS

The CDS capital cost is similar to the LSD: the lower cost slaking system and the avoidance of the slurry system, with its recycle silo, slurry tank(s) and slurry pumps is offset by the tall structure to support the elevated baghouse.

To assure that the FGD systems are available 100% of the time, spare (redundant) equipment is provided for numerous system components. Assumed redundancy is described along with the system descriptions below.

4.2.1 Reagent Handling and Preparation

For dry FGD systems, it is assumed that quicklime is received by truck and pneumatically conveyed to storage. Silo storage capacity is assumed to be 14 days, and from there the lime is pneumatically conveyed to a 24-hour capacity day bin.

For the LSD, the day bin and a gravimetric feeder supplies the lime to a 2 x 100% slaking system. A horizontal ball mill lime slaker system is used. Two 100% capacity slurry transfer pumps are used to provide high reliability to transfer the slurry to the slurry tank. Process makeup water is added to the slaker to produce 20% solids slurry.

If the process requires a more dilute slurry, water is added on line, prior to injection into the absorber. The slurry is fed to the absorber by a dedicated reagent feed pump (100% spare capacity provided).

Most CDS systems utilize dry calcium hydroxide, which can be either purchased directly from commercial suppliers or produced at the plant by the use of on-site hydrator systems. For smaller systems, it is possible that commercial suppliers can produce hydrate products more economically than can be produced on-site in limited quantities. Since the design basis for this example is for a relatively large unit, the assumption has been made that all hydrate for the system will be produced on-site. For this approach quicklime is conveyed from a day bin by a gravimetric feeder to a 2 x 100% atmospheric hydrating system. Water is added to the hydrator to convert the quicklime into hydrated lime. Hydrated lime is pneumatically conveyed to a silo (24-hour capacity), and is fed to the CDS absorber with a rotary screw feeder.

A third option for providing hydrated lime for a CDS system is to size the hydration system to supply all hydrated lime needs, but not expend the capital to have 100% back-up on this system. Unless the system is in a very remote area, back-up supply for hydrated lime can typically be secured from a commercial supplier.

4.2.2 Absorber/Reaction System

For the 400 MW dry FGD systems, a single carbon steel absorber is provided to achieve 95% SO_2 removal efficiency for the LSD, and 98% SO_2 removal for the CDS (slightly lower with PRB coal). For the LSD unit, the absorber is a vertical, open chamber with cross-current contact between the flue gas and lime slurry. The ultrafine slurry droplets partially complete three activities before being swept out of the absorber: 1) Absorption of SO_2 from the flue gas, 2) Reaction of SO_2 with the lime in the droplet, and 3) Drying of the moisture in the droplet. The hopper in the bottom of the absorber collects large particles that may drop out in the absorber. The absorber is typically operated at a 30°F approach to adiabatic saturation temperature to assure that no particle is wet by the time it reaches the wall of the absorber or leaves the absorber, under all operating conditions. Pressure drop across the absorber is approximately 4 in. WC.

In the CDS, the absorber is a fluidized bed reactor in which the solids are fluidized by the high velocity of the flue gas. Flue gas is introduced to the absorber through a battery of venturis, to accelerate the flue gas to the necessary velocity. Humidification water is injected above the venturis using high-pressure atomizer nozzles. Hydrated lime, along with recycled by-product, is introduced just above the venturis (or in some cases, into the venturis). Co-current flow of the humidified flue gas and the solids provides long residence time and significant turbulence to enhance particle flue gas interaction. This helps achieve high SO_2 removal efficiency. The absorber is typically operated at a 30°F approach to adiabatic saturation temperature to assure that no solids will adhere to the walls of the absorber, under all operating conditions. Pressure drop across the absorber is approximately 8 in. WC.

For the 500 MW dry FGD systems, two absorbers, each treating 50% of the flue gas, are provided to achieve 95% SO_2 removal efficiency for the LSD and 98% removal for the CDS (again, slightly lower with PRB coal). This also requires inlet and outlet plenums to distribute gas to the absorbers, which drives up the capital cost, compared to a single-module arrangement. Two absorbers provides an opportunity to service one absorber while the other is providing part-load operation; however, this feature is available only if an inlet damper and an outlet damper is provided on each absorber module. For this evaluation, it has been assumed that no dampers are provided; the absorbers operate as one.

4.2.3 Baghouse

The dry FGD absorber is installed between the air heater outlet and particulate collector. Most existing units have very short ductwork between the air heater outlet and ESP inlet. This makes it very difficult to take the gas from the air heater outlet to the FGD equipment and return it to the ESP inlet. Also, most existing ESPs are not designed to handle increased particulate loading resulting from FGD waste. For these reasons, both the retrofit and new unit cases are based on flue gas from the dry FGD absorber being treated in a new baghouse.

For both dry FGD systems, a conventional pulse-jet baghouse is provided. For the LSD, an air-to-cloth ratio of 3.5 ft/min is used. The much higher recycle rate of the CDS means the dust loading is much higher, so a coarse particle knockout chamber is often provided at the entrance to the baghouse. The CDS unit is also provided with a somewhat larger baghouse (i.e., air-to-cloth ratio of 3.2 ft/min).

For both dry systems, the baghouse is provided with a spare compartment for off-line maintenance while the unit is in service. For the LSD, by-product is conveyed from the baghouse in the usual way, except the flow is split, part being conveyed to the by-product silo and part to the recycle silo. For the CDS, the large recycle volume requires simple handling. This is achieved by installing the baghouse at a higher elevation than the absorber venturis. Air slides (aerated inclines) allow gravity return of the solids to the absorber. A slipstream from the air slides fills a hopper for by-product removal. Also, the fresh lime is dropped into the air slide for injection into the absorber. The material in the by-product hopper is conveyed to the by-product silo.

In either case, the by-product silo is assumed to have a 3-day storage capacity, in accordance with typical utility design.

4.2.4 Flue Gas System/Stack

Flue gas from the air preheater is ducted to the absorbers. Gases from the absorber flow to the baghouse to collect the by-product and fly ash. The booster fan is sized to provide an additional 14 in. WC (12 in. WC operating) pressure drop for LSD, and 18 in. WC (16 in. WC operating) for CDS, through the absorber and baghouse.

Due to the dry condition of the flue gas, a carbon steel stack liner is suitable. This means that for retrofit cases, flue gas can usually be returned to the existing stack. A small allowance is included for lining a 20-foot high section of the existing stack. Although retrofit units save money on the stack, there are some special costs involved in a retrofit. Booster fans, including foundations, switchgear and cable, are a large cost item. Also, retrofit ductwork, including dampers, support structures, foundations and insulation, can be costly.

4.2.5 Baghouse Dust Handling

The by-product is collected in the baghouse. For the LSD, a portion of the by-product is stored in a recycle silo, which is then used to mix with the lime slurry. Recycling the by-product gives the residual lime a second chance to react, which improves the overall reagent utilization. Similarly, for the CDS, a portion of the by-product (fly ash, reaction products, and unreacted reagent) is recycled to the absorber with air slides to increase SO_2 capture capability and reagent utilization. The recycle ratio in the LSD falls in the range of 2–3:1, whereas the recycle ratio for a CDS can be as high as 100:1. This extreme recycle ratio is the reason some CDS suppliers provide a streamlined baghouse, and why they all use the simple air slide system for reinjection of the solids.

In either case, the fraction leaving the system is pneumatically conveyed to an elevated by-product silo. This silo fills railcars or trucks by gravity. Pug mill mixers (2 x 100%) are provided to condition (moisten) the by-product before it is loaded to control dust.

4.3 SUBSYSTEMS FOR WET LIMESTONE FGD

4.3.1 Reagent Handling and Preparation

Limestone is received by truck and stored in a 30-day capacity bulk storage limestone pile. The reclaim system includes a yard hopper with a vibrating feeder and conveyor system to transfer limestone to a 24-hour capacity day bin. The day bin and a gravimetric feeder supply the limestone to a 2 x 100% capacity horizontal ball mill system. This allows 100% availability for the limestone grinding system. Two 100% capacity classification pumps are used to provide high reliability of the classification system. Process makeup water or recycle water is added to the ball mills to produce 70% solid slurry. The slurry is diluted to 30% solids in the classification process and is stored in a 16-hour limestone slurry tank prior to being fed as reagent makeup into the absorber.

4.3.2 SO_3 Control

A new issue with wet FGD systems is the existence of sulfur trioxide (SO_3) in the flue gas entering the air quality control systems. It can lead to corrosion of particulate control systems. Then, the moisture of the wet FGD system converts the SO_3 to sulfuric acid mist, which becomes an issue leaving the stack. For the purposes of this study, it is assumed that SO_3 is not an issue because this study is limited to low- and medium-sulfur coals, which typically produce insufficient SO_3 to be a problem. Also for the purposes of this study, oxidation of SO_2 to SO_3 across an SCR system is likewise assumed to be insignificant.

4.3.3 Absorber/Reaction System

For the wet limestone FGD, a single 100% capacity absorber is provided. The absorber is a vertical open-spray tower, with countercurrent contact between the flue gas and scrubber slurry. To achieve 98% SO_2 removal, multiple spray levels are provided, each with a dedicated pump. Each individual pump is dedicated to a spray level in the absorber. One spare spray level is provided along with a dedicated pump with each design. Entrained slurry droplets are removed from the flue gas through a chevron-type mist eliminator. Scrubber slurry drains into a reaction tank located in the bottom of the absorber. The tank is agitated and is sized for a minimum of 15 hours of solids residence time. Compressed air is injected below the reaction tank agitator through a sparger network. To achieve greater than 99% oxidation of $CaSO_3$ to $CaSO_4$, a 3:1 stoichiometric ratio of oxygen to absorbed SO_2 (moles O/moles SO_2) is used to size the oxidation air compressors. Bleed slurry from the absorber is pumped to the dewatering area. Makeup limestone slurry is added to the reaction tank.

4.3.4 Particulate Control

Particulate control is a pivotal issue in selecting dry vs. wet limestone FGD systems because particulate control is not a necessary part of the wet limestone FGD process. The particulate control system is located upstream of the wet limestone FGD system, and as long as its performance is reasonable, the type of particulate control system used is of no concern to the wet limestone FGD system. In the case of new units, a particulate control system must be installed whether a dry FGD system or a wet limestone FGD system is installed, so when comparing costs between dry and wet systems, it is fair to ignore the cost of a baghouse. However, in the case of a retrofit FGD system, the existing particulate control system is frequently not suitable to support a dry FGD system. This is because retrofits are typically made to plants that have electrostatic precipitators (ESPs) that are sized too small to meet modern emission limits. In addition to being too small, ESPs produce inferior desulfurization results, as less desulfurization occurs on ESP collection plates than when the gas must pass through the filter cake on a bag. As a result, retrofit dry FGD systems must be assessed the cost of a baghouse that is not needed with a retrofit wet limestone FGD system.

4.3.5 Chimney

Unlike dry FGD systems, wet limestone FGD technology is typically installed between the outlet of the existing particulate collector and chimney. However, most retrofit wet limestone FGD units are not able to use the existing chimney, as the chimney was typically designed for hot flue gas at approximately 100 ft/sec exit velocity. To accommodate saturated flue gas from a wet limestone FGD system, wet stacks are typically designed using a corrosion-resistant flue with a gas velocity of 55 to 70 ft/sec. The lower gas velocity is required to prevent condensed moisture from being carried out the top of the stack. Most retrofit units, therefore, require a new chimney with a corrosion-resistant flue.

The gas path of the system starts at the discharge of the existing ID fans, through the new booster ID fans and absorber, and discharges into a new chimney with FRP liner. The booster ID fan is sized to provide 5 to 6 inches water column additional pressure drop through the wet limestone absorber.

For wet stack operation, traditional carbon steel or brick chimney flues are not suitable, due to acid attack. An FRP liner is used, at a cost similar to steel. However, in retrofit cases, a new chimney is usually required, as the outage to replace only the flue would take too long.

Inlet ductwork to the absorber system, including the booster ID fan, is fabricated of carbon steel. The outlet ductwork from the absorber to the stack breeching is constructed of alloy, alloy-lined carbon steel, or FRP, for corrosion protection.

4.3.6 Gypsum Dewatering/Handling and Wastewater Treatment

For the wet limestone FGD process where the gypsum is to be sold for manufacture of wallboard, the bleed stream from the reaction tanks (15% solids) is pumped to a set of hydroclones for primary dewatering. Hydroclone overflow is recycled back to the process. Hydroclone underflow, containing 50% solids, is fed to a vacuum belt filter for secondary dewatering. Two 100% capacity horizontal belt filters are provided, designed to achieve 90% solids in the gypsum by-product. Vacuum pumps draw the liquid through the fabric belt, leaving the solids on the belt. Filtrate from the belt filter is returned to the process.

Chlorine may be present in the coal, the limestone and/or the FGD makeup water. The entering chlorine shows up as chloride in the scrubbing liquor. A low level of chloride is permissible in commercial gypsum. Chloride cannot be allowed to build up in the scrubbing liquor – a maximum chloride concentration must be selected and all the materials in the FGD system must be selected to withstand the selected chloride concentration (commonly 12,000 ppm). The gypsum cake must be rinsed with fresh water on top of the filter belt to reduce residual chlorides to a level of less than 100 ppm in the dry solids. The residual chloride leaving in the cake is not sufficient to balance the incoming chloride, so a small slipstream from the filtrate is blown down to prevent chlorides from accumulating in the process. For purposes of this study, we have assumed that this chloride purge must be treated in a wastewater treatment facility to make the wastewater suitable for discharge.

The wastewater treatment concerns are usually suspended solids (gypsum fines) and heavy metals (which depend on what is in the coal). For the purposes of this study, the wastewater treatment system consists of feed systems for two different polyelectrolytes, a clarifier, and a filter press. The chemical feed systems each consist of a pad for a chemical tote and a chemical metering pump to dose the polyelectrolyte into the purge line going to the clarifier. The polyelectrolytes precipitate the heavy metals. The clarifier serves to settle out the metal precipitates and the gypsum fines. The clarifier overflow is discharged. The underflow is dewatered in the filter press. The filter cake drops into a roll-off container, which is periodically removed to a licensed landfill.

Dewatered gypsum is discharged from the belt filters to belt conveyors that transfer the gypsum to a covered storage area. The covered storage area is sized for 7 days. The gypsum is assumed to be trucked to an offsite user.

For the wet limestone FGD process where the gypsum is to be landfilled, the dewatering process is similar, with some key differences. Primary dewatering is the same, with the bleed from the reaction tanks (15% solids) pumped to hydroclones. Hydroclone overflow is recycled to the process and 50% solids hydroclone underflow is fed to a vacuum filter. However, the dryness requirement for the by-product may be only 85% solids, rather than 90%, which allows use of simpler vacuum drum filters. Two 100% vacuum drum filters have been assumed, designed to achieve 85% solids in the gypsum by-product. Vacuum pumps draw the liquid through the fabric on the drum, leaving the solids on the fabric. Filtrate from the drum filter is returned to the process.

When producing gypsum for disposal, the goal is to balance the incoming chlorine with the chlorides leaving in the gypsum. Thus, no rinsing is done at the vacuum filter. With 12,000 ppm chloride times 15% moisture, there is often sufficient chloride leaving in the cake that no chloride purge stream is required to achieve chloride balance. If chloride purge is required, the purge stream is much smaller than for the commercial gypsum case. The purge stream requires a similar, but much smaller, wastewater treatment system. For purposes of this study, we have assumed that sufficient chloride is lost with the by-product gypsum that no chloride purge is necessary.

Dewatered gypsum is discharged from the drum filters to belt conveyors that transfer the gypsum to a stackout pile. The pile contains up to three days storage and is designed to facilitate a large, articulated wheel loader loading the gypsum into 100-ton off-road dump trucks. The gypsum is assumed to be hauled to an on-site landfill.

Table 4.3–1 contrasts the major equipment used in the dry and wet limestone FGD systems.

TABLE 4.3-1
EQUIPMENT USED IN DRY & WET LIMESTONE FGD SUBSYSTEMS

	DRY SYSTEMS		**WET LSFO**	
	400 MW	**500 MW**	**400 MW**	**500 MW**
Flue Gas System				
Chimney • Retrofit unit • New unit	Re-use existing chimney Carbon steel liner		New chimney with FRP liner; may use existing as bypass FRP liner; no bypass	

ID fans (capacity) • Retrofit unit • New unit	Booster fans (2 x 50%) Size ID fans to include FGD ΔP		Booster fans (2 x 50%) Size ID fans to include FGD ΔP
Absorber inlet ductwork/dampers	Carbon steel		
Absorber outlet ductwork/dampers	Carbon steel		Lined carbon steel
Reagent Hdlg and Prep			
Reagent storage (capacity)	Steel silo for lime (14 days)		Pile for limestone (30 days)
Transport /reclaim system	Pneumatic conveyor		Belt conveyor
Storage bin (capacity)	Lime (24 hours)		Limestone (24 hours)
Ball mill/slaker (capacity)	LSD: Ball mill for quicklime; CDS: hydrator (both 2 x 100%)		Ball mill for limestone (2 x 100%)
Reagent storage (capacity)	LSD: Lime slurry tank (16 hours); CDS: Hydrated lime silo (24 h)		Limestone slurry tank (16 hours)
Lime slurry feed pump	LSD only: 2 x 100%		
Absorber/Reaction System			
Absorbers (capacity) • LSD, CDS • LSFO	1 x 100%	2 x 50%	1 x 100%, with spargers
Injection of solids/water: • LSD: atomizers • CDS: HP water nozzles • LSFO: slurry spray nozz.	3 x 50% 3 x 50%	2 x [3 x 25%] 2 x [3 x 25%]	Low S: 4 x 33% Medium S: 5 x 25%
Particulate Collection			
Pulse jet baghouse (air to cloth ratio)	LSD: 3.5 ft/min CDS: 3.2 ft/min + knockout chamb.		Retrofit unit: none required New unit: 4.0 ft/min.

By-Product Prep and Hdlg		
Dewatering	Not required	Hydroclone (20% spare), vacuum filters (2 x100%); proc. water tank
By-product buffer (capacity)	16 hours	
Handling/conveying system	LSD: pneumatic or other; CDS: air slide recirc; pneum. disp.	Belt conveyors (2 x 100%)
Bulk storage (capacity)	Steel silo (3 days)	Shed (7 days)
Water		
Makeup water tank	1 x 100%	1 x 100%
Makeup water pumps	2 x 100%	2 x 100%
Wastewater treatment plant	Not applicable	Suspended solids, heavy metals

4.4 GENERAL SUPPORT

The general support equipment not listed above includes all the typical balance-of-plant subsystems including the seal water system, instrument air system, makeup water system, distributed control system (DCS), electrical auxiliary power system, and accommodations for the FGD controls in the main control room. Additionally, a typical FGD facility will require plant site accommodations for roads, storm sewers, sanitary systems, and fire protection systems. These accommodations are included in the design and cost estimate.

5. KEY OPERATING COST ISSUES

Operating costs are dominated by the cost of the reagent and auxiliary power. In many cases, by-product management costs are also significant, and thus are reviewed here as well.

5.1 REAGENT COST

Reagent utilization is the efficiency with which the reagent fed to the FGD system reacts with the SO_2. Reagent utilization depends on many variables. It is generally better with wet processes than with dry, and better at low SO_2 removal efficiencies than high ones. Utilization also depends on the concentration of SO_2 entering the scrubber. Reagent utilization is usually expressed in terms of stoichiometric ratio (SR). Stoichiometric ratio is defined as moles of reagent per mole of SO_2. The reader is cautioned when comparing wet and dry processes that the wet FGD industry and the dry FGD industry evolved different norms for calculation of SR. Wet FGD stoichiometric ratio is calculated as moles of reagent per mole of SO_2 *removed*. Dry FGD stoichiometric ratio is calculated as moles of reagent per mole of SO_2 *at the inlet* to the FGD system. A comparison of SRs for the nine cases is shown in Table 5.1-1.

TABLE 5.1-1
REAGENT STOICHIOMETRIC RATIO
Moles of calcium per mole of sulfur *

Fuel	Low sulfur Appalachian	Medium sulfur Appalachian	Low sulfur Powder River Basin
Lime spray dryer @ 95%	1.40 (inlet basis)	1.60 (inlet basis)	1.10 new unit 1.20 retrofit
Circulating dry scrubber @ 98%	1.80 (inlet basis)	2.00 (inlet basis)	1.40 new unit 1.50 retrofit
LSFO (comm'l gypsum) @ 98%	1.03 (removed basis)	1.03 (removed basis)	1.03 (removed basis)

* Practice in the industry is to report wet technology stoichiometric ratios on the basis of moles calcium per mole of sulfur removed, whereas dry technology stoichiometric ratios are reported on the basis of moles calcium per mole of sulfur at the inlet to the FGD system.

The variation in lime utilization is affected in part by the concentration of the SO_2 in the flue gas, but for dry technologies it is also influenced by the alkalinity of the fly ash that is present in the absorber.

Of course, the unit cost of the reagents is also very important in the ultimate cost of reagent. Limestone is a basic mined material, with value added only through sizing. On the other hand, lime is basically limestone upgraded by application of energy. Lime has substantially higher cost, and its transportation cost may also be higher with respect to limestone in that there may be quarries in closer proximity to the plant than there are lime kilns. Table 5.1-2 shows the reagent costs used in this study.

TABLE 5.1-2
COST OF REAGENT

	Reagent	Delivered Cost
Lime spray dryer (LSD)	High-calcium lime	$75/ton
Circulating dry scrubber (CDS)	High-calcium lime	$75/ton
Wet Limestone/Forced Ox (LSFO)	Limestone	$15/ton

No sensitivity analysis has been performed here, due to the infinite variability of both the purchase cost and the transportation cost. Purchase cost can vary depending on available capacity in the quarry and kiln, energy costs, etc. Transportation cost varies with distance between the producer facility and the power plant. This distance varies widely. The process selection for any specific power plant should be made based upon a survey of the actual suppliers in the area. This activity is essential because the outcome is highly sensitive to cost of reagent.

5.2 AUXILIARY POWER

For LSD systems, the major energy consumption is that portion of ID fan power attributable to the pressure drop across the absorber, as well as the power for the atomizer or atomizing air compressor. About 60% to 70% of the energy required is due to an increase in draft (12 in. WC, including absorber, baghouse, and inlet and outlet ductwork) and 15% to 25% of the energy required is for the atomizers. For CDS systems, nearly 80% of the energy required is due to the increase in draft (16 in. WC including absorber, baghouse, inlet and outlet ductwork), with 20% of the energy required for the rest of the subsystems. For the LSFO system, energy consumption results mainly from the ID fan power attributable to the draft loss (8 in. WC, including absorber and inlet and outlet ductwork), the power requirement for the slurry recycle pumps, and the power to grind the stone. The power cost used in this analysis includes both the electrical energy cost and the capacity charge.

TABLE 5.2-1
AUXILIARY POWER COMPARISON
Typical, for illustrative purposes only. Based on 500 MW unit burning coal with 2.0 lb SO_2/MBtu

	LSD	CDS	Wet LSFO

Absorber ΔP (operating)	12 in. WC	16 in. WC	8.2 in. WC
ID Fan Incremental Power	3,240 kW	4,330 kW	2,220 kW
Recycle L/G, gal/1000 acf	--	--	80
Recycle Pump	--	--	2,690 kW
Other FGD Auxiliaries, Margin	2,260 kW	1,220 kW	2,090 kW
Total FGD Auxiliary Power	5,500 kW	5,500 kW	7,000 kW

TABLE 5.2-2
AUXILIARY POWER COMPARISON
Percent of gross plant output

Fuel	LSD	CDS	Wet LSFO
• Low Sulfur Appalachian	1.10	1.00	1.40
• Medium Sulfur Appalachian	1.15	1.05	1.70
• Low Sulfur PRB	1.20	1.10	1.30

5.3 BY-PRODUCT MANAGEMENT

Management of coal combustion by-products (CCBs) varies by region as well as locale. Potential by-products include:

- Fly ash
- FGD by-product

These may be kept separate, but dry FGD generally mixes them unless the installation is a retrofit where the existing particulate removal system will remain in place. Dry FGD by-product is a dry, granular material (much like fly ash). Wet limestone FGD gypsum is a clay-like material, either commercial-grade, made with high-calcium limestone and dewatered to less than 10% moisture, or disposal-grade, made with low-cost limestone and dewatered to 15% moisture.

Quantities of CCBs are very large. For instance, a 500 MW plant burning Powder River Basin coal for 40 years would produce sufficient fly ash to fill some 60 acres to a height of 80 feet (depending on the ash content of the specific coal). A spray dryer would produce an additional 4 to 10 acres (depending on the sulfur content of the coal). A wet limestone FGD system would produce 4 to 10 acres of gypsum.

There are several potential dispositions for these by-products:

Fly Ash Collected Separately

- Cement manufacturing
- Concrete filler/enhancer
- Landfill/mine reclamation

Dry FGD By-Product Collected Separately

- Soil conditioner
- Landfill/mine reclamation

Combined Dry FGD By-Product and Fly Ash

- Aggregate or block manufacture
- Landfill/mine reclamation

Commercial-Grade Gypsum

- Wallboard manufacturing
- Plaster or stucco manufacturing
- Cement manufacturing
- Soil conditioner
- Landfill/mine reclamation

Disposal-Grade Gypsum

- Soil conditioner
- Landfill/mine reclamation

While the list of uses looks long, two facts must be accepted: a) These materials are heavy, so they are expensive to ship any distance, and b) These are relatively low-value commodities. The combined result is that any use must be in close proximity to the power plant. For instance, the houses in which the wallboard will be installed must be within a few hundred miles. If there is no growing city within a few hundred miles of the plant, re-use of the CCBs is unlikely.

Where fly ash is collected separately, it has commonly been sold, for cement or concrete applications. At one time, the majority of fly ash was recycled. However, over the last 30 years, ash production has grown rapidly while ash use has been nearly constant. The result is that most of the fly ash collected separately is landfilled.

For dry FGD systems, the predominant mode has been disposal of combined wastes. This is partly because the combined wastes have fewer uses and command lower prices, and partly because many dry FGD systems are at the remote sites in the West, where transportation costs exceed the value of the by-product.

The predominant mode for wet limestone FGD over the last 20 years has been to produce gypsum for wallboard. This has been so successful that gypsum is no longer mined east of the Mississippi River. In the East, all wallboard is produced from synthetic gypsum made in FGD systems. However, this fact means that sales of gypsum can increase no faster than the wallboard market grows. With the wave of new scrubber installations much gypsum will have to be landfilled. West of the Mississippi, wet limestone FGD is rare due to the low-sulfur fuels.

Fortunately, these CCBs all serve as excellent landfill materials. They are structurally stable, having good properties for supporting light construction on the landfill. They are chemically stable, demonstrating very low leachability of their constituents to the ground water. There is no objectionable odor nor any biological impacts. In the case of mine-mouth power plants (which includes the majority of remote facilities), CCBs represent a valuable resource for the necessary reclamation of the mine to its original contours.

Disposal costs vary widely. They must include the cost of developing the fill site and haul road, as well as the cost of transporting and placing the material. A landfill site today is an engineered venue, including extensive grading and development of berms, development of an underdrain system and peripheral monitoring wells and installation of a heavy plastic liner. Preferably, the landfill can be reached without crossing any road, railroad or stream, so the material can be carried in off-highway trucks. The haul road must be graded and compacted, with soil remediation as necessary to support the heavy trucks. Depending on land cost, local construction costs, depreciation time and recovery rate, cost of the facility may translate to $10 to $50 per ton. Transportation and placement in an onsite landfill may cost $2 to $5 per ton. Where the by-product must be hauled on the highway, these costs escalate rapidly.

For the purposes of this study, cost of by-product handling was assumed as shown in Table 5.3-1. The base case is typical of practice today, where dry systems usually produce a by-product that must be landfilled at a cost of $20 per ton, but wet limestone FGD is often able to sell gypsum at a price that equals the cost to haul the gypsum to the wallboard manufacturing plant (net $0 per ton). Two sensitivities, reflecting the new realities of the marketplace, are explored as well. Case S1 recognizes the saturation of the gypsum market and assigns the full cost of disposal to all three technologies. Case S2 recognizes the drive to use CCBs in lower value applications. These are not as profitable as making wallboard from gypsum, but when that option is not available, use of CCBs in low-value applications is more attractive, both economically and environmentally, than disposal.

TABLE 5.3-1
COST OF BY-PRODUCT MANAGEMENT

	base case	sensitivity S1	sensitivity S2
Lime spray dryer (LSD)	Disposal $20/ton	Disposal $20/ton	Low-Value Use $3/ton
Circulating dry scrubber (CDS)	Disposal $20/ton	Disposal $20/ton	Low-Value Use $3/ton
Wet Limestone/Forced Ox (LSFO)	Commercial Use $0/ton	Disposal $20/ton	Low-Value Use $3/ton

6. RETROFIT VERSUS NEW UNITS

6.1.1 Dry FGD Systems

In the case of a new power plant, a baghouse must be installed regardless of whether it is part of the dry FGD system or freestanding, ahead of the wet limestone FGD system. The wet limestone FGD equipment is far more expensive than the dry scrubber, exclusive of the baghouse. However, in a retrofit situation, this large capital cost item is eliminated. The wet limestone FGD system can follow the existing particulate control, whereas the dry FGD requires a new particulate control system, in addition to, or in place of the existing one. Partially offsetting this advantage, the retrofit wet limestone FGD will require a new chimney, whereas the dry FGD typically does not.

6.1.2 Wet Limestone FGD Systems

Although wet limestone FGD systems are at a capital cost disadvantage in retrofit applications, they retain some advantages. The new chimney adds cost, but it substantially frees the layout from being constrained to return the flue gas to the existing chimney. The wet limestone FGD ductwork may be significantly shorter than the ductwork for the dry FGD. In the end, though, the deciding factor will be the sulfur content of the fuel and its impact on reagent cost and by-product management cost. As we will see in Section 7, the effect of the higher capital cost is simply to change the point at which wet limestone FGD becomes more attractive than dry FGD. With the assumptions used in this study, for a 500 MW plant, that point shifts from 2.2 lb SO_2/MBtu (new plant) to 2.0 lb SO_2/MBtu (retrofit facility). For a 400 MW plant, it remains at 3.0 lb SO_2/MBtu (new plant or retrofit).

6.1.3 Retrofit Issues

Regardless of whether wet or dry, retrofits entail significant extra costs beyond those of a new unit:

Process/Mechanical

- Booster fans required
- New chimney with wet liner required (LSFO)
- Inefficient ductwork routing
- Compromised layout of absorber area and reagent/by-product area
- Upgrades required to service water supply, compressed air supply, etc.

Civil/Structural

- Earthwork, pilings in close quarters vs. new unit done in efficient sequence
- Foundations work around chimney foundation, buried obstacles

Electrical/Controls

- Major upgrade of auxiliary power supply vs. new unit sized for total
- Typically sub-optimal power distribution
- DCS may not be able to handle incremental needs

7. COSTS ANALYSIS

7.1 CAPITAL COSTS

Much of the study work in the power industry is performed using 500 MW size as typical. However, any study that includes either the spray dryer FGD technology or the circulating dry scrubber technology can be misleading because these technologies are limited to a maximum absorber size of 400 MW, whereas the wet limestone FGD technology is not. This study includes cases for 400 MW units to show the very different comparison at that size. Also, costs are shown for a range of coals, to demonstrate how sulfur content affects the process selection. Sargent & Lundy has estimated capital cost for the full array of 36 cases studied here. Estimated capital costs were determined for new and retrofit applications, which includes the equipment, materials, structural, and electrical components associated with the retrofit installation of these technologies.

Costs were developed using Sargent & Lundy's database as well as price quotes obtained from manufacturers for the equipment/work needed. A significant change in the capital requirements for a conventional FGD system has occurred since the 2002 study. The major influence on the capital cost of the FGD system has stemmed from a unprecedented strain on virtually all the resources that are brought to bear on major construction projects, especially FGD systems. Resources including pumps, piping, alloys, fans, transformers, motor controls, steel, cable, mills, filters, engineering, etc., and most notably construction craft labor are being pushed to and beyond their historical capacity. All of these forces have increased the cost of equipment, material, and labor, and have significantly stretched out construction schedules.

FGD prices have seen a minimum of 25% inflation in the past year. Some recent contracts have been signed at prices over 300% higher than the market of 5 years ago. The costs for this study have been prepared on a consistent, uniform basis and show a level that some buyers achieved in mid-2006. Sargent & Lundy cautions the reader that the costs provided herein are not indicative of any cost one may actually achieve. However, we believe the costs are valid for comparative purposes. These costs should *not* be used for any of these purposes:

- Planning the cost of an FGD project
- Budget requests or allocations
- Solicitation of pollution control bonds

In today's marketplace, it is impossible to determine capital cost of an FGD system until the contract is signed with the supplier.

The capital cost estimates provided herein are "total plant cost," and include the following:

- Equipment and material
- Direct field labor
- Indirect field costs and engineering
- Contingency
- Owner's cost
- Allowance for funds during construction (AFUDC)
- Initial inventory and spare parts (1% of the process capital)
- Startup and commissioning

The capital cost estimates provided do not include sales taxes and property tax. License fees and royalties are not expected for the proposed control strategies.

Additionally, the underlying assumption is that the contracting arrangement for the project is large, multiple lump sum work packages. If the owner prefers to execute the project on an engineer, procure, construct (EPC), or turnkey, basis a separate risk allocation of 15% to 20% should be added to the estimate to accommodate the EPC/Turnkey contractor's mark-up, contingencies, and risk allocations.

Tables 7.1-1 and 7.1-2 present the capital costs for new units, 500 MW and 400 MW cases, respectively. Tables 7.1-3 and 7.1-4 cover the retrofit cases. These costs are based on "all else being equal," although in actual practice, "all else" is seldom equal. In particular, the current marketplace is a "seller's market." Thus, even when the seller's costs agree with the costs that these tables were based on, the seller's price may include a factor of 20% that reflects his diminished desire to capture the contract. This diminished desire may alternatively be expressed as a refusal to offer any price. Many suppliers are declining to bid on contracts they deem too small, too different from their experience base, too short a schedule, too difficult a labor environment, or too commercially risky. The capital cost tables reflect a mainstream project with competitive bids.

Tables 7.1-1 through 7.1-4 are all based on systems designed to produce by-products for disposal. A system designed to produce a commercial by-product will include some additional capital equipment. For instance, a LSFO system producing gypsum for use in wallboard manufacture must have vacuum belt filters instead of vacuum drum filters, at significantly higher cost. The belt filter must have a clean water supply for washing the gypsum. The filtrate will have to be purged at a much higher rate, requiring a wastewater treatment system. The by-product, which may be stacked out to a pile in the disposal scenario, must by stacked out to a covered shelter if the by-product is to be sold to a wallboard producer. Sargent & Lundy estimates that the capital cost for the commercial by-product cases is 8% higher than for the disposal-grade by-product cases.

TABLE 7.1-1
CAPITAL COST FOR FGD ON NEW 500 MW UNIT
$ Millions

	Low-Sulfur Powder River Basin (1.44 lb/MBtu)			**Low-Sulfur Appalachian** (2.0 lb/MBtu)			**Medium-Sulfur Appalachian** (3.0 lb/MBtu)		
	LSD	CDS	LSFO	LSD	CDS	LSFO	LSD	CDS	LSFO
Reagent Feed System (installed)	8.4	7.6	9.4	9.2	8.3	10.4	9.7	9.2	12.0
SO_2 Removal System (installed)	22.6	20.4	34.9	21.6	19.4	35.3	22.0	20.2	35.7
Baghouse System (installed)	20.7	26.8	20.2	19.7	25.6	19.7	19.7	25.6	19.7
Flue Gas System (installed)	14.6	14.6	16.0	14.0	14.0	15.7	14.0	14.0	15.7
Waste Hdlg/Rec. Sys. (installed)	5.4	4.3	5.7	5.6	4.5	6.2	6.7	5.6	6.7
General Supp't Eq't (installed)	2.7	2.7	4.7	2.7	2.7	4.5	2.7	2.7	4.7
Total Process	74.4	76.4	91.0	72.7	74.5	91.8	74.8	77.2	94.5
Project Costs*	39.2	42.2	47.8	38.2	39.1	48.1	39.3	40.6	49.8
Total Construction Cost	113.6	116.6	138.8	110.9	113.6	139.9	114.1	117.8	144.1

* Project costs include General Facilities, Engineering, Construction Management and Contingency, as well as AFUDC, Owner's Cost, Spares and Commissioning costs

TABLE 7.1-2
CAPITAL COST FOR FGD ON NEW 400 MW UNIT
$ Millions

	Low-Sulfur Powder River Basin (1.44 lb/MBtu)			**Low-Sulfur Appalachian** (2.0 lb/MBtu)			**Medium-Sulfur Appalachian** (3.0 lb/MBtu)		
	LSD	CDS	LSFO	LSD	CDS	LSFO	LSD	CDS	LSFO
Reagent Feed System (installed)	7.9	7.1	8.7	8.6	7.7	9.6	9.1	8.6	11.1
SO_2 Removal System (installed)	15.1	13.4	30.5	14.2	12.8	30.9	14.5	13.3	31.2
Baghouse System (installed)	17.7	23.0	16.9	16.8	21.9	16.4	16.8	21.9	16.4
Flue Gas System (installed)	10.9	10.9	14.0	10.5	10.5	13.7	10.5	10.5	13.7
Waste Hdlg/Rec. Sys. (installed)	4.8	3.9	5.2	5.0	4.0	5.6	6.0	5.0	6.0
General Supp't Eq't (installed)	2.7	2.7	4.7	2.7	2.7	4.5	2.7	2.7	4.7
Total Process	59.1	60.9	79.9	57.7	59.6	80.7	59.5	61.9	83.2
Project Costs*	31.0	32.0	41.3	30.4	31.3	42.4	31.3	32.5	43.7
Total Construction Cost	90.1	92.9	122.0	88.1	90.9	123.1	90.8	94.4	126.9

* Project costs include General Facilities, Engineering, Construction Management and Contingency, as well as AFUDC, Owner's Cost, Spares and Commissioning costs

TABLE 7.1-3
CAPITAL COST FOR FGD RETROFITTED TO EXISTING 500 MW UNIT
$ Millions

	Low-Sulfur Powder River Basin (1.44 lb/MBtu)			**Low-Sulfur Appalachian** (2.0 lb/MBtu)			**Medium-Sulfur Appalachian** (3.0 lb/MBtu)		
	LSD	CDS	LSFO	LSD	CDS	LSFO	LSD	CDS	LSFO
Reagent Feed System (installed)	9.4	8.4	10.2	10.2	9.2	11.3	10.8	10.2	13.1
SO_2 Removal System (installed)	25.2	22.7	41.0	24.0	21.6	41.5	24.5	22.5	42.0
Baghouse System (installed)	21.7	28.3	0.0	20.7	26.9	0.0	20.7	26.9	0.0
Flue Gas System (installed)	20.8	20.8	40.0	20.0	20.0	39.2	20.0	20.0	39.2
Waste Hdlg/Rec. Sys. (installed)	6.0	4.8	6.0	6.2	5.0	6.5	7.4	6.2	7.0
General Supp't Eq't (installed)	5.3	5.3	10.0	5.3	5.3	10.0	5.3	5.3	10.5
Total Process	88.4	90.2	107.2	86.4	88.0	108.5	88.6	91.0	111.8
Project Costs*	48.1	49.1	54.8	47.0	47.9	55.5	48.3	49.6	57.1
Total Construction Cost	136.5	139.3	162.0	133.4	135.9	164.0	136.9	140.6	168.9

* Project costs include General Facilities, Engineering, Construction Management and Contingency, as well as AFUDC, Owner's Cost, Spares and Commissioning costs

TABLE 7.1-4
CAPITAL COST FOR FGD RETROFITTED TO EXISTING 400 MW UNIT
$ Millions

	Low-Sulfur Powder River Basin (1.44 lb/MBtu)			**Low-Sulfur Appalachian** (2.0 lb/MBtu)			**Medium-Sulfur Appalachian** (3.0 lb/MBtu)		
	LSD	CDS	LSFO	LSD	CDS	LSFO	LSD	CDS	LSFO
Reagent Feed System (installed)	8.8	7.9	9.5	9.5	8.6	10.5	10.1	9.5	12.1
SO_2 Removal System (installed)	16.8	14.9	35.9	15.7	14.2	36.3	16.1	14.7	36.73
Baghouse System (installed)	18.6	24.2	0.0	17.7	23.0	0.0	17.7	23.0	0.0
Flue Gas System (installed)	15.6	15.6	35.0	15.0	15.0	34.3	15.0	15.0	34.3
Waste Hdlg/Rec. Sys. (installed)	5.4	4.3	5.4	5.5	4.5	5.9	6.6	5.5	6.3
General Supp't Eq't (installed)	4.6	4.6	8.7	4.6	4.6	8.7	4.6	4.6	9.2
Total Process	69.7	71.4	94.5	68.1	69.8	95.7	70.0	72.4	98.6
Project Costs*	37.9	38.8	48.2	37.1	38.1	48.9	38.2	39.5	50.5
Total Construction Cost	107.6	110.2	142.7	105.2	107.9	144.6	108.2	111.9	149.1

* Project costs include General Facilities, Engineering, Construction Management and Contingency, as well as AFUDC, Owner's Cost, Spares and Commissioning costs

7.2 OPERATION AND MAINTENANCE COSTS

Sargent & Lundy has estimated O&M expenses associated with the FGD systems in the various cases. These costs include both fixed and variable operating costs, defined in the following sections.

7.2.1 Fixed O&M Costs

The fixed O&M costs determined for this study consist of operating labor, maintenance labor, maintenance material, and administrative labor.

Staffing levels are dependent on many factors:

- Process selected and complexity of arrangement
- System configuration and plant location
- Work rules and plant organization

For the purposes of this study, S&L has estimated the staffing levels as shown in Table 7.2-1.

TABLE 7.2-1
FGD SYSTEM OPERATING LABOR

	Lime Spray Dryer	Circulating Dry Scrubber	Limestone Forced Oxidation
Operating Personnel			
• New Unit FGD Labor	5	4	8
• Retrofit FGD Labor	8	7	12

The operating labor requirement is similar for the 400 MW and the 500 MW facilities. Maintenance material and labor costs shown herein have been estimated based on technology operating experience in the United States and Europe. The maintenance cost includes periodic replacement of water atomizers and maintenance material for various subsystems, and the labor required to perform the maintenance.

7.2.2 Variable O&M Costs

Variable O&M costs determined for each application include the cost of consumables, including reagent (lime or limestone), by-product management, bag and cage replacement for the baghouse, water, and power requirements. These costs have been calculated at a unit capacity factor of 80%. This value is representative of the older units to which FGD retrofits are typically being made. Early in their operating life, new units will experience higher capacity factors (typically 85% or even higher), but lifetime capacity factor will be closer to the 80%.

Table 7.2-2 summarizes the parameters used for the FGD comparison.

TABLE 7.2-2
PARAMETERS USED FOR FGD COMPARISON

Fuel	**Low-Sulfur Powder River Basin** (1.44 lb/MBtu)	**Low-Sulfur Appalachian** (2.0 lb/MBtu)	**Medium-Sulfur Appalachian** (3.0 lb/MBtu)
General			
Load factor	80		
FGD system life, years	30 (new)/20 (retrofit)		
Capital cost leveling factor, %/year	14.8 (new)/15.7 (retrofit)		
Discount rate, %	9.0		
Inflation rate, %	3.0		
Operating cost levelization factor	1.365		
Lime Spray Dryer (LSD)			
SO_2 removal, %	95%		
SO_2 emission, lb/MBtu	0.072	0.10	0.15
Byproduct	Dry waste		
Power, % of gross output	1.20	1.10	1.15
Reagent	High calcium lime		
Reagent cost, $/ton	$75/ton		
Reagent purity, %	93		
Reagent stoich. ratio (inlet basis)	1.20	1.40	1.60
Circulating Dry Scrubber (CDS)			
SO_2 removal, %	97.2%	98%	98%
SO_2 emission, lb/MBtu	0.04	0.04	0.06
Byproduct	Dry waste		
Power, % of gross output	1.10	1.00	1.05
Reagent	High calcium lime		
Reagent cost, $/ton	$75/ton		
Reagent purity, %	93%		
Reagent stoich. ratio (inlet basis)	1.50	1.80	2.00
Limestone Forced Oxidation (LSFO)			
SO_2 removal, %	98%		
SO_2 emission, lb/MBtu	0.029	0.04	0.06

TABLE 7.2-2
PARAMETERS USED FOR FGD COMPARISON

Byproduct	Commercial-grade Gypsum		
Power, % of gross output	1.30	1.40	1.70
Reagent	Limestone		
Reagent cost, $/ton	$15/ton		
Reagent purity, %	95%		
Reagent stoich. ratio (removed basis)	1.03	1.03	1.03

The cost/revenue of fly ash management is not included in this study as it is assumed that even if the fly ash is currently disposed of or sold, the proposed configuration will not affect the current operation. For new unit operations, if the sale of fly ash could create a significant revenue stream, an ESP can be installed upstream of the dry FGD. For this analysis of new unit applications, it was assumed that the ash would be disposed of along with FGD by-product.

Considering both the regulated and unregulated market and new and retrofit applications, Sargent & Lundy believes that $45/MWh is representative of the future power market.

Tables 7.2-3 and 7.2-4 present the estimated operating and maintenance costs for new 500 MW and 400 MW units, respectively. Costs are broken out to show the differences by FGD process and by coal. Referring back to Table 5.3-1, these tables are for the base by-product management case, where LSD and CDS wastes are landfilled and LSFO by-product is sold. Tables 7.2-5 and 7.2-6 are for similar cases, except they are for sensitivity case S1, where all wastes are landfilled. Tables 7.2-7 and 7.2-8 are likewise similar, except they are for sensitivity case S2, where all the by-products are put to low-value use. It should also be noted that the costs in these tables are at 98% SO_2 removal. For the CDS and LSFO cases, reagent cost and by-product management cost are at 98% SO_2 removal. The LSD is not capable of 98% SO_2 removal, so the reagent cost and by-product management cost are at 95% SO_2 removal. To even up the cases, the LSD cases are assessed the cost to buy SO_2 emission allowances to cover the incremental emissions. The allowances are assumed to cost $600/ton.

TABLE 7.2-3 **O&M COST FOR FGD ON NEW 500 MW UNIT[1]** By-product base case (ref. Table 5.3-1)										
		Low-Sulfur Powder River Basin (1.44 lb/MBtu)			**Low-Sulfur Appalachian** (2.0 lb/MBtu)			**Medium-Sulfur Appalachian** (3.0 lb/MBtu)		
		LSD	CDS	LSFO	LSD	CDS	LSFO	LSD	CDS	LSFO
Fixed O&M Cost	**$M**/yr	2.646	2.446	3.672	2.603	2.402	3.719	2.654	2.465	3.810
Reagent Cost	**$M**/yr	2.031	2.585	0.646	3.462	4.451	0.873	5.934	7.418	1.309
By-Product Mgt. Cost	**$M**/yr	1.192	1.355	0.014	1.794	2.086	0.038	2.889	3.326	0.037
Bag & Cage Replacement	**$M**/yr	0.492	0.569	0.438	0.447	0.517	0.394	0.447	0.517	0.394
Water Cost	**$M**/yr	0.166	0.166	0.221	0.156	0.156	0.208	0.156	0.156	0.208
Auxiliary Power Cost	**$M**/yr	1.892	1.734	2.838	1.734	1.577	2.996	1.829	1.656	3.469
Add'l Allowance Cost	**$M**/yr	0.375	0.000	0.000	0.631	0.000	0.000	1.156	0.000	0.000
Variable O&M Cost	$M/yr	6.148	6.409	4.157	8.224	8.787	4.509	12.411	13.073	5.417
Total O&M Cost	**$M**/yr	8.794	8.855	7.829	10.827	11.189	8.228	15.065	15.538	9.227

1. Only the largest cost items are shown here; all others are too small to affect the process selection.

TABLE 7.2-4 **O&M COST FOR FGD ON NEW 400 MW UNIT[1]** By-product base case (ref. Table 5.3-1)										
		Low-Sulfur Powder River Basin (1.44 lb/MBtu)			**Low-Sulfur Appalachian** (2.0 lb/MBtu)			**Medium-Sulfur Appalachian** (3.0 lb/MBtu)		
		LSD	CDS	LSFO	LSD	CDS	LSFO	LSD	CDS	LSFO
Fixed O&M Cost	**$M**/yr	2.267	2.080	3.416	2.234	2.051	3.457	2.276	2.105	3.541
Reagent Cost	**$M**/yr	1.625	2.068	0.517	2.769	3.561	0.698	4.747	5.934	1.047
By-Product Mgt. Cost	**$M**/yr	0.954	1.084	0.012	1.435	1.668	0.031	2.311	2.661	0.029
Bag & Cage Replacement	**$M**/yr	0.394	0.460	0.350	0.358	0.418	0.315	0.358	0.418	0.315
Water Cost	**$M**/yr	0.132	0.132	0.177	0.125	0.125	.167	0.125	0.125	0.167
Auxiliary Power Cost	**$M**/yr	1.514	1.388	2.271	1.388	1.261	2.397	1.463	1.325	2.775
Add'l Allowance Cost	**$M**/yr	0.300	0.000	0.000	0.505	0.000	0.000	0.925	0.000	0.000
Variable O&M Cost	**$M**/yr	4.919	5.132	3.327	6.580	7.033	3.608	9.929	10.463	4.333
Total O&M Cost	**$M**/yr	7.186	7.212	6.743	8.814	9.084	7.065	12.205	12.568	7.874

1. Only the largest cost items are shown here; all others are too small to affect the process selection.

TABLE 7.2-5 **O&M COST FOR FGD ON NEW 500 MW UNIT[1]** By-product case S1 (ref. Table 5.3-1)										
		Low-Sulfur Powder River Basin (1.44 lb/MBtu)			**Low-Sulfur Appalachian** (2.0 lb/MBtu)			**Medium-Sulfur Appalachian** (3.0 lb/MBtu)		
		LSD	CDS	LSFO	LSD	CDS	LSFO	LSD	CDS	LSFO
Fixed O&M Cost	**$M**/yr	2.646	2.446	3.481	2.603	2.402	3.524	2.654	2.465	3.608
Reagent Cost	**$M**/yr	2.031	2.585	0.646	3.462	4.451	0.873	5.934	7.418	1.309
By-Product Mgt. Cost	**$M**/yr	1.192	1.355	1.435	1.794	2.086	1.938	2.889	3.326	2.907
Bag & Cage Replacement	**$M**/yr	0.492	0.569	0.438	0.447	0.517	0.394	0.447	0.517	0.394
Water Cost	**$M**/yr	0.166	0.166	0.221	0.156	0.156	0.208	0.156	0.156	0.208
Auxiliary Power Cost	**$M**/yr	1.892	1.892	2.838	1.734	1.734	2.996	1.829	1.829	3.469
Add'l Allowance Cost	**$M**/yr	0.375	0.000	0.000	0.631	0.000	0.000	1.156	0.000	0.000
Variable O&M Cost	$M/yr	6.148	6.567	5.578	8.224	8.944	6.409	12.411	13.246	8.287
Total O&M Cost	**$M**/yr	8.794	9.013	9.059	10.827	11.346	9.933	15.065	15.711	11.895

1. Only the largest cost items are shown here; all others are too small to affect the process selection.

TABLE 7.2-6 **O&M COST FOR FGD ON NEW 400 MW UNIT[1]** By-product case S1 (ref. Table 5.3-1)										
		Low-Sulfur Powder River Basin (1.44 lb/MBtu)			**Low-Sulfur Appalachian** (2.0 lb/MBtu)			**Medium-Sulfur Appalachian** (3.0 lb/MBtu)		
		LSD	CDS	LSFO	LSD	CDS	LSFO	LSD	CDS	LSFO
Fixed O&M Cost	**$M**/yr	2.267	2.080	3.246	2.234	2.051	3.285	2.276	2.105	3.362
Reagent Cost	**$M**/yr	1.625	2.068	0.517	2.769	3.561	0.698	4.747	5.934	1.047
By-Product Mgt. Cost	**$M**/yr	0.954	1.084	1.148	1.435	1.668	1.550	2.311	2.661	2.326
Bag & Cage Replacement	**$M**/yr	0.394	0.460	0.350	0.358	0.418	0.315	0.358	0.418	0.315
Water Cost	**$M**/yr	0.132	0.132	0.177	0.125	0.125	.167	0.125	0.125	0.167
Auxiliary Power Cost	**$M**/yr	1.514	1.514	2.271	1.388	1.388	2.397	1.463	1.463	2.775
Add'l Allowance Cost	**$M**/yr	0.300	0.000	0.000	0.505	0.000	0.000	0.925	0.000	0.000
Variable O&M Cost	**$M**/yr	4.919	5.258	4.463	6.580	7.160	5.127	9.929	10.589	6.630
Total O&M Cost	**$M**/yr	7.186	7.338	7.709	8.814	9.211	8.412	12.205	12.694	9.992

1. Only the largest cost items are shown here; all others are too small to affect the process selection.

TABLE 7.2-7 **O&M COST FOR FGD ON NEW 500 MW UNIT**[1] By-product case S2 (ref. Table 5.3-1)										
		Low-Sulfur Powder River Basin (1.44 lb/MBtu)			**Low-Sulfur Appalachian** (2.0 lb/MBtu)			**Medium-Sulfur Appalachian** (3.0 lb/MBtu)		
		LSD	CDS	LSFO	LSD	CDS	LSFO	LSD	CDS	LSFO
Fixed O&M Cost	**$M**/yr	2.646	2.446	3.481	2.603	2.402	3.524	2.654	2.465	3.608
Reagent Cost	**$M**/yr	2.031	2.585	0.646	3.462	4.451	0.873	5.934	7.418	1.309
By-Product Mgt. Cost	**$M**/yr	0.179	0.203	0.215	0.269	0.313	0.291	0.433	0.499	0.436
Bag & Cage Replacement	**$M**/yr	0.492	0.569	0.438	0.447	0.517	0.394	0.447	0.517	0.394
Water Cost	**$M**/yr	0.166	0.166	0.221	0.156	0.156	0.208	0.156	0.156	0.208
Auxiliary Power Cost	**$M**/yr	1.892	1.892	2.838	1.734	1.734	2.996	1.829	1.829	3.469
Add'l Allowance Cost	**$M**/yr	0.375	0.000	0.000	0.631	0.000	0.000	1.156	0.000	0.000
Variable O&M Cost	**$M**/yr	5.135	5.415	4.538	6.669	7.171	4.762	9.955	10.403	5.816
Total O&M Cost	**$M**/yr	7.781	7.861	7.839	9.302	9.573	8.286	12.609	12.868	9.424

1. Only the largest cost items are shown here; all others are too small to affect the process selection.

TABLE 7.2-8 **O&M COST FOR FGD ON NEW 400 MW UNIT**[1] By-product case S2 (ref. Table 5.3-1)										
		Low-Sulfur Powder River Basin (1.44 lb/MBtu)			**Low-Sulfur Appalachian** (2.0 lb/MBtu)			**Medium-Sulfur Appalachian** (3.0 lb/MBtu)		
		LSD	CDS	LSFO	LSD	CDS	LSFO	LSD	CDS	LSFO
Fixed O&M Cost	**$M**/yr	2.267	2.080	3.246	2.234	2.051	3.285	2.276	2.105	3.362
Reagent Cost	**$M**/yr	1.625	2.068	0.517	2.769	3.561	0.698	4.747	5.934	1.047
By-Product Mgt. Cost	**$M**/yr	0.143	0.163	0.172	0.215	0.250	0.233	0.347	0.399	0.349
Bag & Cage Replacement	**$M**/yr	0.394	0.460	0.350	0.358	0.418	0.315	0.358	0.418	0.315
Water Cost	**$M**/yr	0.132	0.132	0.177	0.125	0.125	.167	0.125	0.125	0.167
Auxiliary Power Cost	**$M**/yr	1.514	1.514	2.271	1.388	1.388	2.397	1.463	1.463	2.775
Add'l Allowance Cost	**$M**/yr	0.300	0.000	0.000	0.505	0.000	0.000	0.925	0.000	0.000
Variable O&M Cost	**$M**/yr	4.108	4.337	3.487	5.360	5.742	3.810	7.965	8.327	4.653
Total O&M Cost	**$M**/yr	6.375	6.417	6.733	7.594	7.793	7.095	10.241	10.432	8.015

1. Only the largest cost items are shown here; all others are too small to affect the process selection.

7.3 LEVELIZED COSTS

Levelized costs, also referred to as "life cycle costs," take into account the impacts of capital costs and O&M costs during the operation of a plant over the period of analysis. The levelized fixed charge rate (capital cost recovery) was calculated based on the experience of a typical regulated utility. The levelized fixed charge rate includes depreciation of the property, return on capital (50% debt and 50% equity), income tax, property tax, and insurance. Based on 9.0% discount rate and 30-year or 20-year life expectancy for new or retrofit facilities, respectively, the levelized fixed charge rates are 14.8% (30-year life) and 15.70% (20-year life). The levelized cost analysis was performed based on current 2006 dollars.

The levelized O&M cost factors are also based on the experience of a typical regulated utility and take into account the discount rate, escalation rate, and annuity rate. The levelized O&M cost factors are 1.365 for the 30-year period and 1.274 for the 20-year analysis. Tables 7.3-1, 7.3-2 and 7.3-3 show the buildup of levelized costs for new applications, for LSD, CDS and LSFO units, respectively.

TABLE 7.3-1
LEVELIZED COSTS FOR NEW LIME SPRAY DRYER (LSD) UNITS

	Low Sulfur PRB (1.44 lb/MBtu)		**Low Sulfur Appalachian** (2.0 lb/MBtu)		**Medium Sulfur Appalachian** (3.0 lb/MBtu)	
	400 MW	500 MW	400 MW	500 MW	400 MW	500 MW
Levelized Capital Cost, M$/an.	13.34	16.81	13.04	16.42	13.43	16.88
Levelized O&M Cost, M$/an.	9.81	12.00	12.03	14.78	16.66	20.56
Total Levelized Cost, M$/an.	23.15	28.81	25.07	31.20	30.09	37.45
Total ¢/kWh	0.83	0.82	0.89	0.89	1.07	1.07
Efficiency, $/T SO_2 Removed	1,140	1,135	915	910	730	730

Notes: 1. Capital Costs per Table 7.1-1 and 7.1-2
2. O&M Costs per Tables 7.2-3 and 7.2-4 (base by-product management case)
3. Note that this technology is capable of no more than 95% SO_2 removal. In most parts of the US, this means that a permit could not be obtained for the 3.0 lb/MBtu coal. In some locations, it may not be possible for the other coals as well.
4. For comparison purposes a cost of $600/ton has been assumed for the allowances necessary to compare with the 98% SO_2 removal of the other two FGD processes.

TABLE 7.3-2
LEVELIZED COSTS FOR NEW CIRCULATING DRY SCRUBBER (CDS) UNITS

	Low Sulfur PRB (1.44 lb/MBtu)		Low Sulfur Appalachian (2.0 lb/MBtu)		Medium Sulfur Appalachian (3.0 lb/MBtu)	
	400 MW	500 MW	400 MW	500 MW	400 MW	500 MW
Levelized Capital Cost, M$/an.	13.75	17.25	13.45	16.82	13.97	17.44
Levelized O&M Cost, M$/an.	10.02	12.30	12.57	15.49	17.33	21.45
Total Levelized Cost, M$/an.	23.77	29.55	26.02	32.31	31.30	38.89
Total ¢/kWh	0.85	0.84	0.93	0.92	1.12	1.11
Efficiency, $/T SO_2 Removed	1,171	1,160	950	940	760	755

Notes: 1. Capital Costs per Table 7.1-1 and 7.1-2
2. O&M Costs per Tables 7.2-3 and 7.2-4 (base by-product management case)

TABLE 7.3-3
LEVELIZED COSTS FOR NEW WET LIMESTONE/FORCED OXIDATION (LSFO) UNITS

	Low Sulfur PRB (1.44 lb/MBtu)		Low Sulfur Appalachian (2.0 lb/MBtu)		Medium Sulfur Appalachian (3.0 lb/MBtu)	
	400 MW	500 MW	400 MW	500 MW	400 MW	500 MW
Levelized Capital Cost, M$/an.	19.15	21.79	18.22	20.71	19.92	22.62
Levelized O&M Cost, M$/an.	8.59	9.97	9.64	11.23	10.03	11.76
Total Levelized Cost, M$/an.	27.74	31.77	27.86	31.94	29.95	34.37
Total ¢/kWh	0.99	0.91	0.99	0.91	1.07	0.98
Efficiency, $/T SO_2 Removed	1,560	1,250	1,160	930	835	670

Notes: 1. Capital Costs per Table 7.1-1 and 7.1-2
2. O&M Costs per Tables 7.2-3 and 7.2-4 (base by-product management case)

Similarly, Tables 7.3-4, 7.3-5 and 7.3-6 show the buildup of levelized costs for retrofit applications, for the same three technologies.

TABLE 7.3-4
LEVELIZED COSTS FOR RETROFIT LIME SPRAY DRYER (LSD) UNITS

	Low Sulfur PRB (1.44 lb/MBtu)		Low Sulfur Appalachian (2.0 lb/MBtu)		Medium Sulfur Appalachian (3.0 lb/MBtu)	
	400 MW	500 MW	400 MW	500 MW	400 MW	500 MW
Levelized Capital Cost, M$/an.	16.89	21.43	16.51	20.94	16.99	21.49
Levelized O&M Cost, M$/an.	10.01	12.12	11.85	14.41	16.17	19.81
Total Levelized Cost, M$/an.	26.91	33.55	28.36	35.35	33.16	41.30
Total ¢/kWh	0.96	0.96	1.01	1.01	1.18	1.18
Efficiency, $/T SO_2 Removed	1,325	1,320	1,035	1,030	805	805

Notes: 1. Capital Costs per Table 7.1-3 and 7.1-4
2. O&M Costs per Tables 7.2-3 and 7.2-4 (base by-product management case), adjusted for operating labor according to Table 7.2-1

TABLE 7.3-5
LEVELIZED COSTS FOR RETROFIT CIRCULATING DRY SCRUBBER (CDS) UNITS

	Low Sulfur PRB (1.44 lb/MBtu)		Low Sulfur Appalachian (2.0 lb/MBtu)		Medium Sulfur Appalachian (3.0 lb/MBtu)	
	400 MW	500 MW	400 MW	500 MW	400 MW	500 MW
Levelized Capital Cost, M$/an.	17.31	21.87	16.94	21.33	17.56	22.07
Levelized O&M Cost, M$/an.	10.21	12.40	12.35	15.08	16.79	20.62
Total Levelized Cost, M$/an.	27.52	34.27	29.29	36.41	34.35	42.69
Total ¢/kWh	0.98	0.98	1.04	1.04	1.23	1.22
Efficiency, $/T SO_2 Removed	1,352	1,345	1,065	1,060	835	830

Notes: 1. Capital Costs per Table 7.1-3 and 7.1-4
2. O&M Costs per Tables 7.2-3 and 7.2-4 (base by-product management case), adjusted for operating labor according to Table 7.2-1

TABLE 7.3-6 LEVELIZED COSTS FOR RETROFIT WET LIMESTONE/FORCED OXIDATION (LSFO) UNITS						
	Low Sulfur PRB (1.44 lb/MBtu)		**Low Sulfur Appalachian** (2.0 lb/MBtu)		**Medium Sulfur Appalachian** (3.0 lb/MBtu)	
	400 MW	500 MW	400 MW	500 MW	400 MW	500 MW
Levelized Capital Cost, M$/an.	22.41	25.43	22.70	25.74	23.40	26.51
Levelized O&M Cost, M$/an.	8.39	8.92	8.62	9.80	9.65	11.08
Total Levelized Cost, M$/an.	30.80	34.35	31.33	35.55	33.06	37.59
Total ¢/kWh	1.10	0.98	1.12	1.01	1.18	1.07
Efficiency, $/T SO_2 Removed	1,515	1,230	1,140	1,035	800	730

Notes: 1. Capital Costs per Table 7.1-3 and 7.1-4
2. O&M Costs per Tables 7.2-3 and 7.2-4 (base by-product management case), adjusted for elimination of baghouse maintenance expenses

Tables 7.3-7 through 7.3-12 provide summaries of the levelized cost, in terms of impact on the cost of electricity from the generating unit, for new and retrofit projects, respectively.

TABLE 7.3-7 SUMMARY OF LEVELIZED COSTS FOR NEW UNITS base by-product management case (ref. Table 5.3-1) (¢/kWh)						
	Low Sulfur PRB (1.44 lb/MBtu)		**Low Sulfur Appalachian** (2.0 lb/MBtu)		**Medium Sulfur Appalachian** (3.0 lb/MBtu)	
	400 MW	500 MW	400 MW	500 MW	400 MW	500 MW
LSD	0.83	0.82	0.89	0.89	1.07	1.07
CDS	0.85	0.84	0.93	0.92	1.12	1.11
LSFO	0.99	0.91	0.99	0.91	1.07	0.98

TABLE 7.3-8
SUMMARY OF LEVELIZED COSTS FOR NEW UNITS
by-product case S1 (ref. Table 5.3-1)
(¢/kWh)

	Low Sulfur PRB (1.44 lb/MBtu)	**Low Sulfur Appalachian** (2.0 lb/MBtu)	**Medium Sulfur Appalachian** (3.0 lb/MBtu)
	400 MW	400 MW	400 MW
LSD	0.83	0.89	1.07
CDS	0.85	0.93	1.11
LSFO	0.98	1.01	1.11

TABLE 7.3-9
SUMMARY OF LEVELIZED COSTS FOR NEW UNITS
by-product case S2 (ref. Table 5.3-1)
(¢/kWh)

	Low Sulfur PRB (1.44 lb/MBtu)	**Low Sulfur Appalachian** (2.0 lb/MBtu)	**Medium Sulfur Appalachian** (3.0 lb/MBtu)
	400 MW	400 MW	400 MW
LSD	0.79	0.83	0.98
CDS	0.80	0.86	1.01
LSFO	0.93	0.94	1.02

TABLE 7.3-10
SUMMARY OF LEVELIZED COSTS FOR RETROFIT UNITS
base by-product management case (ref. Table 5.3-1)
(¢/kWh)

	Low Sulfur PRB (1.44 lb/MBtu)		**Low Sulfur Appalachian** (2.0 lb/MBtu)		**Medium Sulfur Appalachian** (3.0 lb/MBtu)	
	400 MW	500 MW	400 MW	500 MW	400 MW	500 MW
LSD	0.96	0.96	1.01	1.01	1.18	1.18
CDS	0.98	0.98	1.04	1.04	1.23	1.22
LSFO	1.10	0.98	1.12	1.01	1.18	1.07

TABLE 7.3-11
SUMMARY OF LEVELIZED COSTS FOR RETROFIT UNITS
by-product case S1 (ref. Table 5.3-1)
(¢/kWh)

	Low Sulfur PRB (1.44 lb/MBtu)	**Low Sulfur Appalachian** (2.0 lb/MBtu)	**Medium Sulfur Appalachian** (3.0 lb/MBtu)
	400 MW	400 MW	400 MW
LSD	0.96	1.01	1.18
CDS	0.98	1.04	1.23
LSFO	1.10	1.11	1.21

TABLE 7.3-12
SUMMARY OF LEVELIZED COSTS FOR RETROFIT UNITS
by-product case S2 (ref. Table 5.3-1)
(¢/kWh)

	Low Sulfur PRB (1.44 lb/MBtu)	**Low Sulfur Appalachian** (2.0 lb/MBtu)	**Medium Sulfur Appalachian** (3.0 lb/MBtu)
	400 MW	400 MW	400 MW
LSD	0.92	0.96	1.09
CDS	0.94	0.98	1.12
LSFO	1.04	1.05	1.12

To help visualize the sensitivity of the process selection to the sulfur content of the fuel, a series of curves has been prepared. Figure 7.3-1 shows how the capital cost of the FGD system varies with the SO_2 loading of the flue gas it is designed to treat. This figure is for new units and assumes a commercial by-product in the LSFO case, but disposal of the by-products from the dry FGD systems, as has been typical recent practice. It should be noted for all these curves that the lime spray dryer (LSD) cases have been drawn all the way to 3.0 lb SO_2/MBtu, but for new US units, 95% SO_2 removal is insufficient to achieve permit limits for coals producing more than 2.0 lb SO_2/MBtu. It is clear from this curve, that capital cost of a LSFO FGD system is consistently higher than for the dry technologies. It can also be seen that the difference in capital cost between the LSD and CDS technologies is small, and within the band of uncertainty. If the technology decision were made strictly on capital cost, wet limestone FGD technology would rarely be selected.

The FGD system's impact on cost of electricity includes an O&M element as well as a capital cost recovery element. Figures 7.3-2, -3, and -4 show how impact on cost of electricity varies with the SO_2 loading the FGD system is treating. Figure 7.3-2 shows the relationship for the typical situation where the LSFO system produces commercial by-product and the dry cases produce disposable by-product. The break-even points for the 500 MW size (around 2.0 lb SO_2/MBtu when comparing to CDS, slightly higher for LSD) and for the 400 MW size (around 2.5 lb SO_2/MBtu, when comparing to CDS; 3.0 for LSD) are circled. It is clear that where the fuel is high in sulfur, the savings in reagent with LSFO more than offsets the higher capital cost. Figure

7.3-3 shows the same relationship, but for the situation where all the technologies produce a disposable by-product. In this scenario, the break-even point for the 400 MW size has moved to about 2.9 lb SO_2/MBtu for the CDS; The LSD break-even point is off the chart. Figure 7.3-4 examines the situation where all the technologies produce commercial by-product. In this scenario, LSFO is not favored over the dry technologies at any sulfur level.

Figures 7.3-5 through 7.3-8 show the same relationships, but for FGD retrofit projects. The break-even points for these situations are at slightly lower sulfur levels than for the new unit situations.

For the assumptions made in this study, some generalizations can be made. For the 1.44 lb SO_2/MBtu (0.6% sulfur) Powder River Basin coal, either 400 MW size or 500 MW size, regardless of the disposition of the by-product, the FGD technology ranking is:

1. (tie) Lime Spray Dryer
1. (tie) Circulating Dry Scrubber
3. Limestone Forced Oxidation

For the 2.0 lb SO_2/MBtu (1.3% sulfur) Appalachian coal, at the 400 MW size, regardless of the disposition of the by-product, the FGD technology ranking remains:

1. (tie) Lime Spray Dryer
1. (tie) Circulating Dry Scrubber
3. Limestone Forced Oxidation

However, at the 500 MW size, the selection is a three-way tie.

For the 3.0 lb SO_2/MBtu (2.0% sulfur) Appalachian coal, at the 400 MW size, regardless of the disposition of the by-product, the FGD technology ranking is very close, with a slight advantage to the Lime Spray Dryer. The strength of the LSFO technology lies with fuels having higher sulfur than those investigated here.

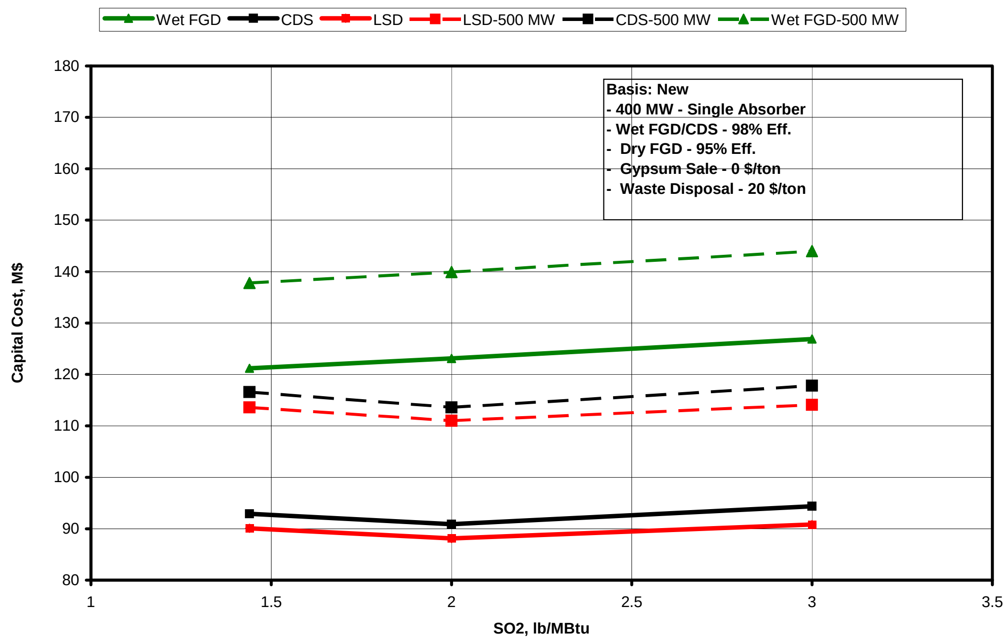

FIGURE 7.3-1: COMPARISON OF FGD TECHNOLOGY FOR NEW UNIT:

CAPITAL COST VS. SO_2 LOADING

base by-product management case

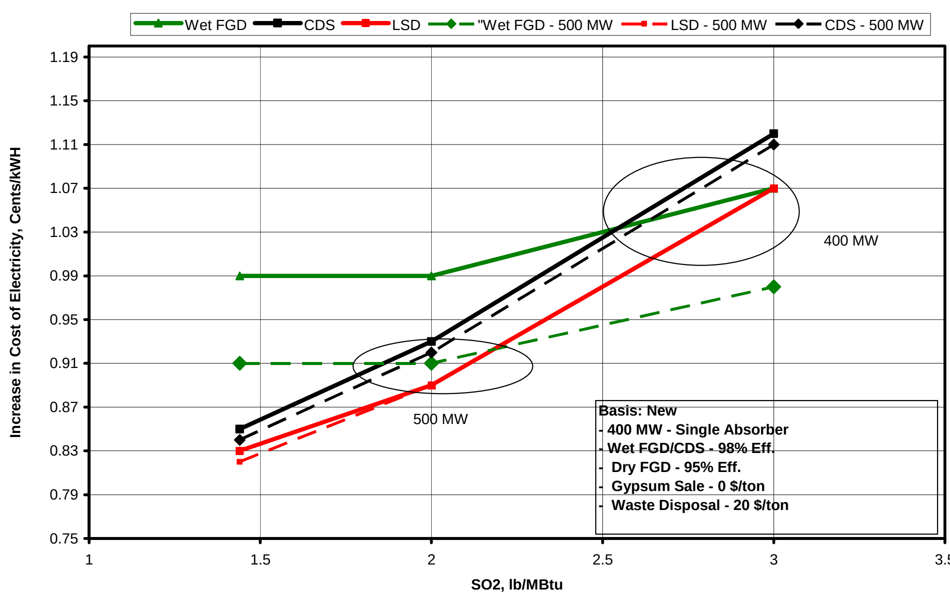

FIGURE 7.3-2: COMPARISON OF FGD TECHNOLOGY FOR NEW UNIT:

Δ COST OF ELECTRICITY VS. SO_2 LOADING

base by-product management case

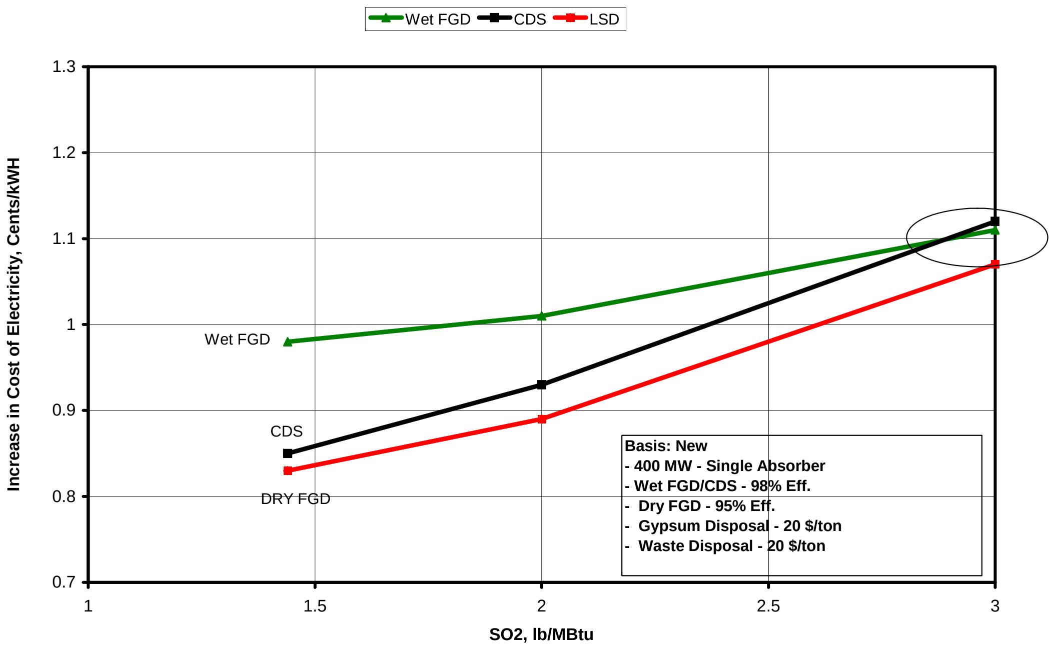

FIGURE 7.3-3: COMPARISON OF FGD TECHNOLOGY FOR NEW UNIT:

Δ COST OF ELECTRICITY VS. SO_2 LOADING

by-product sensitivity S1

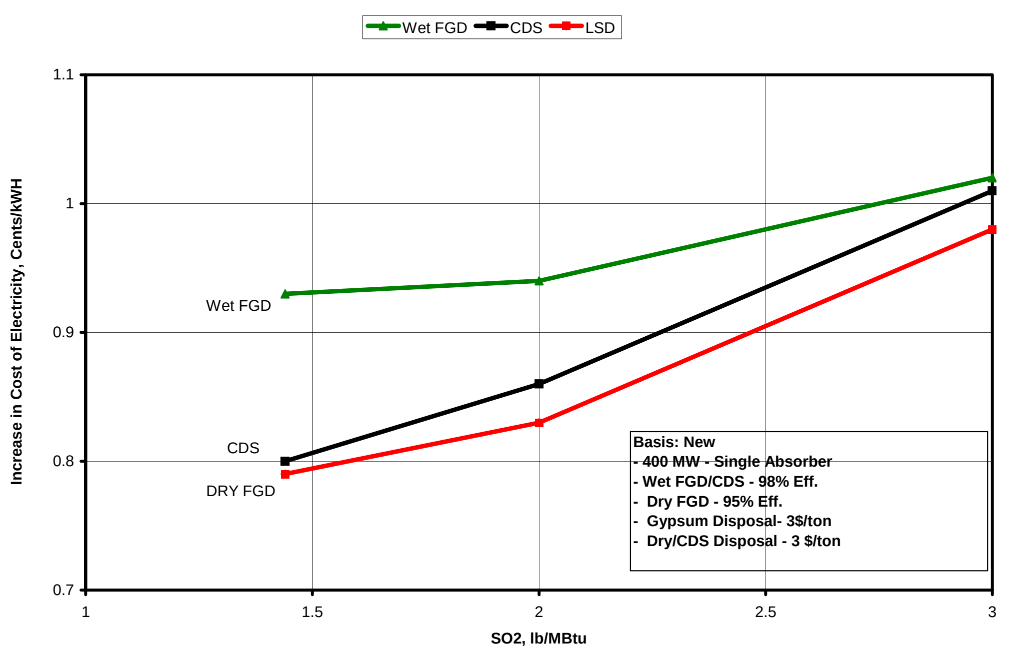

FIGURE 7.3-4: COMPARISON OF FGD TECHNOLOGY FOR NEW UNIT:

Δ COST OF ELECTRICITY VS. SO_2 LOADING

by-product sensitivity S2

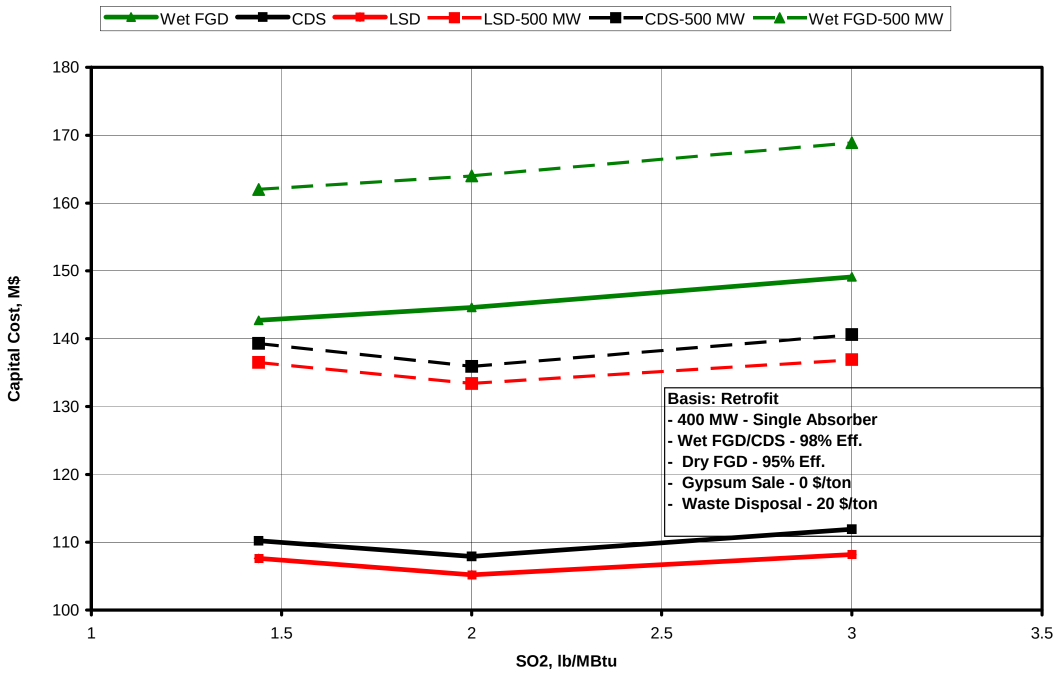

FIGURE 7.3-5: COMPARISON OF FGD TECHNOLOGY FOR RETROFIT UNIT:

CAPITAL COST VS. SO_2 LOADING

base by-product management case

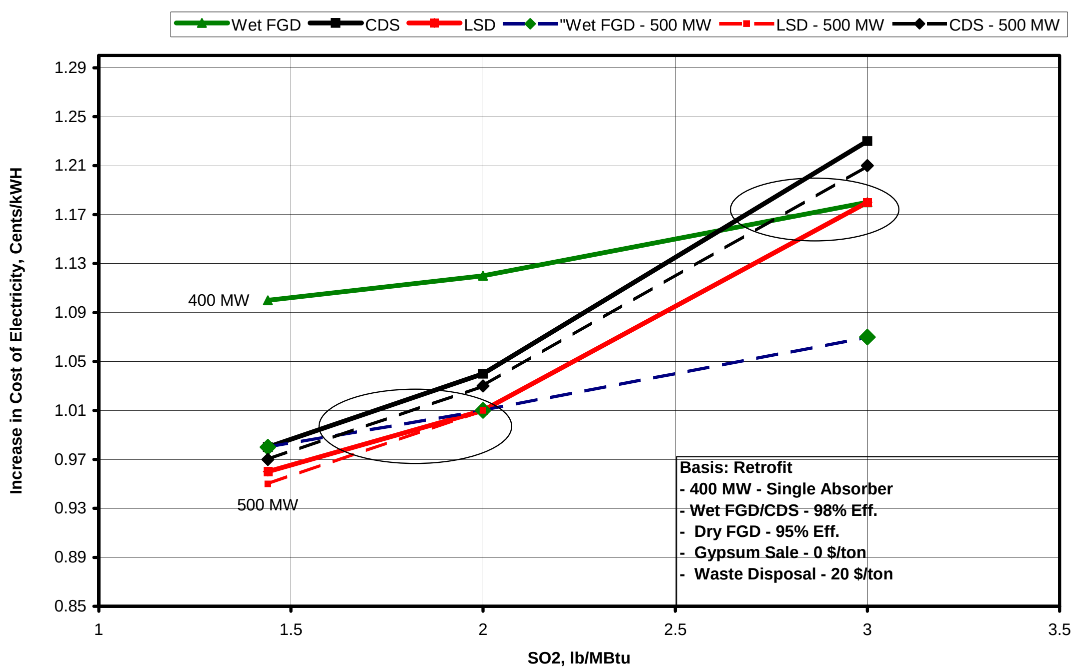

FIGURE 7.3-6: COMPARISON OF FGD TECHNOLOGY FOR RETROFIT UNIT:

Δ COST OF ELECTRICITY VS. SO_2 LOADING

base by-product management case

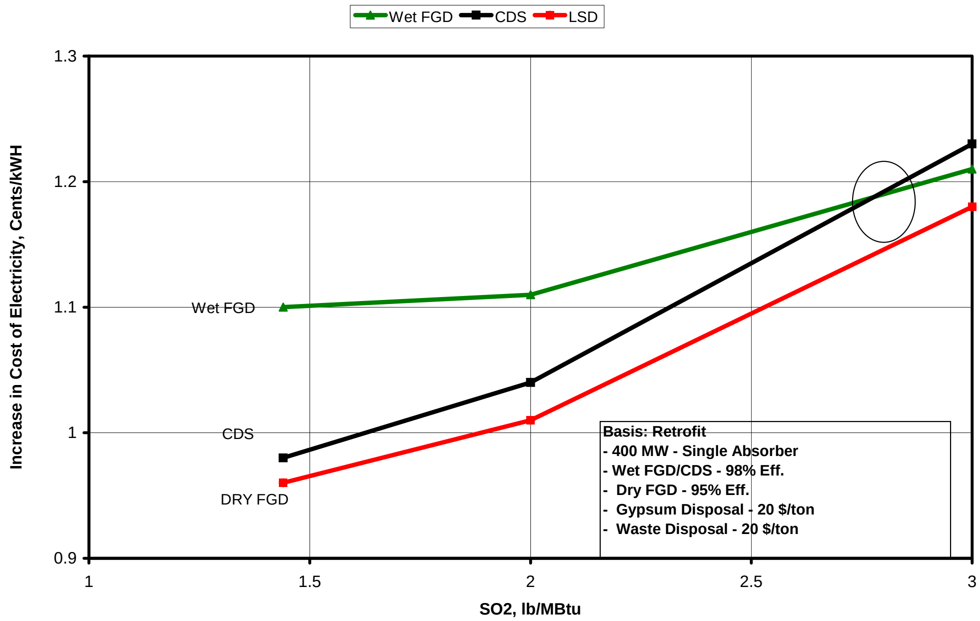

FIGURE 7.3-7: COMPARISON OF FGD TECHNOLOGY FOR RETROFIT UNIT:

Δ COST OF ELECTRICITY VS. SO_2 LOADING

by-product sensitivity S1

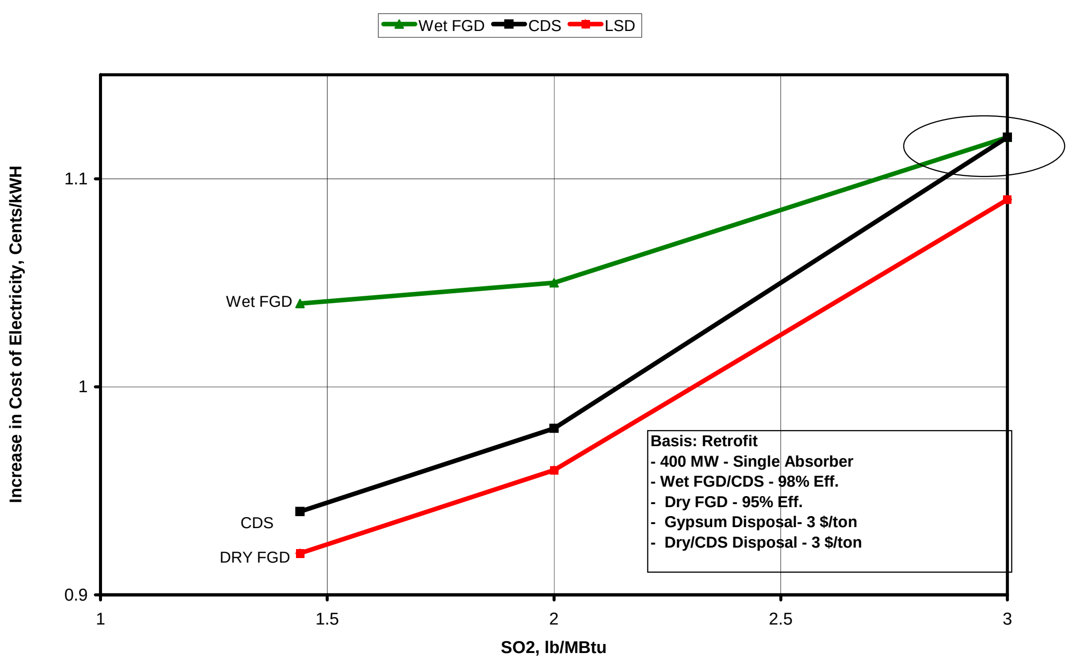

FIGURE 7.3-8: COMPARISON OF FGD TECHNOLOGY FOR RETROFIT UNIT:

Δ COST OF ELECTRICITY VS. SO_2 LOADING

by-product sensitivity S2