WET FLUE GAS DESULFURIZATION TECHNOLOGY EVALUATION

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CONTENTS

SECTION PAGE

CONTENTS

SECTION PAGE

EXHIBITS

APPENDIX

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1. FLUE GAS DESULFURIZATION (FGD) DESCRIPTION

Wet FGD technology, which is based on using limestone or lime as a reagent, is a wet scrubbing process and has been the FGD technology most frequently selected for sulfur dioxide (SO_2) reduction from coal-fired utility boilers. The wet FGD flue gas treatment system is typically located after removal of particulate matter from flue gas either by a baghouse or by an electrostatic precipitator. The cleaned gas is discharged to the stack. This type of FGD system removes SO_2 by scrubbing the flue gas with either a limestone or lime (reagent) slurry. The wet FGD process is considered a commercially mature technology and is offered by a number of suppliers. This report presents the results of Sargent & Lundy's evaluation of both limestone-based (limestone with forced oxidation [LSFO]) and lime-based (magnesium-enhanced lime [MEL] with forced oxidation) FGD technologies.

Flue gas is treated in an absorber by passing the flue gas stream through a limestone or lime slurry spray. In typical absorber designs, the gas flows upward through the absorber countercurrent to the spray liquor flowing downward through the absorber. However, other designs are also available, including co-current and countercurrent designs, and where the gas is forced through the liquor in a froth-type bubbling absorber. In a typical design, slurry is pumped through banks of spray nozzles to atomize it to fine droplets and uniformly contact the gas. The droplets absorb SO_2 from the gas, facilitating the reaction of the SO_2 with reagent in the slurry. Hydrogen chloride present in the flue gas is also absorbed and neutralized with reagent, causing an accumulation of chloride ions in the process liquid. Some of the water in the spray droplets evaporates, cooling the gas at the inlet from approximately 300°F to 125°F-130°F, and saturating the flue gas with water. The desulfurized flue gas passes through mist eliminators to remove entrained droplets before the flue gas is sent to the stack.

In most wet FGD systems, SO_2 collection efficiency is controlled by selecting appropriate design features for the system. For example, the quantity of liquid sprayed relative to flue gas is related to the SO_2 collection efficiency needed and is referred to as liquid-to-gas (L/G) ratio. Higher L/G ratios improve SO_2 removal by exposing the gas to more absorbing liquor. However, higher L/G ratios also consume more power, and this design feature must be factored against other important design features, including type of reagent. An important goal of this study will be to quantify the impact that selection of reagent (limestone or lime) will have on the design of the system and the resultant capital and operations & maintenance (O&M) costs.

After contacting the gas, the slurry collects in the bottom of the absorber in a reaction tank. The slurry is agitated to prevent settling. Limestone or lime consumed in the process is replenished by adding fresh limestone or lime slurry to the reaction tank.

In the LSFO process, the slurry is also aerated in the reaction tank to oxidize calcium sulfite hemihydrate ($CaSO_3 \bullet ½ H_2O$) to calcium sulfate dihydrate ($CaSO_4 \bullet 2H_2O$), or gypsum, which precipitates. This is where the term "forced oxidation" originates and it distinguishes this process from older, more troublesome limestone-based "natural oxidation" technology. The oxidized slurry is then recirculated to the spray headers. A portion of the slurry is withdrawn to remove the precipitated gypsum. Typically, this 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 gypsum slurry is returned to the process. A portion of this water is removed from the system as wastewater to limit accumulation of corrosive chloride salts in the process liquid.

In the MEL process, the slurry is aerated for the same reason, but in a separate tank, ultimately producing a gypsum cake similar to the LSFO process. Water removed from the gypsum and soluble magnesium salts are recycled to the process with a portion removed as wastewater for chloride control.

1.1 PROCESS CHEMISTRY

1.1.1 LSFO Process

A simplified explanation of the LSFO process (Figure 1) is that the SO_2 absorbed in the recirculated slurry reacts with dissolved limestone ($CaCO_3$) in the slurry to form calcium sulfite hemihydrate ($CaSO_3 \bullet \frac{1}{2}H_2O$) according to the following reaction:

$$SO_2 + CaCO_3 + \frac{1}{2} H_2O \Rightarrow\rightarrow CaSO_3 \bullet \frac{1}{2}H_2O + CO_2$$

Carbon dioxide formed from reaction of limestone with SO_2 is released into the flue gas.

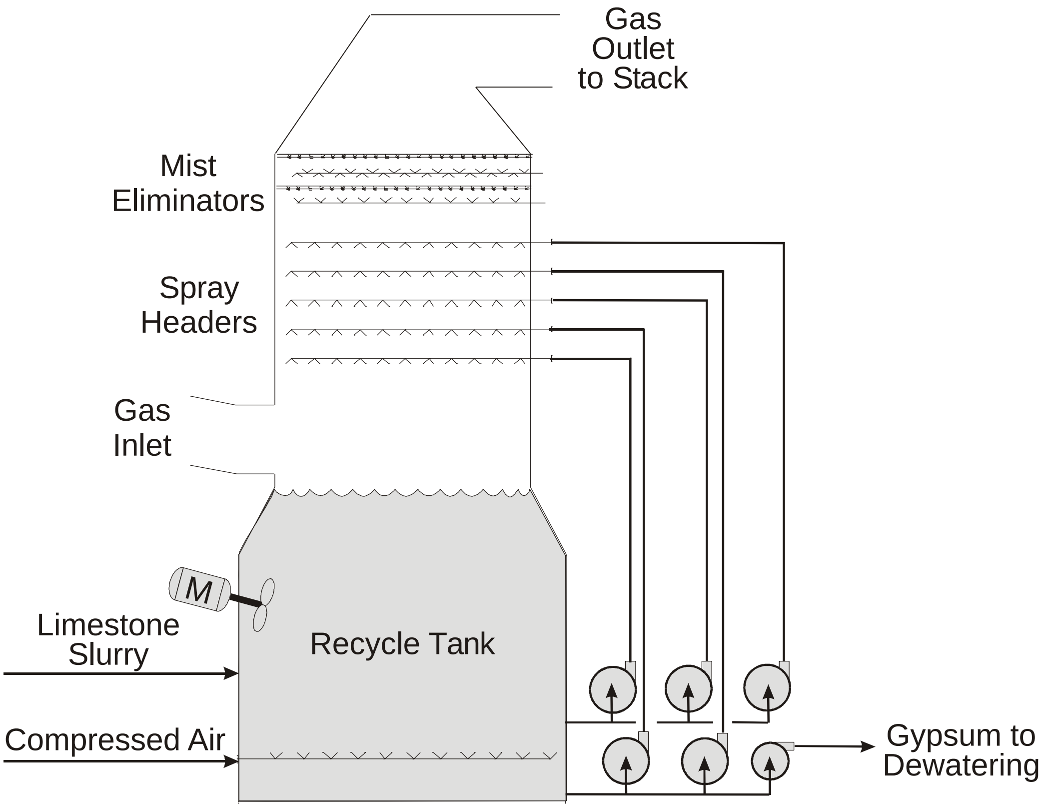
Oxidation air is bubbled through the slurry to convert $CaSO_3 \bullet \frac{1}{2}H_2O$ to gypsum ($CaSO_4 \bullet 2H_2O$) according to the following reaction:

$$CaSO_3 \bullet \frac{1}{2}H_2O + \frac{1}{2}O_2 + 1.5\, H_2O \Rightarrow\rightarrow CaSO_4 \bullet 2H_2O$$

Aeration oxidizes all the calcium sulfite to calcium sulfate and forces precipitation to occur on existing gypsum crystals in the reaction tank. This minimizes tendency for gypsum to precipitate on surfaces in the absorber and cause plugging of pipes and nozzles by maintaining gypsum concentration in absorber slurry liquid at 80-90% of the concentration at which scale would form. The LSFO process also improves slurry dewatering because the gypsum crystal size and shape dewaters more freely than $CaSO_3 \bullet \frac{1}{2}H_2O$ crystals produced by older natural oxidation systems. Physical stability of dewatered gypsum is also improved by the more granular particles, allowing gypsum to be disposed of in a landfill (if no sales opportunities exist) without requiring fly ash or lime fixatives.

SO_2 collection efficiency and energy consumption of an LSFO system can be improved by adding organic carboxylic acids to the reaction tank. These additives serve as a buffer to stabilize the pH of the slurry as it

Figure 1: LSFO Process SO_2 Absorber



absorbs the acidic SO_2. This improves absorber performance by increasing solubility of SO_2 in the slurry which allows operation at a lower L/G ratio. It also allows operation at a lower slurry pH, which increases the rate of dissolution of limestone and increases reagent utilization. (Reagent utilization is the percentage of reagent fed to the FGD system that reacts with SO_2. A portion of the reagent does not react, which increases required reagent feed rate.) The organic acid is not consumed by the absorption process, but replenishment is required to replace acid lost through gradual oxidative degradation in the absorber and to replace acid that is carried out in moisture remaining with the gypsum cake and in wastewater.

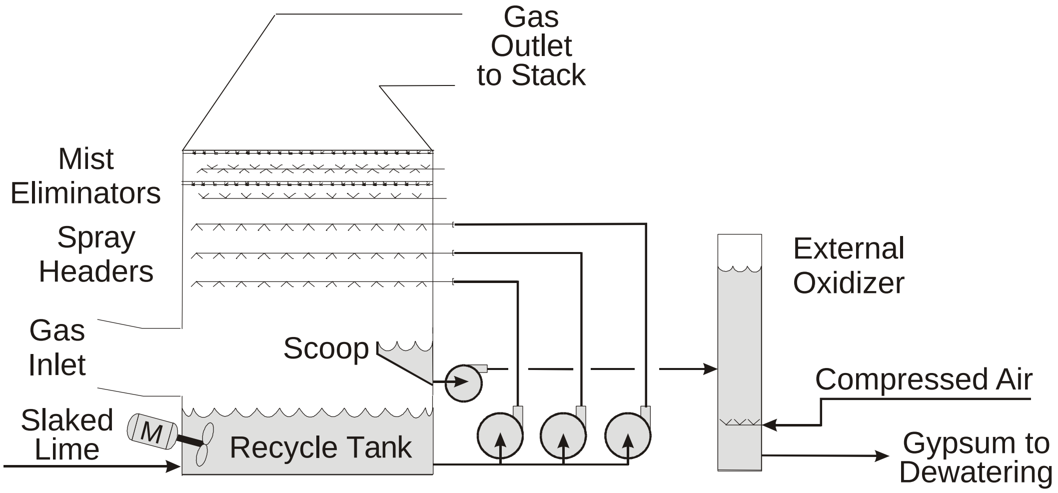
A number of organic acids have been found to be effective. These acids include adipic, glycolic, maleic, acrylic, and formic acids. Dibasic acid (DBA), a by-product from the manufacture of adipic acid, is usually selected instead of other acids because of its lower cost (about $460 per ton). (DBA is a mixture of adipic, glutaric, and succinic acids.) If wastewater treatment is required, organic acids have a biological oxygen demand, which may require removal in an additional water treatment plant.

In general, LSFO systems are designed without depending on use of organic acids. Provision is sometimes included in designs to allow future addition in case SO_2 collection efficiency is inadequate. Disadvantages of organic acids include additional operating cost, uncertainty in long-term supply and pricing of DBA by-products, possible contamination of gypsum by-product, and increased wastewater treatment cost.

1.1.2 MEL Process

In the MEL process, slaked lime, containing calcium hydroxide [$Ca(OH)_2$] and a portion of magnesium hydroxide [$Mg(OH)_2$], is used to react with SO_2 (Figure 2). As with the LSFO process, the slurry is added to a recycle tank at the bottom of the absorber under pH control to replenish reagent consumed. Overall, SO_2, which is a strong acid, reacts with and is neutralized by slaked lime. The overall reactions between lime and SO_2 in the absorber are shown in reactions (1) and (2). Calcium hydroxide in the slurry reacts with most of the SO_2 to precipitate calcium sulfite ($CaSO_3 \bullet \frac{1}{2} H_2O$). Magnesium hydroxide reacts with the remainder of the SO_2 to form soluble magnesium salts, magnesium sulfite and magnesium bisulfite [$MgSO_3$, $Mg(HSO_3)_2$].

Figure 2: MEL Process SO_2 Absorber with Forced Oxidation



These soluble magnesium salts greatly increase SO_2 capture and allow reduction in power consumption and equipment costs. When magnesium sulfite is present in slurry in contact with flue gas, it buffers (prevents from falling sharply) pH of the slurry as it absorbs acidic SO_2. This improves absorber performance by increasing solubility of SO_2 in the slurry, which allows operation at a lower L/G ratio. It also allows operation at a lower slurry pH, near 6.0, which improves reagent utilization to near 100%. Magnesium sulfite is not consumed by the absorption process, but is replenished by addition of fresh lime slurry to the reaction tank. This provides the MEL FGD processes a greater capacity to absorb SO_2 than LSFO. Since magnesium sulfite performs the same function as described for organic acids in LSFO systems, no organic acid addition is required to improve SO_2 collection efficiency. A more detailed description of MEL process chemistry is given in Appendix B.

The magnesium salts also prevent formation of build-ups on surfaces in the absorber. The salts suppress solubility of hard-scale-forming substances like calcium sulfate. This eliminates tendency for calcium sulfate to precipitate on surfaces in the absorber and cause plugging of pipes and nozzles by maintaining the concentration of calcium sulfate in absorber slurry liquid at about 10% of the concentration at which scale would form.

$$SO_2 + Ca(OH)_2 \rightarrow CaSO_3 \bullet \tfrac{1}{2} H_2O + \tfrac{1}{2} H_2O \tag{1}$$

$$4SO_2 + 3Mg(OH)_2 \rightarrow 2MgSO_3 + Mg(HSO_3)_2 + 2H_2O \tag{2}$$

A portion of recirculated slurry containing absorbed SO_2 in the form of calcium sulfite and magnesium salts is collected in a scoop from recirculating slurry that has just contacted the flue gas (Figure 2). Slurry collected in the scoop has lower pH (~5.5), which makes it more suitable for the subsequent oxidation step. The slurry is pumped to an external forced oxidation tank where it is contacted with air. Calcium sulfite is converted to gypsum and magnesium salts are oxidized to $MgSO_4$, according to reactions (3) and (4) & (5), respectively. The gypsum precipitates in a crystalline form while $MgSO_4$ remains in solution.

The gypsum in the slurry effluent from the oxidizer undergoes primary and secondary dewatering to produce gypsum by-product cake in the same way as in the LSFO process. Liquid containing $MgSO_4$ is returned to the absorber after dewatering the gypsum.

$$CaSO_3 \bullet ½H_2O + ½O_2 + 3/2\ H_2O \rightarrow CaSO_4 \bullet 2H_2O \tag{3}$$

$$MgSO_3 + ½O_2 \rightarrow MgSO_4 \tag{4}$$

$$Mg(HSO_3)_2 + O_2 \rightarrow MgSO_4 + H_2SO_4 \tag{5}$$

1.2 REAGENTS AND BY-PRODUCTS

1.2.1 LSFO Process

Preparation of limestone slurry reagent involves grinding a high-calcium, crushed (less than 1 inch) limestone to a fine size [95% smaller than 325 mesh (44 microns)] in a wet ball mill. The fine size provides for more complete use of the limestone to maximize reagent utilization and to minimize the amount of unreacted limestone in the gypsum product. The ground limestone is slurried with water and held in an agitated tank for use. The slurry reagent is fed to the absorber to replenish limestone consumed in the reaction and the feed rate is typically controlled based on pH of the circulating slurry with feed-forward inlet SO_2 and flue gas flow rate signals.

The by-product gypsum, after dewatering to 90% solids (10% moisture), can be sold or disposed in a landfill, either alone or blended with fly ash. Gypsum can be used for making wallboard, as an additive for concrete, or as an agricultural soil conditioner.

1.2.2 MEL Process

Lime used in the process is composed of 83 to 91 weight percent calcium oxide (CaO) and 2 to 8 weight percent magnesium oxide (MgO). Total oxides content is equivalent in SO_2 neutralizing value to about 94% CaO. Magnesium-enhanced lime can be produced from limestone with suitable magnesium content or by blending high-calcium and dolomitic (~40% MgO) limes. Magnesium oxide content of lime is adjusted by the supplier to match FGD systems' requirements for SO_2 collection efficiency. This includes adjusting MgO content so the amount added to the FGD system exceeds the equivalent amount of hydrogen chloride absorbed. (For example, for 3% sulfur and 0.12% chlorine in coal, 1.2% MgO in lime is required to accommodate hydrogen chloride; for coal with 1.3% sulfur and 0.1% chlorine, 2.3% MgO is required.) Magnesium oxide which reacts with hydrogen chloride is not available to react with SO_2 to form magnesium sulfite, which has the beneficial effects on SO_2 collection efficiency and scale prevention discussed earlier.

To prepare it for use, the lime is mixed with water in a slaker where it is hydrated to form calcium hydroxide [$Ca(OH)_2$] and magnesium hydroxide [$Mg(OH)_2$]. The resulting slurry is added to the absorber recycle tank to replenish lime consumed in the process; pH is controlled to a set point near 6.0. At this pH nearly 100% of the lime reacts with SO_2.

As in the LSFO process, the by-product gypsum, after dewatering to 10% moisture, can be sold or disposed of in a landfill, either alone or blended with fly ash. This gypsum is also used for manufacturing wallboard, as an additive for concrete, or as an agricultural soil conditioner. FGD wastewater containing magnesium sulfate can be directed to a wastewater treatment plant. Magnesium sulfate has no biological oxygen demand, so no additional (biological) wastewater treatment is required. In this study, wastewater treatment requirements are assumed to be the same for the MEL and the LSFO processes. No additional cost is assumed for disposing magnesium sulfate. Optionally, a second by-product, magnesium hydroxide, can be produced by treating the FGD wastewater with additional lime prior to directing it to the wastewater treatment plant. Magnesium hydroxide recovery reduces salinity and toxic metals content of wastewater, which reduces wastewater treatment requirements. Magnesium hydroxide has a number of commercial uses, including neutralization of sulfur trioxide in flue gases in the furnace or at the economizer outlet, thereby reducing corrosion in the air heater, and lowering sulfuric acid plume potential.

1.3 SO_2 REMOVAL PERFORMANCE

1.3.1 LSFO Process

LSFO systems have achieved SO_2 removal efficiencies as high as 98% in power plants firing a variety of high- and low-sulfur fuels. For the LSFO absorber, the higher removal efficiencies are typically attained through the use of additives such as DBA in the recirculating slurry or by recirculating larger quantities of slurry (larger L/G ratio) in the absorber. This portion of the study focuses on the use of L/G ratios to achieve design SO_2 capture rather than being dependent on DBA , which avoids potential by-product contamination and additional wastewater treatment.

Most of the scrubbers installed in Phase 1 of the CAAA title IV program (1995) were designed for and achieved 95% efficiency with L/Gs of 90-130 and inlet sulfur dioxide up to 8 lb/MBtu. Demonstrations and testing by the major FGD process developers, including Alstom, Mitsubishi, Babcock & Wilcox, and Wheelabrator, have shown that a 130 L/G is adequate to achieve 98% efficiency in a typical open-spray tower design on the 4.72 lb/MBtu inlet SO_2 basis (high-sulfur coal case) of this work. This is further verified by recent guarantees offered by FGD vendors for new unit applications. It should be noted that 98% SO_2 removal for a LSFO system is, in most cases, not the optimal design point for the process. Efficiencies above 95% require exponentially greater mass transfer (e.g., L/G) capability and will only be pursued if air permit requirements necessitate very low SO_2 emissions. Note also that these very low SO_2 emissions have been required on some recent air permit applications.

This being the case, for the Appalachian high- and low-sulfur coals, we have estimated that L/G ratios of 130 and 80, respectively, will be required to achieve 98% SO_2 removal efficiency without the use of organic acid in a typical LSFO absorber design. Limestone reagent utilization of at least 97% is necessitated by gypsum quality requirements.

Regarding emission averaging period, typically, system design and guarantee requirements are requested to be demonstrated over a short and manageable test period. This test period is done usually on a one-time or two-time basis, and consists of 3 two- to four-hour tests. Margin is specified into the design (efficiency, equipment redundancy, etc.) to meet the plant's emission averaging requirements.

1.3.2 MEL Process

MEL forced oxidation systems have achieved a better level of performance than the LSFO process, with SO_2 removal efficiencies between 98% to 99% in power plants also firing a variety of high- and low-sulfur coals. Because of the higher SO_2 absorption capacity available in the magnesium-enhanced lime system compared with the LSFO system, we estimate that L/G ratios of 40 and 30 will be required to achieve 98% SO_2 removal efficiency for the Appalachian high- and low-sulfur fuels, respectively.

The difference in the L/G ratio requirements between the LSFO and MEL processes has a major impact both on capital and O&M cost differences between the two competing technologies.

1.4 COMMERCIAL STATUS

Both the LSFO and the MEL FGD systems are operating successfully at many coal-fired power facilities, ranging in size from less than 100 MW to 1000 MW. Applications include upgrading commercial designs for coal sulfur content well above 5%. Many competing designs are available in the marketplace, from a number of well-respected and viable suppliers, including:

- Alstom
- Babcock & Wilcox
- Babcock Power, Inc.

- Chiyoda
- Hamon Research Cottrell
- Mitsubishi
- Wheelabrator Air Pollution Control

2. WET FGD PROCESS ADVANTAGES AND DISADVANTAGES COMPARED TO DRY FGD TECHNOLOGY

2.1 PROCESS ADVANTAGES

The LSFO and MEL technologies have the following advantages when compared with other FGD technologies:

1. Well-established FGD technology on a variety of world coals with proven reliability.
2. SO_2 removals of 95% are common and removals as high as 98% can be attained.
3. Adequate and commercially viable suppliers offer the technology.
4. Reagents used by the process are plentiful and readily available.
5. Waste gypsum is stable for landfills without blending with fly ash and lime.
6. Can be designed to produce wallboard-grade gypsum as a saleable by-product.
7. The FGD system is not sensitive to boiler operational upsets and typical operating modes, such as cycling duty.

2.2 PROCESS DISADVANTAGES

The LSFO and MEL technologies can have the following disadvantages when compared with other FGD technologies:

1. The LSFO process circulates large quantities of slurry with the attendant high pumping power consumption.
2. The pressure drop across the absorber increases the induced draft (ID) fan power consumption.
3. These processes can produce a large volume of gypsum. The salability of this by-product is dependent on a sufficiently sized gypsum market near the plant.
4. The high potential for corrosion requires extensive use of costly corrosion-resistant alloys or nonmetallic liners as materials of construction for the absorber and other system components.

3. DESIGN BASIS

3.1 SPECIFIC DESIGN CRITERIA

Table 3.1 -1 lists the specific design criteria.

TABLE 3.1-1
SPECIFIC DESIGN CRITERIA

	LSFO	**LSFO**	**MEL**	**MEL**
Unit capacity	500 MW	500 MW	500 MW	500 MW
Heat input to boiler, MBtu/hr	5000	5000	5000	5000
Fuel	High-sulfur Appalachian	Low-sulfur - Appalachian	High-sulfur - Appalachian	Low-sulfur - Appalachian
Fuel analysis, % wt.:				
Moisture	3.1	6.0	3.1	6.0
Ash	9.0	9.1	9.0	9.1
Carbon	69.82	72.6	69.82	72.6
Hydrogen	5.0	4.8	5.0	4.8
Nitrogen	1.26	1.4	1.26	1.4
Sulfur	3.0	1.3	3.0	1.3
Oxygen	8.7	4.7	8.7	4.7
Chlorine	0.12	0.1	0.12	0.1
High heating value, Btu/lb	12,720	13,100	12,720	13,100
SO_2 generation*, lb/MBtu	4.72	2.0	4.72	2.0
Coal feed rate, tons/hr	197	191	197	191
Flue gas flow at FGD inlet**, macfm	1.75	1.70	1.75	1.70
Flue gas temperature at FGD inlet, °F	300	280	300	280
Flue gas flow at FGD outlet, macfm	1.52	1.50	1.52	1.50

TABLE 3.1-1
SPECIFIC DESIGN CRITERIA

	LSFO	LSFO	MEL	MEL
Flue gas temperature at FGD outlet, °F	130	130	130	130
SO_2 reduction efficiency, %	98	98	98	98
SO_2 outlet, lb/MBtu	0.094	0.04	0.094	0.04
Mercury concentration in coal, ppmd	0.08 to 0.15	0.08 to 0.15	0.08 to 0.15	0.08 to 0.15

* All sulfur in coal is assumed to be converted to SO_2 (no SO_3), and that there is no sulfur removed in the either the bottom ash or electrostatic precipitator ash.

** The air leakage for the new plant and retrofit plants are assumed to be the same over a long duration.

Table 3.1-2 summarizes the parameters used for the FGD comparison.

TABLE 3.1-2
PARAMETERS USED FOR FGD COMPARISON

	LSFO	LSFO	MEL	MEL
Unit capacity	500 MW	500 MW	500 MW	500 MW
Heat input to boiler, MBtu/hr	5000	5000	5000	5000
Fuel	High-sulfur - Appalachian	Low-sulfur - Appalachian	High-sulfur - Appalachian	Low-sulfur - Appalachian
SO_2 removal, %	98%	98%	98%	98%
SO_2 emission, lb/MBtu	0.094	0.04	0.094	0.04
By product	Gypsum	Gypsum	Gypsum	Gypsum
Power consumption*, % of MW_{net}	2.0	1.3	1.4	1.0
Reagent	Limestone	Limestone	MEL	MEL
Reagent cost, $/ton	15	15	50	50
Reagent purity, %	95	95	94	94
L/G ratio, gpm per 1000 acfm at absorber outlet	130	80	40	30
Reagent stoichiometry**, moles of reagent/mole of sulfur removal	1.03	1.03	1.02	1.02

TABLE 3.1-2
PARAMETERS USED FOR FGD COMPARISON

	LSFO	LSFO	MEL	MEL
SO_2 oxidation stoichiometry (O/SO_2 removed)	3.0	3.0	3.0	3.0
Load factor, %	80	80	80	80
FGD system life, years:				
New unit application	30	30	30	30
Retrofit application	20	20	20	20
Capital cost leveling factor, %/year:				
New unit application	14.5	14.5	14.5	14.5
Retrofit application	15.43	15.43	15.43	15.43
Discount rate, %	8.75	8.75	8.75	8.75
Inflation rate, %	2.5	2.5	2.5	2.5
Operating cost levelization factor:				
New unit application	1.30	1.30	1.30	1.30
Retrofit application	1.22	1.22	1.22	1.22

*Auxiliary power, including estimated calculations, is summarized in Appendix A.

** If the LSFO process is to produce a saleable grade gypsum, a reagent stoichiometry (moles reagent per mole SO_2 absorbed, equal to inverse of reagent utilization) of 1.03 or lower is required, along with greater than 95% limestone purity. The capital cost takes into account the size of the reaction tank to achieve higher reagent utilization and crystal growth for the LSFO process.

3.2 SYSTEM COMPONENTS

The design criteria and features selected both for the LSFO and the MEL FGD systems are considered typical for today’s standard practices. The following descriptions present important features of the components and assumptions used in this study:

3.2.1 Reagent Handling and Preparation Systems

3.2.1.1 LSFO Process

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 16-hour capacity day bin. The limestone day bin and a gravimetric feeder supply the limestone to a 150% capacity ball mill system. This allows two-shift operation of the reagent preparation system with the boiler continuously operating at 100% load. The ball mill grinds the limestone to 90% less than 325 mesh and uses a wet recycle classification loop to ensure proper size distribution in the product. Two 100% capacity classification pumps are used to provide high reliability of the classification system. The process makeup water or the 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 limestone slurry tank prior to being fed as reagent makeup into an absorber.

3.2.1.2 MEL Process

Lime is received by truck and stored in a 14-day capacity bulk storage lime silo. The lime is pneumatically conveyed to a 16-hour capacity day bin. The lime day bin and a gravimetric feeder supply the lime to a 150% capacity slaking system. This will allow two-shift operation of the reagent preparation system with the boiler continuously operating at 100% load. A modern, conventional lime slaker with a high efficiency grit removal and lime recovery system is used. Two 100% 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 classification to produce 20% solids product slurry. The slurry is fed to the absorber by a dedicated reagent feed pump (100% spare capacity provided).

A 100-gallon capacity sulfuric acid tank with sulfuric acid feed pumps is also provided.

3.2.2 SO_2 Removal System

3.2.2.1 LSFO Process

One 100% capacity absorber is provided to achieve 98% SO_2 removal efficiency. The absorber is a vertical open-spray tower, with countercurrent contact between the flue gas and scrubber slurry. To achieve 98% SO_2 reduction with high-sulfur coal, 4 operating pumps will be provided to achieve a 130 L/G ratio. Three operating pumps will be provided for low-sulfur coal to achieve an 80 L/G ratio. Each individual pump will be dedicated to a spray level in the absorber. One spare spray level is provided along with a dedicated pump with each design. The entrained slurry droplets are removed from the flue gas through a chevron-type mist eliminator. The scrubber slurry drains into a reaction tank located in the bottom of the absorber. This tank is agitated and is sized for a minimum of 15 hours of solid 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 oxidized air compressors. The bleed slurry from the absorber is pumped to the dewatering area. Makeup limestone slurry is added to the reaction tank. The material of construction for the absorber is 317 LMN steel. The chloride levels will be maintained to less than 8,000 ppm.

The majority of water loss is through evaporation, which is assumed to be the same for both processes. Since it is also assumed that both processes remove the same amount of sulfur from the gas, converting it into a saleable grade gypsum, both processes will have same water losses in the gypsum by-product. The blowdown is based on the chloride level, which, again, will be the same for both processes. Therefore, the water losses and consumption will be the same for both wet processes.

3.2.2.2 MEL Process

One 100% capacity absorber is provided to achieve 98% SO_2 removal efficiency. The absorber is a vertical open-spray tower, with countercurrent contact between the flue gas and scrubber slurry. To achieve 98% SO_2 reduction with high-sulfur coal, 2 operating pumps will be provided to achieve a 40 L/G ratio. Two operating pumps will also be provided for low-sulfur coal to achieve a 30 L/G ratio. Each individual pump will serve a dedicated spray level in the absorber. One spare spray level is provided, along with a dedicated pump, for both

L/G designs. The absorber is constructed of 317 LMN steel or equivalent. The chloride levels will be maintained to less than 8,000 ppm. The entrained slurry droplets are removed from the flue gas through a chevron mist eliminator. Makeup lime slurry is added to the recycle tank. A trough (scoop) system is provided to direct a small portion of the recirculated slurry to an oxidation tank for the conversion of $CaSO_3$ to $CaSO_4$. The remaining slurry falls into the recycle tank.

The oxidation tank is sized for 4 hours of solid residence time. To achieve greater than 99% oxidation of $CaSO_3$ to $CaSO_4$, a 3:1 stoichiometric ratio of oxygen to removed SO_2 is used in the oxidation compressor design. Compressed air is injected below the oxidation tank agitator through a sparger network. The bleed slurry from the oxidation tank is pumped to the dewatering area.

An Oxallizer™, by EIMCO, is currently being demonstrated and has the potential to reduce overall costs. EIMCO has installed four Oxallizers™ at the Lowman power station, which uses the MEL process. Those Oxallizers™ produce a total of 12-15 t/hr of gypsum, using a bleed stream from the absorber. An Oxallizer™ unit is also being demonstrated at a confidential site. The Oxallizers™ may effectively compete against sparged oxidizers and, thus, must be given consideration to determine the overall capital and O&M costs during the process selection.

S&L contacted EIMCO to discuss the application of the Oxallizers™ process to the hypothetical applications of the MEL process in this study. On a total cost of electricity impact basis, the Oxallizers™ process is approximately 10% more costly than the conventional process under the assumptions used in this analysis.

The capital costs presented in this report are based on the conventional sparged oxidation system rather than the Oxallizer™.

3.2.3 Flue Gas System/Stack

3.2.3.1 LSFO Process

The gas path of the system will start at the discharge of the existing ID fans, through the new booster ID fans and absorber, and discharge into a new acid brick chimney. The draft requirements of the LSFO FGD system will be accommodated by the booster ID fans. The booster ID fan is sized to provide an additional 10" H_2O (9" operating) and 9” H_2O (8" operating) pressure drop through the absorber for high- and low-sulfur applications, respectively. The inlet ductwork to the absorber system, including the booster ID fan, will be fabricated with carbon steel. The outlet ductwork from the absorber to the stack bleeding will be constructed of a carbon steel substrate with a 1/16-inch “wallpaper” of C-276 alloy for corrosion protection.

3.2.3.2 MEL Process

The flue gas system for the MEL process will be identical to the LSFO process, with the exception that the draft requirements will be reduced to 8” H_2O (7” operating).

3.2.4 By-Product Treatment and Handling System

The by-product system will be similar both for the LSFO and MEL processes.

The bleed stream from the reaction tanks for the LSFO process and from the oxidation tank for the MEL process is pumped to a set of hydroclones for primary dewatering. The hydroclone overflow is recycled back to the process. The hydroclone underflow, containing 50% solid, is fed to a belt filter for secondary dewatering. Two 100% belt filters are designed to achieve 90% solids in the gypsum by-product. The gypsum cake is washed with fresh water on the belt filters to remove any residual chlorides to a level of less than 100 ppm of chloride in the dry solids. The filtrate from the horizontal belt filter is returned to the process. A small blowdown from the process filtrate is required to remove chloride from the process. For purposes of this study, we have assumed that this chloride purge stream can be disposed of in an onsite landfill/pond or used as a wetting agent for ash transportation.

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

3.2.5 General Plant Support Systems

The general support equipment includes modifications and/or additions to the plant seal water system, instrument air system, makeup water system, distributed control system (DCS), and electrical auxiliary power system and accommodations in the BTG 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.

Table 3.2.5–1 lists the equipment used in each subsystem.

TABLE 3.2.5 -1 **EQUIPMENT USED IN EACH SUBSYSTEM**
Reagent Handling and Preparation
LSFO Process:
Truck unloading system
Limestone storage pile (30 days' storage)
Limestone reclaim system
Limestone day bins (16 hours' storage)
Limestone crushers
Limestone ball mills (150% capacity) system
Slurry storage tank (16 hours' storage)
MEL Process:
Truck unloading system
Lime bulk storage steel silo (14 days' storage)
Lime live reclaim system
Lime day bins (16 hours' storage)
Lime slakers (150% capacity)
Slurry storage tank (16 hours' storage)
SO_2 Removal System
LSFO Process:
1 x 100% module with reaction tank and spargers
5 recirculation pumps (4 operating, 1 spare)
2 x 100% air compressors
MEL Process:

1 x 100% module with reaction tank and slurry trough
3 recirculation pumps (2 operating, 1 spare)
Oxidation tank with spargers
2 x 100% air compressors
By-Product Treatment and Handling System
Hydroclones (20% spare hydroclones)
Hydroclone underflow surge tank
Belt filters (2 x 100%)
Conveyor to storage shed
Gypsum storage shed
Recycle process water tank
Flue Gas System
Booster ID fans (2 x 50%)
Absorber inlet ductwork/dampers
Absorber outlet ductwork/dampers
Acid brick lined stack for retrofit applications
Bypass duct to existing stack
General Plant Support Systems
Makeup water tank, pumps, piping
Instrumentation/plant air compressors (2 x 50%)
Auxiliary power transformers/switchgear/motor controls, etc.
Electrical conduit, duct banks, wiring, cables, etc.
Control system (DCS)
Site accommodations:
Roads
Storm sewers
Sanitary systems
Fire protection

4. TECHNOLOGY APPLICATION CONSTRAINTS

Summarized below are the typical application constraints for conventional wet FGD technology.

4.1 UNIT/ABSORBER SIZE

Conventional wet FGD systems are in successful operation at a large number of coal-fired facilities ranging in size from small, individual applications to large utility applications of more than 1000 MW. Single absorbers capable of treating the fuel gas flow from a 1000 MW bituminous coal-fired plant are commercially demonstrated and available from industry suppliers.

4.2 COAL SULFUR CONTENT

Wet FGD systems have been successfully applied to coals of all sulfur levels. Most of these systems are installed on medium- to high-sulfur fuels (>2%); however, a number of systems are also installed on low-sulfur fuels (<1%) in the western areas of the United States.

4.3 PERFORMANCE EXPECTATIONS

The SO_2 removal guarantees of up to 98% are available from the system suppliers and have been demonstrated in commercial applications.

4.4 SO_2 REDUCTION

Suppliers of FGD systems have guaranteed SO_2 reduction efficiency up to 98% with the inlet SO_2 concentration up to 10.0 lb/MBtu. To the best of our knowledge, the lowest outlet SO_2 loading guaranteed by a supplier of FGD systems is 0.06 lb/MBtu.

4.5 REAGENT UTILIZATION

Reagent utilization is the percentage of reagent fed to the FGD system that reacts with SO_2. Reagent utilization in a wet FGD system is typically very close to the required theoretical level when considered on a removed SO_2 basis. Limestone reagent utilization in LSFO systems is typically greater than 97%. This high level of reagent utilization is also required to maintain the quality of gypsum if it is to be used for wallboard manufacturing. Lime reagent utilization is near 100% due to high reactivity of lime and small particle size of slaked lime (3 microns average).

"Stoichiometry" is a related measure of reagent utilization. Stoichiometry is the ratio of reagent fed to SO_2 absorbed (mole/mole). Stoichiometry is equal to 100 divided by reagent utilization. For example, a reagent utilization of 97% is equal to a stoichiometry of 1.03.

4.6 BY-PRODUCT QUALITY

The by-product from either the wet LSFO process or the MEL process can be fashioned for use in the cement industry, for wallboard manufacturing, or for agricultural use. The by-product quality requirements are much more stringent for wallboard manufacturing compared to the other uses.

4.7 ENERGY CONSUMPTION

The process energy consumption results primarily from the booster ID fan power required to overcome the draft loss across the absorber and the power requirement for recirculation pumps. The MEL process has an advantage over the LSFO process because the power requirement for slurry recirculation is 71% lower (high-sulfur case) than for the LSFO process and the power requirement for the booster ID fan is 22% lower (high-sulfur case) than for the LSFO process. Overall, the MEL process will require approximately 0.6% less auxiliary (% of MW_{net}) power for high-sulfur coal and approximately 0.3% less auxiliary power auxiliary (% of MW_{net}) for low-sulfur coal compared to the LSFO process.

4.8 RETROFIT VERSUS NEW UNITS

In new unit and retrofit applications, wet FGD technology is typically installed between the electrostatic precipitator/baghouse outlet and the stack.

Most of the retrofit units will have booster fans to overcome the pressure drop across the FGD absorber, which will be located after the existing ID fans. However, new units are expected to be installed with ID fans large enough to overcome the pressure drop across the FGD absorber. This feature will typically result in a lower capital cost for the draft system on a new unit application compared with a retrofit application.

Most retrofit units will not be able to use existing stacks, as these stacks are typically designed for hot flue gas approximately at a 100 ft/sec exit velocity. To accommodate the saturated flue gas from a wet FGD system, wet stacks are typically designed for a highly corrosion-resistant material with a gas velocity of between 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 of the retrofit units will, therefore, require a new stack.

5. COSTS ANALYSES

5.1 CAPITAL COSTS

Estimated capital costs for the high- and low-sulfur coal applications both of the LSFO and the MEL wet FGD systems were developed for new and retrofit applications. These estimates encompass all of the required equipment, materials, and construction associated with the new and retrofit installations of these technologies.

The costs were developed primarily from Sargent & Lundy's cost database, which is continuously updated from ongoing work in the area of FGD.

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

- Equipment and material (FGD system and balance-of-plant)
- 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 encompass sales tax or property tax. License fees and royalties are not expected for these FGD technologies. Owner's cost includes owner's project management, engineering, accounting, billing, etc., as well as any "corporate" charges that may apply to capital projects. In many cases, owners will also place capital financing charges in this category. We have assumed in our analysis that the capital cost estimate is an "overnight" cost and therefore no financing charges or escalation are included.

Salient features of each capital cost estimate are described in Section 3 of this report.

Based on our experience and the level of effort applied to this study, we estimate that the relative accuracy of the estimate prepared is ±20%.

Additionally, an underlying assumption is that the contracting arrangement for the implementation of an FGD project is based on multiple lump sum specialty work packages. If the FGD project is to be executed on an engineer, procure, construct (EPC) or turnkey basis, a separate risk allocation should be added to the estimate of 10% to 20% (1.1 or 1.2 multiplier) to cover the EPC contractor's contingency, pass-through mark-up, and profit.

Exhibit 5–1 and Exhibit 5–2 present the capital costs for new units and retrofit units, respectively.

5.2 OPERATIONS AND MAINTENANCE COSTS

Exhibit 5–3 through Exhibit 5–6 present the estimated O&M costs associated with the wet and dry FGD systems. These include both fixed and variable operating costs, defined as follows:

5.2.1 Fixed O&M Costs

The fixed O&M costs consist of O&M labor, maintenance material, and administrative labor.

For purposes of this study, the installation of the FGD system has been anticipated to add an additional 8 operators to the current pool of operating labor for new units and 12 operators for the retrofit application. It is assumed the plant layout for retrofit application is not optimized, which would require more operating labor than would a new unit. Four extra personnel are added for a retrofit application to cover the impact of a non-labor-efficient plant layout and control room.

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 equipment, the required maintenance material, and the labor required to perform the maintenance.

5.2.2 Variable O&M Costs

Variable O&M costs developed for each FGD system include the cost of all consumables, including reagent (limestone or lime), water, and power requirements. We have assumed that gypsum will be "sold" for $0/ton.

Auxiliary power costs reflect the power requirements associated with the operation of the fans (incremental ID fan power for new units and booster ID fan power for retrofit units) as well as the power consumption for ball mills/slakers, recirculation slurry pumps, compressor for oxidation air, gypsum dewatering system, and various electrical and control users typically needed for FGD operations.

The FGD system auxiliary power requirement has two types of impact on a coal plant:

1. An increase in the net plant heat rate, resulting in higher annual consumption of fuel and the other materials and services that are directly proportional to furnace heat input, such as any required FGD reagents and ash disposal.
2. A reduction in net plant MW capacity.

The economic impacts of Item 1 are a direct function of the delivered fuel price, gross turbine heat rate, boiler efficiency, coal ash content, coal sulfur content, ash disposal cost, reagent costs, by-product values, etc.

The economic impact of Item 2, can be considered either from the standpoint of an independent power producer (IPP) on a competitive grid or from the standpoint of an integrated electric utility operating a power grid under regulation. The economic impact would essentially be identical under an IPP or integrated utility scenario; however, each situation must be considered on a case-by-case basis.

From the IPP point of view, the capacity reduction represents (a) loss of any capacity revenues (including ancillary services associated with reserve margin support) that the plant would have received during the hours when the plant was operating at full load, plus (b) loss of contribution margin during that period represented by the hourly marginal price of power on the system, less variable operating costs of the unit under study. From the integrated utility point of view, the economic losses include the cost of buying replacement capacity and replacement energy, where the net replacement energy cost would be the market price of power, less the variable costs (primarily fuel costs) that would be avoided. Either value will be a function of the hours per year that the plant operates at 100% load. The hours per year that the coal plant will operate at full load are a function of the load-duration curve of the system the plant is serving and the competitiveness of this plant relative to other plants connected to this grid. For this reason, each situation must be considered individually in the context of those parameters.

In the short term, capacity may have very little value in some markets, as indicated Table 5.2.2 -1 below of recent total spot market costs for electricity:

TABLE 5.2.2 - 1
RECENT TOTAL SPOT MARKET ELECTRICITY COSTS

	$/MWh	
Region	**On Peak**	**Off Peak**
Western United States:		
Mid-Columbia	26.42	22.67
CA-OR Border	31.19	24.42
NP 15	38.22	26.26
SP 15	37.05	25.42
Palo Verde	29.24	20.15
Gulf/Southeast:		
SPP	29.03	14.57
ERCOT	33.75	24.50
SERC	28.72	16.58
FRCC	31.50	21.00
Midwest:		
ECAR	26.72	14.10
MAIN	27.59	13.71
MAPP	31.99	16.26
East Coast:		

NEPOOL	46.75	34.00
NY Zone J	68.38	41.70
PJM West	28.62	18.33

The above table shows a range of slightly less than \$15/MWh to just over \$40/MWh, with an average of \$20/MWh for the off peak periods (most of the hours of the day). Over the longer term, electricity prices are likely to be higher in most regions, and a portion of those prices will be attributable to the value of capacity. For an optimized power system, the market value of capacity typically will be approximately equal to the carrying charges on the lowest cost type of capacity (simple cycle combustion turbines), equal to about \$55/kW/year. Depending on unit size, fuel, financing, etc., the levelized total cost of electricity for a modern coal plant will be in the neighborhood of \$45/MWh, covering capital, O&M, and fuel costs. Market prices for base-load electricity must reach that general level over the long term for cost-effective installation of new coal-fired generating plants.

Although local power prices are depressed in some markets today, for *long-term* FGD studies involving new coal-fired plants that are expected to be economically viable over the long term, a reasonable assumption for the cost of auxiliary power in feasibility studies is the full-cost-recovery figure for the plant being studied (i.e., \$45/MWh). This implies that that cost is adequate for installation of the power plant to be viable in the first place. This is the appropriate auxiliary power penalty because those revenues are what would be forgone due to the auxiliary power impact. The forgone revenues for a very old coal plant backfitted with an FGD system could, arguably, be lower, since the required capacity charges would be lower.

Note that this full-cost methodology is what EPRI included in its Technical Assessment Guide when that guide was only published in paper format.

Considering these generalized conditions, Sargent & Lundy believes that \$30/MWh is fairly representative of the future mix of the power market. We would propose using a range of \$20/MWh to \$45/MWh to demonstrate the sensitivity of the technology selection to this variable. Auxiliary power cost for retrofit applications tend toward the lower end of the range; whereas, for new applications, these costs tend toward the higher end of the range.

Exhibit 5–3 and Exhibit 5–4 represent the fixed and variable O&M costs for new and retrofit applications of the LSFO FGD technology, respectively. Exhibit 5-5 and Exhibit 5-6 represent the fixed and variable O&M costs for new and retrofit applications of the MEL FGD technology, respectively.

5.3 LEVELIZED COSTS

Levelized costs, also referred to as "life cycle costs," take into account the impacts of capital costs and O&M cost during the operation of a plant over the period of analysis. The levelized fixed charge rate (impact due to capital cost) was calculated based on an assumption that a typical customer is a 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 an 8.75% discount rate and 30-year or 20-year life expectancy for new or retrofit facilities, respectively, the levelized fixed charge rates are 14.50% (30-year life) and 15.43% (20-years life). The levelized cost analysis was performed based on current dollars, as most regulated utilities base their analysis on current dollars.

The levelized O&M cost factor takes into account the discount rate, escalation rate, and annuity rate. The levelized O&M cost factors were 1.30 for the 30-year period and 1.22 for the 20-year analysis.

APPENDIX A

AUXILIARY POWER SUMMARY

One of the major components of auxiliary power is additional power required by ID fans to overcome system pressure drop caused by flue gas flow through the absorber. The system pressure drop was estimated as follows in Table A-1:

TABLE A-1
ESTIMATED FGD SYSTEM PRESSURE DROP

	LSFO High-Sulfur	LSFO Low-Sulfur	MEL High-Sulfur	MEL Low-Sulfur
Ductwork to absorber (“w.c.)*	1	1	1	1
Absorber inlet expansion (“w.c.)	0.5	0.5	0.5	0.5
Absorber spray (“w.c.)	3.12	1.92	0.96	0.72
L/G	130	80	40	30
“w.c. loss per 25 L/G (assumed)	0.6	0.6	0.6	0.6
Mist eliminator (“w.c.)	0.8	0.8	0.8	0.8
Absorber outlet contraction (“w.c.)	0.5	0.5	0.5	0.5
Ductwork to stack (“w.c.)	0.2	0.2	0.2	0.2
Additional stack pressure due to wet stack (“w.c.)	2	2	2	2
Margin (“w.c.)	1	1	1	1
Total	**9.1**	**7.9**	**7.0**	**6.7**
***”w.c.=pressure loss in inches of water column**				

The power requirements with flue gas handling, based on the above pressure drop, and the power requirement for other FGD subsystems were calculated as follows in Table A-2:

TABLE A-2
POWER REQUIREMENTS WITH FLUE GAS HANDLING AND OTHER FGD SUBSYSTEMS

	LSFO High-Sulfur	**LSFO Low-Sulfur**	**MEL High-Sulfur**	**MEL Low-Sulfur**
Reagent Preparation (kW)	530	225	330	140
SO_2 Absorption (kW)	3391	2010	990	742
No. operating spray levels	4	3	2	2
#4 (highest) header elevation (ft)	52.5	--	--	--
#3 header elevation (ft)	47.5	47.5	--	--
#2 header elevation (ft)	42.5	42.5	42.5	42.5
#1 (lowest) header elevation (ft)	37.5	37.5	37.5	37.5
Average header elevation (ft)	45	42.5	40	40
Nozzle pressure loss (psi)	10	10	10	10
Specific gravity of pumped slurry	1.09	1.09	1.16	1.16
GPM slurry	196000	120615	58000	43500
Pump efficiency (%)	85.00	85.00	85.00	85.00
Motor efficiency (%)	95	95	95	95
Oxidation Air Compression (kW)	2235	947	2235	947
Flue Gas Handling (kW)	2420	2042	1882	1765
By-product Handling (kW)	420	178	420	178
General (kW)	60	60	60	60
Total (kW)	**9056**	**5462**	**5917**	**3832**

Parasitic load (% of generation)	1.81	1.09	1.18	0.77
Margin (%)	0.2	0.2	0.2	0.2
Total (%)	**2.01**	**1.29**	**1.38**	**0.97**
Parasitic load used in study(%)	**2.00**	**1.30**	**1.40**	**1.00**

The booster fan and slurry pump efficiencies used in the study are 85% and the associated motor drives are 95%, respectively. The LSFO slurry pump curves at 49,000 gpm show a somewhat higher pump efficiency of 87-89%, whereas the MEL pump curves at 29,000 gpm show a lower efficiency of 81-83% in the area of required developed head. It is assumed, for the purposes of this study, that the same efficiency pumps will be available for both services.

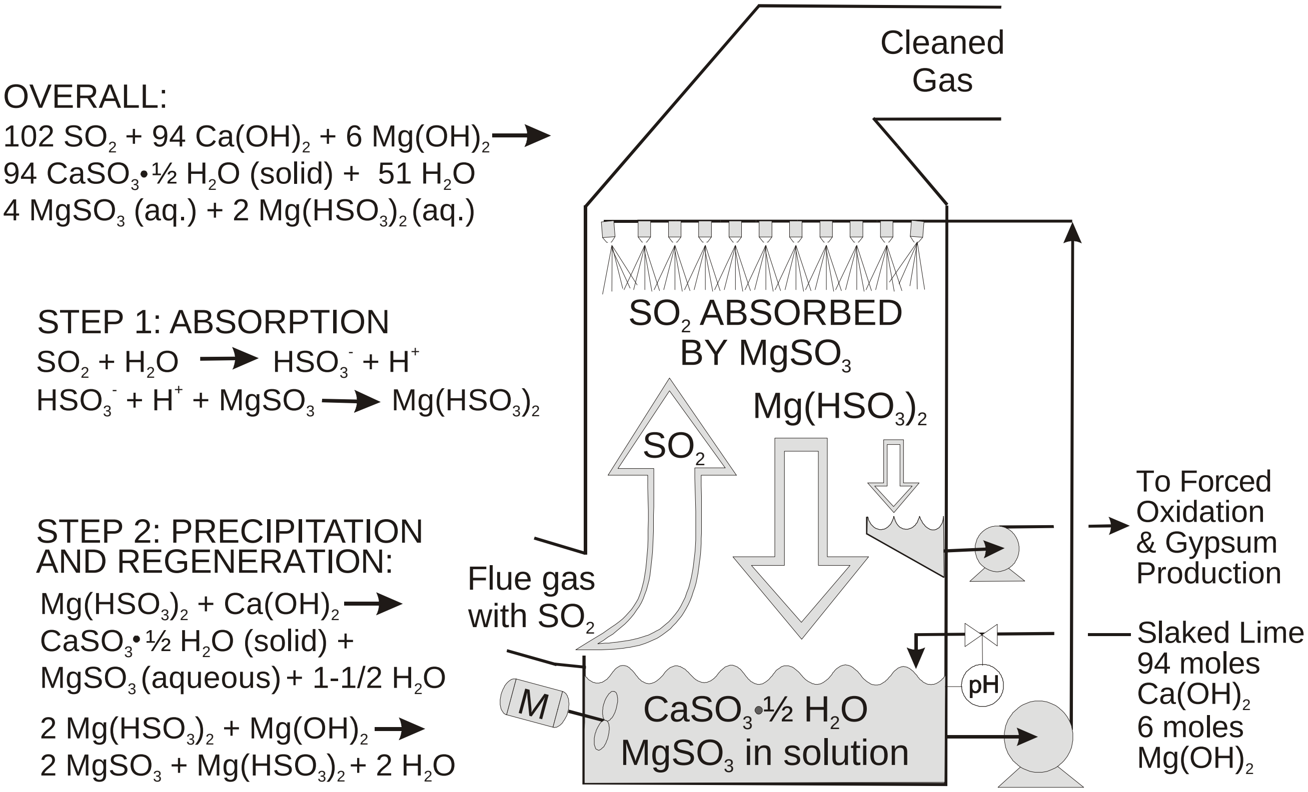
APPENDIX B

DETAILED DESCRIPTION OF MEL PROCESS CHEMISTRY

Steps in Absorption of SO_2 in MEL Process

In the MEL process, slaked lime, containing calcium hydroxide [$Ca(OH)_2$] and a portion of magnesium hydroxide [$Mg(OH)_2$], is used to react with SO_2. Slaked lime is added to the recycle tank of an SO_2 absorber to replenish reagent consumed, as shown in Figure 1.

Figure 1: Reactions in MEL SO_2 Absorber



OVERALL:

$$102\,SO_2 + 94\,Ca(OH)_2 + 6\,Mg(OH)_2 \rightarrow 94\,CaSO_3 \cdot \tfrac{1}{2}H_2O\ (solid) + 51\,H_2O\ 4\,MgSO_3\ (aq.) + 2\,Mg(HSO_3)_2\ (aq.)$$

STEP 1: ABSORPTION

$$SO_2 + H_2O \rightarrow HSO_3^- + H^+$$

$$HSO_3^- + H^+ + MgSO_3 \rightarrow Mg(HSO_3)_2$$

STEP 2: PRECIPITATION AND REGENERATION:

$$Mg(HSO_3)_2 + Ca(OH)_2 \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O\ (solid) + MgSO_3\ (aqueous) + 1\text{-}1/2\ H_2O$$

$$2\,Mg(HSO_3)_2 + Mg(OH)_2 \rightarrow 2\,MgSO_3 + Mg(HSO_3)_2 + 2\,H_2O$$

The overall reaction in the absorber of lime with SO_2, for a magnesium-enhanced lime with 6 moles of magnesium hydroxide and 94 moles of calcium hydroxide, is shown in reaction (1):

(1) $102\ SO_2 + 94\ Ca(OH)_2 + 6\ Mg(OH)_2 \rightarrow 94\ CaSO_3 \bullet \frac{1}{2}\ H_2O$ (solid) $+ 4\ MgSO_3$ (aqueous) $+ 2\ Mg(HSO_3)_2$ (aqueous) $+ 51\ H_2O$

Although reaction (1) shows the overall reaction, SO_2 does not react directly with calcium or magnesium hydroxide. SO_2 capture with lime occurs in two steps. In the first step, SO_2 is absorbed into the liquid phase of slurry recirculated from the recycle tank through banks of spray nozzles. In the second step, SO_2 absorbed into the slurry and in the form of an intermediate reaction product falls into the recycle tank where it reacts with slaked lime.

In the first step, SO_2 is absorbed by recirculated slurry whose liquid phase contains magnesium sulfite ($MgSO_3$). Magnesium sulfite is alkaline with respect to SO_2, which is highly acidic. When SO_2 is absorbed into the liquid, it combines with water to form sulfurous acid, which dissociates into hydrogen and hydrogen sulfite ions (HSO_3^-), shown in reaction (2). $MgSO_3$ reacts with the hydrogen ions, shown in reaction (3), which buffers pH so that slurry pH falls less than 0.5 pH units from the recycle tank set point of 6. This buffering prevents accumulation of SO_2 in the liquid by keeping reaction (1) driven far to the right. By preventing accumulation of SO_2 in the liquid, maximum driving force for mass transfer of SO_2 from flue gas to the liquid is maintained.

(2) $SO_2 + H_2O \rightarrow H^+ + HSO_3^-$

(3) $H^+ + HSO_3^- + MgSO_3 \rightarrow Mg(HSO_3)_2$

In the second step, slurry containing captured SO_2 in the form of $Mg(HSO_3)_2$ falls into the recycle tank. Its lower pH depresses slightly the pH of the contents of the recycle tank, and lime slurry containing both calcium and magnesium hydroxides is added to restore pH to the set point. Calcium hydroxide reacts with most of the $Mg(HSO_3)_2$ to precipitate calcium sulfite hemihydrate ($CaSO_3 \bullet \frac{1}{2}H_2O$) and regenerate $MgSO_3$, according to reaction (4). Magnesium hydroxide in the slaked lime reacts with a smaller portion of

$Mg(HSO_3)_2$ to form additional $MgSO_3$ as shown in reaction (5). A portion of captured SO_2 remains in the form of $Mg(HSO_3)_2$.

(4) $$Mg(HSO_3)_2 + Ca(OH)_2 \rightarrow CaSO_3 \bullet \tfrac{1}{2}H_2O \text{ (solid)} + MgSO_3 \text{ (aqueous)} + 1\tfrac{1}{2}H_2O$$

(5) $$2Mg(HSO_3)_2 + Mg(OH)_2 \rightarrow 2MgSO_3 \text{ (aqueous)} + Mg(HSO_3)_2 \text{ (aqueous)} + 2H_2O$$

A portion of the $MgSO_3$ formed in reactions (4) and (5) is converted to $MgSO_4$ by oxygen present in the flue gas. $MgSO_4$ does not buffer pH in a range effective for SO_2 removal.

However, a mechanism called co-precipitation allows a portion of this $MgSO_4$ to be converted back to $MgSO_3$. When $CaSO_3 \bullet \tfrac{1}{2}H_2O$ precipitates, it tends to include sulfate ($SO_4^=$) ions in its crystal lattice in place of some sulfite ($SO_3^=$) ions. (This co-precipitation occurs even when the scrubber liquor is unsaturated with gypsum). As a result, magnesium ions which were associated with the $SO_4^=$ in $MgSO_4$ become available to pair with $SO_3^=$ and regenerate additional $MgSO_3$. Rewriting equation (4) to include this effect, and assuming that the mole fraction of $CaSO_4$ in the calcium sulfite crystal lattice is 0.2, shows how this additional $MgSO_3$ is formed:

(6) $$Ca(OH)_2 + Mg(HSO_3)_2 + 0.2\ MgSO_4 \rightarrow Ca(SO_3)_{0.8}(SO_4)_{0.2} \bullet \tfrac{1}{2}H_2O \text{ (co-precipitate)} + 1.2\ MgSO_3 + H_2O$$

As a result of this effect, a substantial amount of $MgSO_3$ is produced in addition to that produced by addition of $Mg(OH)_2$ in slaked lime.

Forced Oxidation and Production of Gypsum

A portion of the recirculated absorber slurry that has just contacted the flue gas is collected in a scoop, located above the level of slurry in the recycle tank (Figure 1). Slurry is withdrawn from this location because it has a lower pH which is more suitable for the subsequent oxidation step. Slurry withdrawn from the absorber via the scoop contains calcium sulfite solid and magnesium sulfite and magnesium bisulfite in the liquid; the combined molar rate of withdrawal is equal to the molar absorption rate of SO_2.

The slurry is pumped to an external forced oxidation tank where it is contacted with compressed air. $CaSO_3 \bullet ½H_2O$ is converted to gypsum ($CaSO_4 \bullet 2H_2O$), and $MgSO_3$ is oxidized to $MgSO_4$, according to reactions (7) and (8), respectively. The gypsum precipitates while $MgSO_4$ remains in solution. Liquid containing $MgSO_4$ is returned to the absorber after dewatering the gypsum. A portion of $MgSO_4$ returned to the absorber replenishes $MgSO_3$ as described for reaction (6).

$Mg(HSO_3)_2$ in the liquor is oxidized to form $MgSO_4$ and sulfuric acid (H_2SO_4), shown in reaction (9), which reduces pH to below 5.5 necessary to speed oxidation of $CaSO_3 \bullet ½H_2O$. This “free” sulfuric acid replaces purchased sulfuric acid that would otherwise be required to reduce pH in the external oxidizer. Sulfuric acid increases dissolution rate of calcium sulfite, which increases overall oxidation rate. It also neutralizes small amounts of unreacted lime and calcium carbonate carried over from the absorber, shown in reaction (10). As a result, a negligible amount of unreacted lime remains, which allows overall lime reagent utilization to be more than 99.9%.

$$CaSO_3 \bullet \tfrac{1}{2}H_2O + \tfrac{1}{2}O_2 + 3/2\ H_2O \rightarrow CaSO_4 \bullet 2H_2O \text{ (gypsum)} \tag{7}$$

$$MgSO_3 + \tfrac{1}{2}\ O_2 \rightarrow MgSO_4 \tag{8}$$

$$Mg(HSO_3)_2 + O_2 \rightarrow MgSO_4 + H_2SO_4 \text{ (sulfuric acid)} \tag{9}$$

$$CaO + CaCO_3 + 2H_2SO_4 \rightarrow 2\ CaSO_4 \bullet 2H_2O + CO_2 \tag{10}$$

Exhibit 5-1: Capital Cost for New Units using Appalachian Coals - Wet FGD

	Limestone Forced Oxidation				Mangesium Enhanced Lime			
	Appalachian High Sulfur		Appalachian Low Sulfur		Appalachian High Sulfur		Appalachian Low Sulfur	
Subsystems	**Cost, US$**	**$/kW**	**Cost, US$**	**$/kW**	**Cost, US$**	**$/kW**	**Cost, US$**	**$/kW**
Reagent Feed System	7,373,000	14.7	6,215,000	12.4	5,600,000	11.2	4,824,000	9.6
SO2 Removal System	17,712,000	35.4	15,000,000	30.0	14,290,000	28.6	13,011,000	26.0
Flue Gas System	7,857,000	15.7	7,250,000	14.5	7,769,000	15.5	7,150,000	14.3
By-product Treatment and Handling	7,419,000	14.8	6,018,000	12.0	7,419,000	14.8	6,018,000	12.0
General Support Equipment	2,011,000	4.0	1,710,000	3.4	2,011,000	4.0	1,710,000	3.4
(Inclu. miscellaneous Equipment)								
TOTAL PROCESS CAPITAL (TPC)	42,372,000	85	36,193,000	72	37,089,000	74	32,713,000	65
General Facilities (5% of TPC)	2,119,000	4.2	1,810,000	3.6	1,854,000	3.7	1,636,000	3.3
Engineering and Construction Management	4,237,000	8.5	3,619,000	7.2	3,709,000	7.4	3,271,000	6.5
Project Contingency (15%)	7,309,000	14.6	6,243,000	12.5	6,398,000	12.8	5,643,000	11.3
TOTAL PLANT COST (TPLC)	56,037,000	112.1	47,865,000	95.7	49,050,000	98.1	43,263,000	86.5
Allowance for Funds (AFUDC - 3.2% of TPLC)	1,793,000	3.6	1,532,000	3.1	1,570,000	3.1	1,384,000	2.8
Owner's Cost (5% of TPLC)	2,802,000	6.0	2,393,000	5.0	2,453,000	5.0	2,163,000	4.0
TOTAL PLANT INVESTMENT (TPI)	60,632,000	121.7	51,790,000	103.8	53,073,000	106.2	46,810,000	93.3
Inventory Capital (Spare, 1% of TPI)	606,000	1.2	518,000	1.0	531,000	1.1	468,000	0.9
Initial Chemicals and Commissioning (2% of TPI)	1,213,000	2.4	1,036,000	2.1	1,061,000	2.1	936,000	1.9
Royalties	0	0	0	0	0	0	0	0
TOTAL CAPITAL REQUIREMENT (TCR)	**62,451,000**	**125**	**53,344,000**	**107**	**54,665,000**	**109**	**48,214,000**	**96**

Notes:

1.0 Accuracy of Estimate +-20%
2.0 Labor cost based on normal 5 x 8 shift operation
3.0 Project implementation based on multiple lump sum specialty contracting

Exhibit 5-2: Capital Cost for Retrofit Units using Appalachian Coals - Wet FGD

	Limestone Forced Oxidation				Mangesium Enhanced Lime			
	AppalachianHigh Sulfur		Appalachian Low Sulfur		AppalachianHigh Sulfur		Appalachian Low Sulfur	
Subsystems	**Cost, US$**	**$/kW**	**Cost, US$**	**$/kW**	**Cost, US$**	**$/kW**	**Cost, US$**	**$/kW**
Reagent Feed System	7,588,000	15.2	6,400,000	12.8	6,114,000	12.2	4,970,000	9.9
SO2 Removal System	22,133,000	44.3	18,540,000	37.1	17,100,000	34.2	15,679,000	31.4
Flue Gas System with new stack	15,379,000	30.8	15,370,000	30.7	15,264,000	30.5	15,260,000	30.5
By-product Treatment and Handling	7,900,000	15.8	6,490,000	13.0	7,900,000	15.8	6,490,000	13.0
General Support Equipment	5,800,000	11.6	5,300,000	10.6	5,800,000	11.6	5,300,000	10.6
Including Miscellaneous Equipment (Additional Transformer, Switchgear)								
TOTAL PROCESS CAPITAL (TPC)	58,800,000	118	52,100,000	104	52,178,000	104	47,699,000	95
General Facilities (5% of TPC)	2,940,000	5.9	2,605,000	5.2	2,609,000	5.2	2,385,000	4.8
Engineering and Construction Management	5,880,000	11.8	5,210,000	10.4	5,218,000	10.4	4,770,000	9.5
Project Contingency (15%)	10,143,000	20.3	8,987,000	18.0	9,001,000	18.0	8,228,000	16.5
TOTAL PLANT COST (TPLC)	77,763,000	155.5	68,902,000	137.8	69,006,000	138.0	63,082,000	126.2
Allowance for Funds (AFUDC - 3.2% of TPLC)	2,488,000	5.0	2,204,864	4.4	2,208,000	4.4	2,018,624	4.0
Owner's Cost (5% of TPLC)	3,888,000	8.0	3,445,000	7.0	3,450,000	7.0	3,154,000	6.0
TOTAL PLANT INVESTMENT (TPI)	84,139,000	168.5	74,551,864	149.2	74,664,000	149.4	68,254,624	136.2
Inventory Capital (Spare, same as new)	606,000	1.2	518,000	1.0	531,000	1.1	468,000	0.9
Initial Chemicals and Commissioning (same as new)	1,213,000	2.4	1,036,000	2.1	1,061,000	2.1	936,000	1.9
Royalties	0	0	0	0	0	0	0	0
TOTAL CAPITAL REQUIREMENT (TCR)	**85,958,000**	**172**	**76,105,864**	**152**	**76,256,000**	**153**	**69,658,624**	**139**

Notes:

1.0 Accuracy of Estimate +-20%
2.0 Labor cost based on normal 5 x 8 shift operation
3.0 Booster ID fan/motor and electrical cost is included
additional tranformers and switchgears are also included
4.0 Medium Retrofit Difficulty assumed
5.0 Project implementation based on multiple lump sum specialty contracting

Exhibit 5-3: Fixed and Variable O&M Cost/Levelized Costs (New Units)

LSFO FGD System

Input for O&M Costs

		Appalachian High S	Appalachian Low S
1	Number of Operators (40 hrs/wk)	8	8
2	Operating labor Cost, $/hr	50	50
3	Reagent Purity, %	95	95
4	Reagent Stoichiometry	1.03	1.03
5	Reagent Cost, $/ton	15	15
6	Reagent Requirement, t/h	19.59	8.30
7	SO2 Removal Efficiency, %	98	98
8	SO2 Removed, t/h	11.56	4.90
9	By-product Generated - dry, t/h	32.63	13.83
10	By-product Sale Cost - dry, $/ton	0	0
11	Water Requirement, gpm	660	660
12	Water Cost, $/1000 gal	0.75	0.75
13	Aux. Power Requirement, MW	10.0	6.5
14	Aux. Power Cost, $/MWH	30	30
15	Load Factor, %	80	80

	Appalachian High S	Appalachian Low S
Fixed O&M Costs		
1. Operating Labor Cost ($/yr)	$832,000	$832,000
2. Maintenance Materials Cost ($/yr)	$1,525,000	$1,303,000
3. Maintenance Labor Cost ($/yr)	$1,017,000	$869,000
4. Administrative and Support Labor =	$555,000	$510,000
Total Yearly Fixed O&M Cost =	$3,929,000	$3,514,000
Variable Operating Costs		
1. Reagent Costs =	$2,059,000	$873,000
2. Waste Disposal Cost for FGD System = (Dry basis)	$0	$0
3. Credit for Byproduct =	$0	$0
4. Water Cost=	$208,000	$208,000
5. Additional Power Costs =	$2,102,000	$1,367,000
Total Yearly Variable O&M Cost =	4,369,000	2,448,000
TOTAL YEARLY FIXED AND VARIABLE O&M CO	8,298,000	5,962,000

Levelized Costs

Inputs for Levelized Costs

		Appalachian High S	Appalachian Low S
1	FGD System Life, years	30	30
2	Capital Cost Levelization Factor	14.5	14.5
3	Discount rate, %/yr	8.75	8.75
4	Inflation Rate, %	2.5	2.5
5	Operating Cost Levelization Factor	1.30	1.30
	Total Capital Cost, M$	62.5	53.3
	Levelized capital Cost, MM$/yr	**9.06**	**7.73**
	Levelized O&M Cost, MM$/yr	**10.79**	**7.75**
	Total Levelized Cost, MM$/yr	**19.84**	**15.49**
	Total cents/kW-hr	**0.57**	**0.44**

Exhibit 5-4: Fixed and Variable O&M Cost/Levelized Costs (Retrofit Units)

LSFO FGD System

Input for O&M Costs

		Appalachian High S	Appalachian Low S
1	Number of Operators (40 hrs/wk)	12	12
2	Operating labor Cost, $/hr	50	50
3	Reagent Purity, %	95	95
4	Reagent Stoichiometry	1.03	1.03
5	Reagent Cost, $/ton	15	15
6	Reagent Requirement, t/h	19.59	8.30
7	SO2 Removal Efficiency, %	98	98
8	SO2 Removed, t/h	11.56	4.90
9	By-product Generated - dry, t/h	32.63	13.83
10	By-product Sale Cost - dry, $/ton	0	0
11	Water Requirement, gpm	660	660
12	Water Cost, $/1000 gal	0.75	0.75
13	Aux. Power Requirement, MW	10.0	6.5
14	Aux. Power Cost, $/MWH	30	30
15	Load Factor, %	80	80

	Appalachian High S	Appalachian Low S
Fixed O&M Costs		
1. Operating Labor Cost ($/yr)	$1,248,000	$1,248,000
2. Maintenance Materials Cost ($/yr)	$1,525,000	$1,303,000
3. Maintenance Labor Cost ($/yr)	$1,017,000	$869,000
4. Administrative and Support Labor =	$680,000	$635,000
Total Yearly Fixed O&M Cost =	$4,470,000	$4,055,000
Variable Operating Costs		
1. Reagent Costs =	$2,059,000	$873,000
2. Waste Disposal Cost for FGD System = (Dry basis)	$0	$0
3. Credit for Byproduct =	$0	$0
6. Water Cost=	$208,000	$208,000
7. Additional Power Costs* =	$2,102,000	$1,367,000
Total Yearly Variable O&M Cost =	4,369,000	2,448,000
TOTAL YEARLY FIXED AND VARIABLE O&M CO	8,839,000	6,503,000

Levelized Costs

Inputs for Levelized Costs

		Appalachian High S	Appalachian Low S
1	FGD System Life, years	20	20
2	Capital Cost Levelization Factor	15.43	15.43
3	Discount rate, %/yr	8.75	8.75
4	Inflation Rate, %	2.5	2.5
5	Operating Cost Levelization Factor	1.22	1.22
	Total Capital Cost, M$	86.0	76.1
	Levelized capital Cost, MM$/yr	**13.26**	**11.74**
	Levelized O&M Cost, MM$/yr	**10.78**	**7.93**
	Total Levelized Cost, MM$/yr	**24.05**	**19.68**
	Total cents/kW-hr	**0.69**	**0.56**

Exhibit 5-5: Fixed and Variable O&M Cost/Levelized Costs (New Units)

MEL FGD System

Input for O&M Costs

		Appalachian High S	Appalachian Low S
1	Number of Operators (40 hrs/wk)	8	8
2	Operating labor Cost, $/hr	50	50
3	Reagent Purity, %	94	94
4	Reagent Stoichiometry	1.02	1.02
5	Reagent Cost, $/ton	50	50
6	Reagent Requirement, t/h	10.98	4.65
7	SO2 Removal Efficiency, %	98	98
8	SO2 Removed, t/h	11.56	4.90
9	By-product Generated - dry, t/h	32.63	13.83
10	By-product Sale Cost - dry, $/ton	0	0
11	Water Requirement, gpm	660	660
12	Water Cost, $/1000 gal	0.75	0.75
13	Aux. Power Requirement, MW	7.0	5.0
14	Aux. Power Cost, $/MWH	30	30
15	Load Factor, %	80	80

	Appalachian High S	Appalachian Low S
Fixed O&M Costs		
1. Operating Labor Cost ($/yr)	$832,000	$832,000
2. Maintenance Materials Cost ($/yr)	$1,335,000	$1,178,000
3. Maintenance Labor Cost ($/yr)	$890,000	$785,000
4. Administrative and Support Labor =	$517,000	$485,000
Total Yearly Fixed O&M Cost =	$3,574,000	$3,280,000
Variable Operating Costs		
1. Reagent Costs =	$3,847,000	$1,630,000
2. Waste Disposal Cost for FGD System = (Dry basis)	$0	$0
3. Credit for Byproduct =	$0	$0
6. Water Cost=	$208,000	$208,000
7. Additional Power Costs* =	$1,472,000	$1,051,000
Total Yearly Variable O&M Cost =	5,527,000	2,889,000
TOTAL YEARLY FIXED AND VARIABLE O&M COST	9,101,000	6,169,000

Levelized Costs

Inputs for Levelized Costs

		Appalachian High S	Appalachian Low S
1	FGD System Life, years	30	30
2	Capital Cost Levelization Factor	14.5	14.5
3	Discount rate, %/yr	8.75	8.75
4	Inflation Rate, %	2.5	2.5
5	Operating Cost Levelization Factor	1.3	1.3
	Total Capital Cost, M$	54.7	48.2
	Levelized capital Cost, MM$/yr	**7.93**	**6.99**
	Levelized O&M Cost, MM$/yr	**11.83**	**8.02**
	Total Levelized Cost, MM$/yr	**19.76**	**15.01**
	Total cents/kW-hr	**0.56**	**0.43**

Exhibit 5-6: Fixed and Variable O&M Cost/Levelized Costs (Retrofit Units)

MEL FGD System

Input for O&M Costs

		Appalachian High S	Appalachian Low S
1	Number of Operators (40 hrs/wk)	12	12
2	Operating labor Cost, $/hr	50	50
3	Reagent Purity, %	94	94
4	Reagent Stoichiometry	1.02	1.02
5	Reagent Cost, $/ton	50	50
6	Reagent Requirement, t/h	10.98	4.65
7	SO2 Removal Efficiency, %	98	98
8	SO2 Removed, t/h	11.56	4.90
9	By-product Generated - dry, t/h	32.63	13.83
10	By-product Sale Cost - dry, $/ton	0	0
11	Water Requirement, gpm	660	660
12	Water Cost, $/1000 gal	0.75	0.75
13	Aux. Power Requirement, MW	7.0	5.0
14	Aux. Power Cost, $/MWH	30	30
15	Load Factor, %	80	80

	Appalachian High S	Appalachian Low S
Fixed O&M Costs		
1. Operating Labor Cost ($/yr)	$1,248,000	$1,248,000
2. Maintenance Materials Cost ($/yr)	$1,335,000	$1,178,000
3. Maintenance Labor Cost ($/yr)	$890,000	$785,000
4. Administrative and Support Labor =	$641,000	$610,000
Total Yearly Fixed O&M Cost =	$4,114,000	$3,821,000
Variable Operating Costs		
1. Reagent Costs =	$3,847,000	$1,630,000
2. Waste Disposal Cost for FGD System = (Dry basis)	$0	$0
3. Credit for Byproduct =	$0	$0
6. Water Cost=	$208,000	$208,000
7. Additional Power Costs* =	$1,472,000	$1,051,000
Total Yearly Variable O&M Cost =	5,527,000	2,889,000
TOTAL YEARLY FIXED AND VARIABLE O&M CO	9,641,000	6,710,000

Levelized Costs

Inputs for Levelized Costs

		Appalachian High S	Appalachian Low S
1	FGD System Life, years	20	20
2	Capital Cost Levelization Factor	15.43	15.43
3	Discount rate, %/yr	8.75	8.75
4	Inflation Rate, %	2.5	2.5
5	Operating Cost Levelization Factor	1.22	1.22
	Total Capital Cost, M$	76.3	69.7
	Levelized capital Cost, MM$/yr	**11.77**	**10.75**
	Levelized O&M Cost, MM$/yr	**11.76**	**8.19**
	Total Levelized Cost, MM$/yr	**23.53**	**18.93**
	Total cents/kW-hr	**0.67**	**0.54**