WET FLUE GAS DESULFURIZATION TECHNOLOGY EVALUATION

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PREPARED BY

Sargent & Lundy LLC

55 East Monroe Street
Chicago, IL 60603-5780 USA

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REPORT PREPARED, REVIEWED, AND APPROVED BY SARGENT & LUNDY LLC:

Prepared by:	Raj Gaikwad Technical Advisor	May 2006 Date
Reviewed by:	Steve Katzberger Technical Advisor	May 2006 Date
Approved by:	William DePriest Project Director	May 2006 Date

1. WET FLUE GAS DESULFURIZATION (FGD) DESCRIPTION

Wet FGD technology, which is based on using limestone or lime as a reagent, is the scrubbing process that 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 the flue gas either by a baghouse or by an electrostatic precipitator. The cleaned gas is discharged directly to the stack with no further treatment such as reheat. Wet FGD processes are considered commercially mature technologies and are 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 designs 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 used to create fine droplets to facilitate intimate and uniform contact with the flue gas. The droplets absorb SO_2 from the gas, facilitating the reaction of the SO_2 with the reagent in the slurry. Hydrogen chloride and other soluble species (e.g. HF, Hg++, etc.) present in the flue gas are also absorbed in the spray droplets, causing an accumulation of chloride ions (and other absorbed species) in the process liquid. Some of the water in the spray droplets evaporates, adiabatically cooling the gas from approximately 300°F at the inlet to 125°F-140°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 discharged into the stack.

In most wet FGD systems, the specific SO_2 collection efficiency required is achieved by selecting a combination of design features for the system. For example, one feature is the quantity of liquid sprayed relative to flue gas flow and is referred to as the 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 such design parameters as absorber velocity,

absorber pressure drop and quantity/type of reagent. An important goal of this study will be to quantify the impact that selection of the 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 chloride salts and other undesirable constituents 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 (sulfites) are recycled to the process with a portion removed as wastewater similar to the LSFO process. A variety of the lime based wet FGD system is known as a hi-calcium wet lime system. This system does not rely on the magnesium sulfite inherent in the MEL system as a source of alkalinity. A high calcium wet lime system includes in-situ forced oxidation and operates in a manner similar to an limestone forced oxidation system. The MEL system is typically considered to be the more attractive wet lime system of the two options and is therefore the process evaluated in this study.

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 ($CaSO_3 \bullet ½H_2O$) according to the following simplified reaction:

$$SO_2 + CaCO_3 + ½\ H_2O \Rightarrow\rightarrow CaSO_3 \bullet ½H_2O + CO_2$$

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

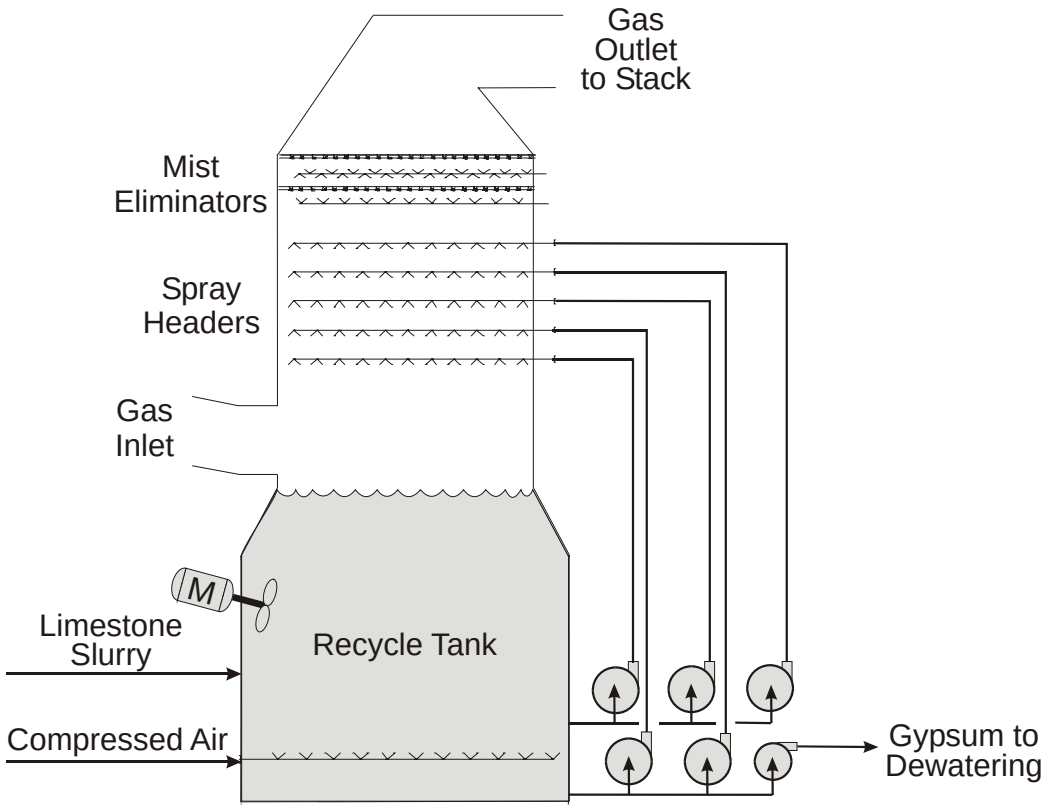
Oxidation air is bubbled through the slurry to convert $CaSO_3 \bullet ½H_2O$ to gypsum ($CaSO_4 \bullet 2H_2O$) according to the following simplified reaction:

$$CaSO_3 \bullet ½H_2O + ½O_2 + 1.5\ H_2O \Rightarrow\rightarrow CaSO_4 \bullet 2H_2O$$

Aeration oxidizes all the calcium sulfite to calcium sulfate and forces precipitation to occur on existing gypsum crystals in the reaction tank. This minimizes the tendency for gypsum to precipitate on surfaces in the absorber, piping, and nozzles by maintaining the gypsum concentration in the liquid phase of the 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 $CaSO_3 \bullet ½H_2O$ 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 the wastewater.

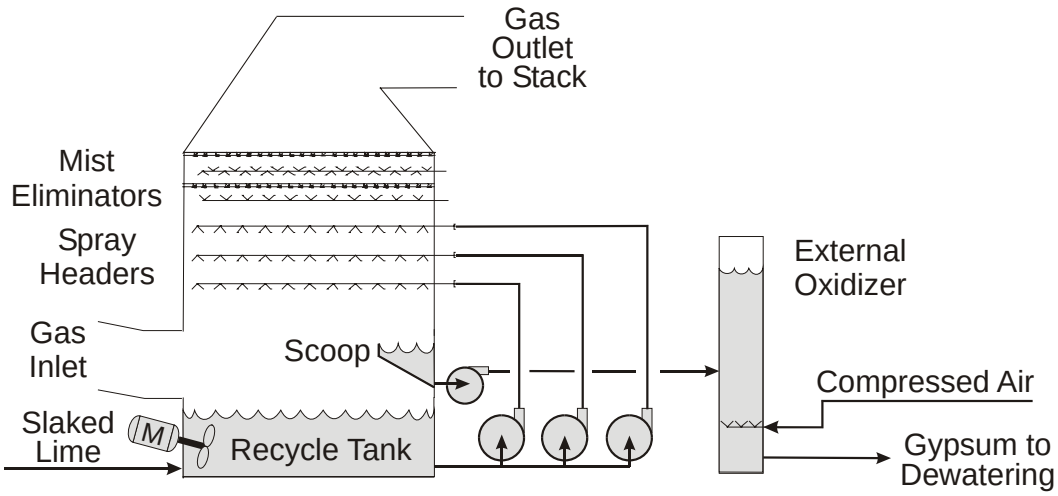
A number of organic acids have been found to be effective. These acids include adipic, glycolic, maleic, acrylic, and formic acids. Dibasic acid (DBA), a by-product from the manufacture of adipic acid, is usually selected instead of other acids because of its lower cost (about $1000 per ton). (DBA is 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 the use of organic acids. Provision is sometimes included in designs to allow future addition of organic acid 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 ($CaSO_3 \bullet ½ H_2O$). 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 enhance SO_2 capture and facilitate the design of a system with lower power consumption and equipment costs. When magnesium sulfite is present in slurry in contact with flue gas, it buffers the pH of the slurry as it absorbs acidic SO_2. This improves absorber performance by increasing solubility of SO_2 in the slurry, which can allow 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 process with a greater capacity to absorb SO_2 per specific quantity of absorbing liquor 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 the 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 the absorber slurry liquid at about 10% of the concentration at which scale would form.

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

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

A portion of the recirculated slurry containing absorbed SO_2 in the form of calcium sulfite and magnesium salts is collected in a scoop from the 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.

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

(4) $$MgSO_3 + \tfrac{1}{2}O_2 \rightarrow MgSO_4$$

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

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 [typically 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. Some LSFO processes claim to be able to accommodate a coarse grind due to a lower operating pH in the absorber. 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 can be used for manufacturing wallboard, as an additive for concrete, or as an agricultural soil conditioner. FGD wastewater containing magnesium sulfate and other soluble species that are undesirable in the gypsum product or in the recirculated scrubber slurry are directed to a wastewater treatment plant similar to that described for the LSFO process. In this study, wastewater treatment requirements are assumed to be the same for the MEL and the LSFO processes. Additional cost is assumed for disposing magnesium sulfate processed through the waste water treatment system. 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 has a number of commercial uses, including injection in the flue gas in the furnace, at the economizer outlet or at the ESP inlet for neutralization of sulfur trioxide (SO_3) in flue gases thereby reducing corrosion in the air heater, and lowering the potential for a visible plume caused by sulfuric acid condensation.

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. Recent contracts for LSFO technology in the U.S. market have included guarantees of 99%. These projects are scheduled for start up in 2008 and beyond. For the LSFO absorber, the higher removal efficiencies are typically attained using a combination of higher L/G's, higher velocity, gas/liquid contactors, or through the use of additives such as DBA in the recirculating slurry . This study employs a design that is not dependent on DBA to achieve the 98% removal requirement.

FGD systems 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 FGD process suppliers, including Alstom, Hitachi, Babcock Power, and Wheelabrator, have shown that an L/G of 130 is adequate to achieve 98% efficiency in a typical open-spray tower design for the high-sulfur coal case (4.72 lb SO_2/MBtu inlet) and an L/G of 80 is appropriate for the low sulfur coal case (2.00 lb SO_2/MBtu inlet). This is further verified by recent guarantees offered by FGD vendors for new and retrofit unit applications on similar sulfur level coals.

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 open spray tower LSFO absorber design. Another important design parameter is that limestone reagent utilization of at least 97% will be necessary to achieve the gypsum quality 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 coals, 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 and 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, but not limited to, the following:

- Alstom
- Babcock & Wilcox
- Babcock Power, Inc.
- Chiyoda
- Marsulex
- Mitsubishi
- Hitachi
- Wheelabrator Air Pollution Control (Siemens)

Although these competing designs vary significantly in their use of the typical mass transfer concepts for achieving high SO_2 removal required of modern FGD systems, recent competition in the U.S. market has shown that the various designs are very similar in life cycle cost (balance of capital and operating costs). Therefore, for purposes of this study, we have chosen a conventional open spray tower design (currently offered by at least five suppliers) as the basis for this work.

2. WET FGD TECHNOLOGY COMPARISON WITH 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 99% have been guaranteed in the last two years.
3. Multiple and commercially viable suppliers offer the technology.
4. Reagents used by the process are plentiful and readily available.
5. Waste gypsum is stable for landfill disposal 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 access to sufficiently sized gypsum market.
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.

5. In most applications, a waste water treatment facility will be required to remove heavy metals from the blowdown stream.

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
Higher 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

* For material balance purposes, all sulfur in coal is assumed to be converted to SO_2 (no SO_3) and there is no sulfur removed in the either the bottom ash or electrostatic precipitator ash.

** The air preheater 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_{gross}	2.10	1.4	1.3	1.0
Reagent	Limestone	Limestone	MEL	MEL
Reagent cost, $/ton	15	15	65	65
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	14.8	14.8	14.8	14.8
Retrofit application	15.7	15.7	15.7	15.7
Discount rate, %	9.0	9.0	9.0	9.0
Inflation rate, %	3.0	3.0	3.0	3.0
Operating cost levelization factor:				
New unit application	1.365	1.365	1.365	1.365
Retrofit application	1.274	1.274	1.274	1.274

*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 24-hour capacity day bin. The limestone day bin and a gravimetric feeder supply the limestone to a 2 x 100% capacity horizontal ball mill system. This will allow 100% availability for the limestone grinding system. 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 16-hour 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 24-hour capacity day bin. The lime day bin and a gravimetric feeder supply the lime to a 2 x 100% capacity horizontal ball mill slaking system. This will allow 100% availability for the slaking system. A modern, conventional horizontal ball mill slaker is used. Two 100% slurry transfer pumps are used to provide high reliability to transfer the slurry to the 16-hour slurry tank. Process makeup water is added to the ball mill slaker 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 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. 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 UNS S32205 steel. The chloride levels will be maintained at less than 15,000 ppm.

The majority of water loss is through evaporation, which is assumed to be the same for both processes (LSFO and MEL). 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 the same water losses in the gypsum by-product. The wastewater 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 UNS S32205 steel or equivalent. The chloride levels will be maintained at less than 15,000 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 employed 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 returns to the recycle tank.

The oxidation tank is sized for 4 hours of solids 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, where a conventional gypsum dewatering system, similar to the LSFO process, is employed.

As an alternative to the conventional dewatering system, 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, including both capital and operating costs, the Oxallizers™ process is approximately 10% more costly than the conventional process considering the assumptions used in this analysis.

Therefore, 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 and 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 FRP lined chimney. The draft requirement of the LSFO FGD system will be accommodated by the booster ID fans. The booster ID fan is sized to provide an additional 10.4" H_2O (9.4" operating) and 9.2" H_2O (8.2" operating) pressure drop through the absorber for high- and low-sulfur coal 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 breeching will be constructed of a carbon steel substrate clad with a 1/16-inch "layer" 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 8.1" H_2O (7.1" operating) and 7.8" H_2O (6.8" operating) for the high- and low-sulfur cases, respectively.

3.2.4 Gypsum Dewatering and Handling System

The gypsum dewatering system will be similar for both 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% solids, is fed to a belt filter for secondary dewatering. Two 100% horizontal 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 vacuum 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 will have to be treated in a wastewater treatment facility to remove heavy metals before discharge.

The 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.

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(24 hours storage)
Limestone crushers
Limestone ball mills (2 x 100% 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 (24 hours storage)
Lime ball mill slakers (2 x 100%% 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
FRP lined stack for retrofit applications
Bypass duct to existing stack with damper
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 flue gas flow from a 1000 MW bituminous coal-fired plant are commercially demonstrated and available from industry suppliers. Therefore, the design basis for this study of a single absorber for a 500 MW facility is well within the current range of commercial applications of FGD technology.

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

SO_2 removal guarantees of up to 99% are available from the system suppliers and have been demonstrated in commercial applications.

4.4 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 the high reactivity of lime and small particle size of slaked lime (3 microns average).

4.5 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 and will typically require the use of a gypsum washing system and subsequent water treatment system to handle soluble impurities not acceptable to wallboard manufacturers.

4.6 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 74% lower (high-sulfur case) than for the LSFO process and the power requirement for the booster ID fan is 27% lower (high-sulfur case) than for the LSFO process. Overall, the MEL process will require approximately 42.8% less auxiliary (% of MW_{gross}) power for high-sulfur coal and approximately 36.9% less auxiliary power auxiliary (% of MW_{gross}) for low-sulfur coal compared to the LSFO process.

4.7 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 at approximately 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 wet stack.

In some cases, retrofit of the FGD equipment will be hampered by the existing site constraints which may result in unique and not well optimized equipment arrangements that would not be the case for new units.

Other than these three key areas, an FGD system itself will be very similar for either the retrofit or the new fit application.

4.8 HG REMOVAL

The 1990 Clean Air Act Amendments required the EPA to study the public health risks, if any, associated with hazardous air pollutant emissions, including mercury, from fossil fuel-fired power plants. The Utility Study was completed in the late 1990's and identified uncertainties with respect to mercury emissions from coal-fired power plants. In order to address uncertainties identified in the mercury reports, EPA issued an information collection request (ICR) requiring all coal-fired power plants to collect and analyze coal samples at their facilities. The ICR also required approximately eighty power plants representing various combinations of fuel types and control technologies to perform mercury speciation tests on stack emissions. This lack of data regarding mercury emissions was a great obstacle to formulating approaches to control mercury emissions from power plants. The ICR was implemented, and a data base of mercury emissions was developed. The existence of the ICR data initiated R/D efforts and eventually revealed some cost effective approaches for reducing mercury emissions.

The ICR data showed that mercury is present in power plant emissions as both solid and vaporous forms. The solid form is a particulate of a size and character that is collected effectively in the existing particulate control equipment (fabric filters and precipitators) typically installed on power plants. Hence, some of the mercury in the coal is already being collected; however, the ICR data also revealed that a majority of the mercury is not in the solid particulate form; rather the mercury is in a vaporous form.

The ICR data indicated that the vaporous form of mercury predominates in power plant exhaust gasses. The vaporous form was categorized as two species, elemental and ionic mercury. Elemental mercury is a molecule that is unreacted, neutral valence, and very stable in the flue gas. The data showed that elemental mercury, Hg0, is only minimally collected in the current power plant emission control equipment. The second category, or species, of mercury is ionic mercury, Hg+2. Ionic mercury is a compound of mercury and other molecules and has a positive valence. The data showed that ionic mercury is collected in some of the existing power plants' pollution control equipment.

The ICR data did not offer a clear correlation between mercury collection and the existing control devices, until researchers identified that power plants firing eastern bituminous coals had greater removal efficiencies and also tended to have greater concentrations of chlorides in their coal analysis. The discovery of this correlation has helped define the problem of controlling power plant mercury emissions. The correlation revealed that chlorides in the coal were associated with higher ionic mercury concentrations. Ionic mercury can be removed with existing pollution control equipment. Coals with higher coal chloride levels exhibited significantly higher mercury removal efficiencies at existing power plants. Further study of ionic mercury compounds supported the correlation, because many of these ionic species were chloride and mercury compounds. The following table illustrates the general tendency found in the ICR data for the speciation of mercury in Eastern Bituminous, PRB Sub-Bituminous, and Lignite coals.

Mercury Speciation in Flue Gas for Various Coal Ranks

Coal Type	Median Hg in Coal	Flue Gas Particulate Bound Hg	Flue Gas Ionic Hg	Flue Gas Elemental Hg	Chloride Content in Coal
Eastern Bituminous	7 lb/TBtu	44.49%	30.25%	25.26%	500-1000 ppm
PRB Sub-Bituminous	5 lb/TBtu	17.56%	17.44%	65.01%	25-200 ppm
Lignite	8 lb/TBtu	17.96%	18.88%	63.16%	25-300 ppm

This correlation explains many of the differing removal efficiencies exhibited at power plants. Plants that fire Powder River Basin (PRB) coals have relatively low amounts of chloride in the coal, have greater concentrations of elemental mercury, and they exhibit low mercury removal efficiencies. Plants firing bituminous coals have higher chloride concentrations in the coal, lower concentrations of elemental mercury, and have greater mercury removal efficiencies.

In support of this correlation, further evaluation of the ICR data identified that baghouses remove considerably more mercury than precipitators, but that the removal depends on the coal type. The removal for baghouses collecting PRB fly ash ranged from 53% to 88%; and 67% to 93% when bituminous coals were fired. It is presumed that baghouses performed better when bituminous coals were fired because they had more ionic mercury in the flue gas, and the baghouses were collecting more of this ionic mercury. Elemental mercury is also being collected in the baghouse but not nearly as efficiently as is the ionic. These results were without any sorbent injection.

Also supporting the correlation, the ICR data indicated wet FGD systems can remove approximately 80% of the inlet mercury from bituminous fired boilers, but only 35% removal on PRB fired boiler. Again, it appears the ionic mercury is collected effectively in the wet FGD system, but the elemental is not collected at all. The PRB has mostly elemental mercury and showed considerable less mercury capture in the wet FGD system.

A second correlation that resulted from evaluation of the ICR data showed greater removal of ionic mercury in the emission reduction equipment that operated at air heater gas outlet temperatures. This was evident when evaluators identified that ionic mercury was not readily collected in hot side precipitators. While the temperature of the flue gas affects mercury collection, this correlation is not nearly as important as the correlation with chlorides. This result has focused investigators efforts at collection at air heater outlet temperatures.

While these two correlations resulted from analysis of the ICR data, the data still reveals inconsistencies and much variation in the removal data. The ICR data dramatically improved the understanding of mercury removal, but there is still much investigation needed to fully understand mercury speciation and mercury capture and how coal chemistry is related to each issue.

The availability of the ICR data resulted in an infusion of grant moneys to evaluate the control of both elemental and ionic mercury emissions from power plants. In the last six years, R/D has identified many emission reduction alternatives and characterized their applicability to coal-fired power plants; however, few of the technologies have yet received long term demonstration in a power plant situation. Thus, there is much still remaining with regard to knowing how technologies will perform in a specific application and the reliability of the systems.

In conclusion, the preponderance of the data on Mercury speciation in the combustion process and the subsequent collection in post combustion emission control technologies indicate that there will be no significant difference between the LSFO and the MEL FGD processes.

5. COSTS ANALYSES

5.1 CAPITAL COSTS

Estimated capital costs for the high- and low-sulfur coal applications for both 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 used during construction (AFUDC)
- Initial inventory and spare parts
- 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. Capital costs are in 2006 dollars and therefore no escalation is 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 costs, contingency, subcontract mark-ups, and project risk profile.

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. For retrofit applications, we have assumed that the plant layout and integration between the existing power generation block and the FGD system is not fully 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. In each case, 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 the FGD system, reagent handling, reagent preparation, gypsum dewatering, gypsum handling, and miscellaneous ball mills/slakers, recirculation slurry pumps, compressor for oxidation air, gypsum dewatering system, and electrical loads and control users typically needed for FGD operations.

Considering both the regulated and unregulated market and new and retrofit applications, Sargent & Lundy believes that $45/MWh is representative of the current power market. 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. Included with each of these Exhibits are the assumptions used for all the components in the fixed and variable O&M costs.

5.3 LEVELIZED COSTS

Levelized costs, also referred to as "life cycle costs," combine the one time capital costs and the yearly O&M cost over the life of the plant. The levelized fixed charge rate (impact due to capital cost) was calculated based on an assumption that a plant owner 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 a 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 15.7% (20-year 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 are 1.365 for a 30-year analysis and 1.274 for a 20-year analysis.

Exhibits 5-7 through 5-10 summarize the levelized costs for the four cases of this study; LSFO on new and retrofit units and MEL on new and retrofit units, respectively. A summary of the levelized costs in ¢/kW-hr for each of the study cases for both the high- and low-sulfur coals is as follows:

Levelized Costs (¢/kW-hr)		
New Units		
	High Sulfur Coal	Low Sulfur Coal
LSFO	0.80	0.66
MEL	0.82	0.64
Retrofit Units		
	High Sulfur Coal	Low Sulfur Coal
LSFO	0.99	0.85
MEL	1.00	0.82

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.)	3.64	2.24	1.12	0.84
L/G	130	80	40	30
“w.c. loss per 25 L/G (assumed)	0.7	0.7	0.7	0.7
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	**9.64**	**8.24**	**7.12**	**6.84**
***”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)	720	305	450	190
SO_2 Absorption (kW)	4,463	2,692	1,161	859
No. operating spray levels	4	3	2	2
#4 (highest) header elevation (ft)	70.0	--	--	--
#3 header elevation (ft)	65.0	65.0	--	--
#2 header elevation (ft)	60.0	60.0	50.0	50.0
#1 (lowest) header elevation (ft)	55.0	55.0	45.0	45.0
Average header elevation (ft)	62.5	60.0	47.5	47.5
Nozzle pressure loss (psi)	10	10	10	10
Specific gravity of pumped slurry	1.09	1.09	1.16	1.16
GPM slurry	197,600	120,615	60,800	45,000
Pump efficiency (%)	85.00	85.00	85.00	85.00
Motor efficiency (%)	95	95	95	95
Oxidation Air Compression (kW)	1,350	572	1,350	572
Flue Gas Handling (kW)	2,592	2,216	1,915	1,839
By-product Handling (kW)	420	178	420	178
General (kW)	60	60	60	60
Total (kW)	**9,440**	**5,865**	**5,355**	**3,699**
Parasitic load (% of generation)	1.89	1.17	1.07	0.74

Margin (%)	0.2	0.2	0.2	0.2
Total (%)	**2.09**	**1.37**	**1.27**	**0.94**
Parasitic load used in study(%)	**2.1**	**1.40**	**1.30**	**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 50,000 gpm show a somewhat higher pump efficiency of 87-89%, whereas the MEL pump curves at 30,000 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.

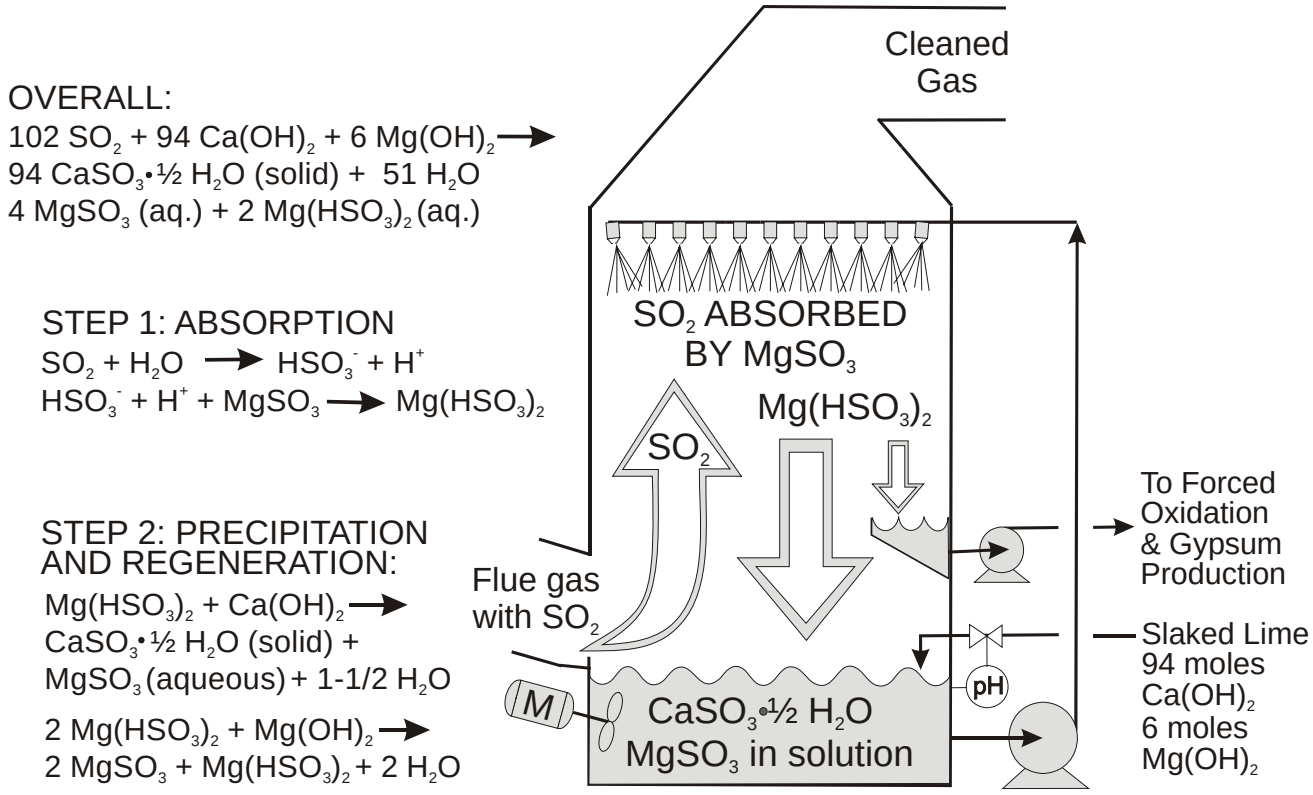
APPENDIX B

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



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) $102\ SO_2 + 94\ Ca(OH)_2 + 6\ Mg(OH)_2 \rightarrow 94\ CaSO_3 \bullet ½\ H_2O$ (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 ($CaSO_3 \bullet ½H_2O$) 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%.

(7) $$CaSO_3 \bullet ½H_2O + ½O_2 + 3/2\ H_2O \rightarrow CaSO_4 \bullet 2H_2O \text{ (gypsum)}$$

(8) $$MgSO_3 + ½\ O_2 \rightarrow MgSO_4$$

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

(10) $$CaO + CaCO_3 + 2H_2SO_4 \rightarrow 2\ CaSO_4 \bullet 2H_2O + CO_2$$

Exhibit 5-1: Capital Cost for 500 MW New Unit using Appalachian Coals - Wet FGD

	Limestone Forced Oxidation				Magnesium 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	12,825,000	25.7	10,432,800	20.9	8,740,000	17.5	7,866,000	15.7
SO2 Removal System	28,475,000	57.0	25,925,000	51.9	25,627,500	51.3	23,332,500	46.7
Flue Gas System	11,500,000	23.0	11,270,000	22.5	11,250,000	22.5	11,025,000	22.1
By-product Treatment and Handling	9,975,000	20.0	9,025,000	18.1	9,975,000	20.0	9,025,000	18.1
General Support Equipment	5,000,000	10.0	4,500,000	9.0	5,000,000	10.0	4,500,000	9.0
Including Miscellaneous Equipment (Additional Transformer, Switchgear, wastewater treatment)								
TOTAL PROCESS CAPITAL (TPC)	67,775,000	136	61,152,800	122	60,592,500	121	55,748,500	111
General Facilities (5% of TPC)	3,389,000	6.8	3,058,000	6.1	3,030,000	6.1	2,787,000	5.6
Engineering and Construction Management	6,778,000	13.6	6,115,000	12.2	6,778,000	13.6	6,115,000	12.2
Project Contingency (15%)	11,691,000	23.4	10,549,000	21.1	10,560,000	21.1	9,698,000	19.4
TOTAL PLANT COST (TPLC)	89,633,000	179.3	80,874,800	161.7	80,960,500	161.9	74,348,500	148.7
Allowance for Funds (AFUDC - 7.0% of TPLC)	6,274,000	12.5	5,661,000	11.3	5,667,000	11.3	5,204,000	10.4
Owner's Cost (5% of TPLC)	4,482,000	9.0	4,044,000	8.0	4,048,000	8.0	3,717,000	7.0
TOTAL PLANT INVESTMENT (TPI)	100,389,000	200.8	90,579,800	181.1	90,675,500	181.3	83,269,500	166.1
Inventory Capital (Spare, 1% of TPI)	1,004,000	2.0	906,000	1.8	907,000	1.8	833,000	1.7
Initial Chemicals and Commissioning (2% of TPI)	2,008,000	4.0	1,812,000	3.6	1,814,000	3.6	1,665,000	3.3
TOTAL CAPITAL REQUIREMENT (TCR)	**103,401,000**	**207**	**93,297,800**	**187**	**93,396,500**	**187**	**85,767,500**	**171**
TOTAL CAPITAL REQUIREMENT (TCR) with BAGHOUSE	**128,401,000**	**257**	**118,297,800**	**237**	**118,396,500**	**237**	**110,767,500**	**222**

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
4.0 Costs are based on year 2006 (No escalation)
5.0 Baghouse is included in the cost for comparison with dry FGD

Exhibit 5-2: Capital Cost for 500 MW Retrofit Unit using Appalachian Coals - Wet FGD

	Limestone Forced Oxidation				Magnesium 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	13,500,000	27.0	11,340,000	22.7	9,200,000	18.4	8,280,000	16.6
SO2 Removal System	33,500,000	67.0	30,500,000	61.0	30,150,000	60.3	27,450,000	54.9
Flue Gas System with new stack	23,000,000	46.0	22,540,000	45.1	22,500,000	45.0	22,050,000	44.1
By-product Treatment and Handling	10,500,000	21.0	9,500,000	19.0	10,500,000	21.0	9,500,000	19.0
General Support Equipment	13,000,000	26.0	12,000,000	24.0	13,000,000	26.0	12,000,000	24.0
Including Miscellaneous Equipment (Additional Transformer, Switchgear, wastewater treatment)								
TOTAL PROCESS CAPITAL (TPC)	93,500,000	187	85,880,000	172	85,350,000	171	79,280,000	159
General Facilities (5% of TPC)	4,675,000	9.4	4,294,000	8.6	4,268,000	8.5	3,964,000	7.9
Engineering and Construction Management	9,350,000	18.7	8,588,000	17.2	9,350,000	18.7	8,588,000	17.2
Project Contingency (15%)	16,129,000	32.3	14,814,000	29.6	14,845,000	29.7	13,775,000	27.6
TOTAL PLANT COST (TPLC)	123,654,000	247.3	113,576,000	227.2	113,813,000	227.6	105,607,000	211.2
Allowance for Funds (AFUDC - 7.0% of TPLC)	8,656,000	17.3	7,950,000	15.9	7,967,000	15.9	7,392,000	14.8
Owner's Cost (5% of TPLC)	6,183,000	12.0	5,679,000	11.0	5,691,000	11.0	5,280,000	11.0
TOTAL PLANT INVESTMENT (TPI)	138,493,000	276.6	127,205,000	254.1	127,471,000	254.6	118,279,000	237.0
Inventory Capital (Spare, same as new)	1,004,000	2.0	906,000	1.8	907,000	1.8	833,000	1.7
Initial Chemicals and Commissioning (same as new)	2,008,000	4.0	1,812,000	3.6	1,814,000	3.6	1,665,000	3.3
TOTAL CAPITAL REQUIREMENT (TCR)	**141,505,000**	**283**	**129,923,000**	**259**	**130,192,000**	**260**	**120,777,000**	**242**

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 Costs are based on year 2006 (No escalation)
6.0 Project implementation based on multiple lump sum speciality contracting

Exhibit 5-3: Fixed and Variable O&M Cost/Levelized Costs (500 MW New Unit)

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	10.5	7.0
14	Aux. Power Cost, $/MWH	45	45
15	Load Factor, %	80	80
16	Waste from Waste water Treatment, t/h	0.4	0.3
17	Waste Disposal cost, $/ton	20.0	20.0

	Appalachian High S	Appalachian Low S
Fixed O&M Costs		
1. Operating Labor Cost ($/yr)	$832,000	$832,000
2. Maintenance Materials Cost ($/yr)	$1,423,000	$1,284,000
3. Maintenance Labor Cost ($/yr)	$949,000	$856,000
4. Administrative and Support Labor ($/yr)	$534,000	$506,000
Total Yearly Fixed O&M Cost =	$3,738,000	$3,478,000
Variable Operating Costs		
1. Reagent Costs ($/yr)	$2,059,000	$873,000
2. Waste Disposal Cost for FGD System ($/yr) (Dry basis)	$49,000	$38,000
3. Credit for Byproduct ($/yr)	$0	$0
4. Water Cost ($/yr)	$208,000	$208,000
5. Auxiliary Power Costs ($/yr)	$3,311,000	$2,208,000
Total Yearly Variable O&M Cost	$5,627,000	$3,327,000
TOTAL YEARLY FIXED AND VARIABLE O&M COSTS	$9,365,000	$6,805,000

Exhibit 5-7: Levelized Costs - LSFO System (500 MW New Unit)

Inputs for Levelized Costs

		Appalachian High S	Appalachian Low S
1	FGD System Life, years	30	30
2	Capital Cost Levelization Factor	14.8	14.8
3	Discount rate, %/yr	9.0	9.0
4	Inflation Rate, %	3.0	3.0
5	Operating Cost Levelization Factor	1.365	1.365

	Appalachian High S	Appalachian Low S
Total Capital Cost, MM$	**103.4**	**93.3**
Total Fixed O&M Cost, $/yr	**3,738,000**	**3,478,000**
Total Variable O&M Cost, $/yr	**5,627,000**	**3,327,000**
Levelized capital Cost, MM$/yr	**15.30**	**13.81**
Levelized O&M Cost, MM$/yr	**12.78**	**9.29**
Total Levelized Cost, MM$/yr	**28.09**	**23.10**
Total cents/kW-hr	**0.80**	**0.66**

Exhibit 5-4: Fixed and Variable O&M Cost/Levelized Costs (500 MW Retrofit Unit)

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	10.5	7.0
14	Aux. Power Cost, $/MWH	45	45
15	Load Factor, %	80	80
16	Waste from Waste water Treatment, t/h	0.4	0.3
17	Waste Disposal cost, $/ton	20.0	20.0

	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)	$1,423,000	$1,284,000
3. Maintenance Labor Cost ($/yr)	$949,000	$856,000
4. Administrative and Support Labor ($/yr)	$659,000	$631,000
Total Yearly Fixed O&M Cost =	$4,279,000	$4,019,000
Variable Operating Costs		
1. Reagent Costs ($/yr)	$2,059,000	$873,000
2. Waste Disposal Cost for FGD System ($/yr) (Dry basis)	$49,000	$38,000
3. Credit for Byproduct ($/yr)	$0	$0
4. Water Cost ($/yr)	$208,000	$208,000
5. Auxiliary Power Costs ($/yr)	$3,311,000	$2,208,000
Total Yearly Variable O&M Cost	$5,627,000	$3,327,000
TOTAL YEARLY FIXED AND VARIABLE O&M COSTS	$9,906,000	$7,346,000

Exhibit 5-8: Levelized Costs - LSFO System (500 MW Retrofit Unit)

Inputs for Levelized Costs

		Appalachian High S	Appalachian Low S
1	FGD System Life, years	20	20
2	Capital Cost Levelization Factor	15.7	15.7
3	Discount rate, %/yr	9.0	9.0
4	Inflation Rate, %	3.0	3.0
5	Operating Cost Levelization Factor	1.274	1.274

	Appalachian High S	Appalachian Low S
Total Capital Cost, MM$	**141.5**	**129.9**
Total Fixed O&M Cost, $/yr	**4,279,000**	**4,019,000**
Total Variable O&M Cost, $/yr	**5,627,000**	**3,327,000**
Levelized capital Cost, MM$/yr	**22.22**	**20.40**
Levelized O&M Cost, MM$/yr	**12.62**	**9.36**
Total Levelized Cost, MM$/yr	**34.84**	**29.76**
Total cents/kW-hr	**0.99**	**0.85**

Exhibit 5-5: Fixed and Variable O&M Cost/Levelized Costs (500 MW New Unit)

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	65	65
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	6.5	5.0
14	Aux. Power Cost, $/MWH	45	45
15	Load Factor, %	80	80
16	Waste from Waste water Treatment, t/h	0.8	0.5
17	Waste Disposal cost, $/ton	20.0	20.0

	Appalachian High S	Appalachian Low S
Fixed O&M Costs		
1. Operating Labor Cost ($/yr)	$832,000	$832,000
2. Maintenance Materials Cost ($/yr)	$1,272,000	$1,171,000
3. Maintenance Labor Cost ($/yr)	$848,000	$780,000
4. Administrative and Support Labor ($/yr)	$504,000	$484,000
Total Yearly Fixed O&M Cost =	$3,456,000	$3,267,000
Variable Operating Costs		
1. Reagent Costs ($/yr)	$5,001,000	$2,119,000
2. Waste Disposal Cost for FGD System ($/yr) (Dry basis)	$111,000	$64,000
3. Credit for Byproduct ($/yr)	$0	$0
4. Water Cost ($/yr)	$208,000	$208,000
5. Auxiliary Power Costs ($/yr)	$2,050,000	$1,577,000
Total Yearly Variable O&M Cost	$7,370,000	$3,968,000
TOTAL YEARLY FIXED AND VARIABLE O&M COSTS	$10,826,000	$7,235,000

Exhibit 5-9: Levelized Costs - MEL System (500 MW New Unit)

Inputs for Levelized Costs

		Appalachian High S	Appalachian Low S
1	FGD System Life, years	30	30
2	Capital Cost Levelization Factor	14.8	14.8
3	Discount rate, %/yr	9.0	9.0
4	Inflation Rate, %	3.0	3.0
5	Operating Cost Levelization Factor	1.365	1.365

	Appalachian High S	Appalachian Low S
Total Capital Cost, MM$	**93.4**	**85.8**
Total Fixed O&M Cost, $/yr	**3,456,000**	**3,267,000**
Total Variable O&M Cost, $/yr	**7,370,000**	**3,968,000**
Levelized capital Cost, MM$/yr	**13.82**	**12.69**
Levelized O&M Cost, MM$/yr	**14.78**	**9.88**
Total Levelized Cost, MM$/yr	**28.60**	**22.57**
Total cents/kW-hr	**0.82**	**0.64**

Exhibit 5-6: Fixed and Variable O&M Cost/Levelized Costs (500 MW Retrofit Unit)

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	65	65
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	6.5	5.0
14	Aux. Power Cost, $/MWH	45	45
15	Load Factor, %	80	80
16	Waste from Waste water Treatment, t/h	0.8	0.5
17	Waste Disposal cost, $/ton	20.0	20.0

	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)	$1,272,000	$1,171,000
3. Maintenance Labor Cost ($/yr)	$848,000	$780,000
4. Administrative and Support Labor ($/yr)	$629,000	$608,000
Total Yearly Fixed O&M Cost =	$3,997,000	$3,807,000
Variable Operating Costs		
1. Reagent Costs ($/yr)	$5,001,000	$2,119,000
2. Waste Disposal Cost for FGD System ($/yr) (Dry basis)	$111,000	$64,000
3. Credit for Byproduct ($/yr)	$0	$0
4. Water Cost ($/yr)	$208,000	$208,000
5. Auxiliary Power Costs ($/yr)	$2,050,000	$1,577,000
Total Yearly Variable O&M Cost	$7,370,000	$3,968,000
TOTAL YEARLY FIXED AND VARIABLE O&M COSTS	$11,367,000	$7,775,000

Exhibit 5-10: Levelized Costs - MEL System (500 MW Retrofit Unit)

Inputs for Levelized Costs

		Appalachian High S	Appalachian Low S
1	FGD System Life, years	20	20
2	Capital Cost Levelization Factor	15.7	15.7
3	Discount rate, %/yr	9.0	9.0
4	Inflation Rate, %	3.0	3.0
5	Operating Cost Levelization Factor	1.274	1.274

	Appalachian High S	Appalachian Low S
Total Capital Cost, MM$	**130.2**	**120.8**
Total Fixed O&M Cost, $/yr	**3,997,000**	**3,807,000**
Total Variable O&M Cost, $/yr	**7,370,000**	**3,968,000**
Levelized capital Cost, MM$/yr	**20.44**	**18.96**
Levelized O&M Cost, MM$/yr	**14.48**	**9.91**
Total Levelized Cost, MM$/yr	**34.92**	**28.87**
Total cents/kW-hr	**1.00**	**0.82**