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**DEVELOPMENT OF THE
INTEGRATED ENVIRONMENTAL CONTROL MODEL**

July - September 1993

Quarterly Progress Report

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MASTER

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INTRODUCTION

The purpose of this contract is to develop and refine the Integrated Environmental Control Model (IECM) created and enhanced by Carnegie Mellon University (CMU) for the U.S. Department of Energy's Pittsburgh Energy Technology Center (DOE/PETC) under contract Numbers DE-FG22-83PC60271 and DE-AC22-87PC79864.

In its current configuration, the IECM provides a capability to model various conventional and advanced processes for controlling air pollutant emissions from coal-fired power plants before, during, or after combustion. The principal purpose of the model is to calculate the performance, emissions, and cost of power plant configurations employing alternative environmental control methods. The model consists of various control technology modules, which may be integrated into a complete utility plant in any desired combination. In contrast to conventional deterministic models, the IECM offers the unique capability to assign probabilistic values to all model input parameters, and to obtain probabilistic outputs in the form of cumulative distribution functions indicating the likelihood of different costs and performance results.

The most recent version of the IECM, implemented on a Macintosh II computer, was delivered to DOE/PETC at the end of the last contract in May 1991. The current contract will continue the model development effort to provide DOE/PETC with improved model capabilities, including new software developments to facilitate model use and new technical capabilities for analysis of environmental control technologies. Integrated environmental control systems involving pre-combustion, combustion, and post-combustion control methods will be considered.

The work in this contract is divided into two phases. Phase I deals with further developing the existing version of the IECM and training PETC personnel on the effective use of the model. Phase II deals with creating new technology modules, linking the IECM with PETC databases, and training PETC personnel on the effective use of the updated model.

The present report summarizes recent progress on the Phase I effort during the period July 1, 1993 through September 30, 1993. This report summarizes our progress to date in refining the existing analytical model of flue gas desulfurization (FGD) technology for SO₂ control. This report presents new performance and cost models for four types of FGD systems to be implemented in the IECM. These models reflect the most recent experience in FGD design, performance and cost.

1.0 Background to FGD Models

In this report we describe the use of systematic sensitivity analysis and multivariate regression for the development of analytical models for the performance and cost of high-performance wet lime/limestone FGD systems and dry lime sprayer systems. Special attention has been paid to FGD systems whose performance with respect to sulfur removal efficiency lies in the 90-98% range. The cost of using additives such as dibasic acid (DBA) as a design option for achieving high performance is also examined. The cost models developed relates the capital costs and the operating and maintenance costs to input variables describing performance parameters and the costs of labor and materials. Systematic sensitivity analysis is used to select and rank the set of input variables and multivariate regression is used for the development of functional relations between the input variables and costs (outputs). This systematic approach along with engineering judgment is used to develop aggregate models for costing high performance FGD systems. These aggregate models are appropriate in size and form for conducting uncertainty analysis using Monte Carlo methods in the IECM.

The report is organized as follows. The next section describes the development of new FGD performance models. Based on a review of the literature some approximate functional relations to describe FGD performance are presented. Section 3 provides an overview of the methodological approach used to derive cost models. Four FGD systems are analyzed: (1) wet limestone with forced oxidation (LSFO), (2) dibasic acid enhanced wet limestone (LS/DBA), (3) a magnesium-enhanced wet lime system, and (4) a spray dryer with lime. The relations for direct process capital cost models are derived for each process area. The operating and maintenance costs of reagent use, power use, waste disposal and other items are also derived. Various statistics to indicate the quality of the model are provided and the response surfaces are provided. This section also examines the effect of different sparing philosophies on the cost of the FGD system. Finally, Section 4 provides a numerical example to illustrate the new models.

2.0 FGD Performance Models

In this section we describe the performance models used to provide key parameter inputs to the cost models described later. We begin with a brief survey of FGD performance literature. Based on the literature, approximate functional relations between performance parameters are developed.

The analysis in this report is geared towards developing updated performance and cost models for four commercial FGD processes:

1) Limestone with Forced Oxidation (LSFO): A limestone slurry is used in an open spray tower with in-situ oxidation to remove SO_2 and form a gypsum sludge. The main advantages as compared to conventional systems is easier dewatering, more economical disposal of scrubber products, and decreased scaling on tower walls.

2) Limestone with Dibasic Acid Additive (LS/DBA): A modification to LSFO where dibasic acid (DBA) is added to act as a buffer/catalyst in the open spray tower. The main advantages are increased SO_2 removal and decreased liquid to gas ratio.

3) Magnesium Enhanced Lime System: A magnesium sulfite and lime slurry (maglime) is used to remove SO_2 and form a precipitate high in calcium sulfite. The high alkalinity of the maglime slurry allows very high SO_2 removal. However, the reagent cost is also higher and solid waste is not easily disposed.

4) Lime Spray Dryer: An atomized spray of a mixture of lime slurry and recycled solids is brought into contact with the hot flue gas. The water in the slurry evaporates leaving dry reaction products and flyash which drops out of the scrubber. A particulate control device such as a baghouse is also used to remove the rest of the dry products from the flue gas before releasing it. The SO_2 removal efficiency is the total of SO_2 removed in the scrubber and the baghouse.

2.1 Wet Limestone FGD Systems

Advanced wet limestone FGD systems are now designed to achieve SO_2 removal efficiencies in excess of 95%. The single loop countercurrent spray tower is the most commonly used device for the removal of SO_2 . The design of spray towers for high efficiency without additives is achieved by using high liquid-to-gas (L/G) ratios and improving gas/liquid contact by spray nozzle design (Bhat, et al., 1993; Rader & Bakke, 1991). On the other hand, organic acids such as dibasic acid or adipic acid are added as buffers to improve performance (Blythe, et al., 1993; Moser & Owens, 1990; Smolenski, et al., 1993; Stevens, et al., 1993). In this section, brief descriptions and approximate relations for high performance LSFO and LS/DBA systems are provided. The approximate relations are gleaned from the literature to reflect US experience with high performance wet limestone FGD systems (Benson, 1993; Dene, et al., 1991; Johnson, 1993; Klingspor, 1993; Laslow, 1993; Moser & Owens, 1990; Noblett Jr., et al., 1990; Noblett Jr., 1993; Rader, 1993; Rader & Bakke, 1991; Smolenski, et al., 1993; Stevens, et al., 1993; Stevens, et al., 1991; Weilert, 1993; Weilert & Ratliff, 1990). The paragraph below provides some background (Corbett, et al., 1977; Noblett Jr., et al., 1990) for the scrubbing process which is useful in understanding the bases for the relations provided later.

The removal of SO₂ from flue gas in a lime/limestone scrubber depends on a gas-liquid-solid mass transfer process. The sulfur dioxide is transferred from the flue gas to the slurry liquid in the scrubber and subsequently precipitated as calcium salts. In this section we provide a very brief discussion of the key factors that affect the SO₂ removal efficiency from a theoretical standpoint. Against this backdrop we examine empirical data which reflects the recent experience with high efficiency scrubbers and develop an approximate response surface parametrized on the key variables.

The basis for the analysis of vapor-liquid mass transfer phenomena is the two-film theory in which the total resistance to mass transfer is expressed as the sum of individual resistances in the vapor and liquid phases near the interface. Based on expressions for molar flux in the gas and liquid films and assuming that the liquid-vapor interface is at equilibrium (Corbett, et al., 1977; Mehta & Rochelle, 1983; Rochelle, 1981; Rochelle, 1981), the overall mass transfer coefficient is written as:

$$K_g = \left(\frac{1}{k_g} + \frac{H}{e \times k_l} \right)^{-1} \quad (1)$$

where k_g (mole/cm²-sec-atm) and k_l (cm/sec) are individual gas and liquid phase mass transfer coefficients, H is a physical constant (atm/mol-liter) from Henry's Law and e is an enhancement factor to account for chemical reactions that permit SO₂ to diffuse through the liquid film as sulfite or bisulfite ions rather than as undissociated SO₂ (Chang and Rochelle 1980). The amount of SO₂ transferred from gas phase per unit time is the integral of the molar flux and the interfacial area for mass transfer. This leads to an expression for SO₂ removal efficiency as follows:

$$\eta_{SO_2} = 1 - \exp\left(-\frac{K_g \times a \times P \times V}{G}\right) \quad (2)$$

where a is the interfacial mass transfer area per unit volume, V the scrubber volume, P the total pressure in scrubber, and G is the molar gas flow rate. The interfacial area is determined by the contactor design, gas distribution in the scrubber and the gas and liquid flow rates (or residence time in the scrubber). V/G is the residence time of flue gas in the scrubber and it depends on superficial velocity and size of the scrubber. The mass transfer coefficient, K_g is influenced by process variables. Any process variable which affects the physical or chemical properties of the two films may affect K_g . The individual mass transfer coefficients k_g and k_l describe the

diffusion rates across two hypothetically stagnant films. In the following paragraphs we discuss the effect of various process parameters on the SO₂ removal efficiency and develop a response surface using the exponential form. All the relations developed in this section are based on a log-linear approximation (as suggested by equation 1) and are developed as follows:

$$\begin{aligned} RTU &= -\ln(1 - \eta_{SO_2}) \\ RTU &= a + \sum_i b_i \times x_i \end{aligned} \quad (3)$$

where a and b_i parameters which represent the linear slope of the effect of process parameters x_i .

L/G is a key variable since it affects SO₂ removal by increasing the available liquid phase alkalinity and the interfacial mass transfer area. Increasing the L/G ratio brings more alkaline materials into the scrubber per mole of SO₂ scrubbed thus decreasing the liquid film mass transfer resistance and improving the SO₂ removal. This effect is magnified by an associated increase in the interfacial mass transfer area which improves efficiency. Industry practice is based on a log-linear relation between η_{SO_2} and L/G ratio (Benson, 1993; Bhat, et al., 1993; Dene, et al., 1991; Klingspor, 1993; Rader, 1993). Typical values for η_{SO_2} and L/G are provided in the table below.

η_{SO_2}	L/G(gpm/Kacfm)
90%	90
95%	130

Based on the values in the table above, an approximate relation¹ between sulfur removal efficiency and L/G is:

$$\eta_{SO_2} = 1 - \exp\{-(0.725 + 0.0175 \times L / G)\} \quad (4)$$

where L/G is measured in gpm/1000 acfm. The effect of lime/limestone scrubbing chemistry on SO₂ removal can be understood by considering the mass transfer resistance from the gas and liquid film separately.

Liquid Film Limited Mass Transfer

¹ All relations developed in this section are based on a wet limestone/lime FGD tower with four spray headers per tower. The default process parameters are as follows: pH=5.3, ϕ =1.03, L/G=90, inlet SO₂=1750 ppm (2.6% S coal), Cl=25,000 ppm (in the slurry) and superficial velocity in tower=10 fps.

Limestone systems are generally at least partially liquid film mass transfer limited. The liquid film mass transfer resistance is a strong function of liquid phase alkalinity. When SO₂ is absorbed by an aqueous scrubbing liquor, sulfurous acid is formed. Increasing the liquid phase or solid phase concentrations of alkaline materials (such as CaCO₃, CaSO₃, Mg⁺⁺) increases the liquid phase alkalinity which in turn improves SO₂ removal efficiency. Increasing the liquid phase alkalinity decreases the Henry's Law constant which in turn decreases the mass transfer resistance $H/(e k_l)$. The enhancement factor depends on the inlet SO₂ concentration and it decreases with increasing SO₂ concentrations thus increasing the mass transfer resistance. The k_l portion of the mass transfer resistance is a function of the hydraulics of the scrubber and depends on gas distribution and the geometry of scrubber.

Increasing the limestone stoichiometry (ϕ) improves liquid phase alkalinity and SO₂ removal. However the effect saturates and little improvement in SO₂ removal is observed over 1.1. Limestone utilization varies inversely with the pH and is a critical factor for the operating costs of SO₂ scrubbing. Most modern scrubbers are designed for high utilization of about 95% or higher. Therefore stoichiometry affects the performance in opposite ways. A detailed quantification of the effect of stoichiometry on pH is not straightforward and needs the development of process simulators (Agrawal & Rochelle, 1993; Noblett Jr., et al., 1990). It is possible to get an approximate idea of how SO₂ removal is affected by pH and in turn by ϕ based on empirical data (Klingspor, 1993; Noblett Jr., et al., 1990; Rader, 1993; Stevens, et al., 1991) as shown below:

η_{SO_2}	pH	ϕ
90%	5.3	1.03
95%	6.0	1.1

Based on these numbers a linear approximation for the effect of stoichiometry can be represented by an additional term in Equation 4 as follows:

$$\eta_{SO_2} = 1 - \exp\{-[0.725 + 0.0175 \times L / G + (10.0 \times \phi - 10.3)]\} \quad (5)$$

Notice that the effect of stoichiometry is represented as an increase in SO₂ removal due to an increase in pH (caused by increasing ϕ) over the default value of 5.3.

Increases in the inlet SO₂ concentrations beyond 1000 ppm decreases the enhancement factor e which in turn increases the mass transfer resistance. The enhancement factor is a function of gas and solution composition (Chang & Rochelle, 1983) which affects the conversion of SO₂ to

bisulfite. Typical values of SO₂ removal for different inlet SO₂ concentrations are provided below based on industry experience (Moser & Owens, 1990; Noblett Jr., et al., 1990; Stevens, et al., 1991).

η_{SO_2}	SO ₂ ppm
90%	2000 (2.8% S)
84%	4000 (4% S)

This can be represented by an additional term in Equation 5 which accounts for the loss of removal efficiency as the inlet SO₂ concentration increases over 2000 ppm.

$$\eta_{SO_2} = 1 - \exp \left\{ \begin{array}{l} -[0.725 + 0.0175 \times L / G + (10.0 \times \phi - 10.3)] \\ - 2.5 \times 10^{-4} (SO_2 - 2000) \end{array} \right\} \quad (6)$$

Similarly calcium chloride accumulation in the slurry leads to a decrease in SO₂ removal. Chloride accumulation suppresses the desirable effects of alkali salts by permitting their accumulation as chloride salts rather than sulfate salts. The effect of accumulated Cl concentrations in the scrubber slurry on the SO₂ removal based on industry experience (Klingspor, 1993; Moser & Owens, 1990; Noblett Jr., et al., 1990; Rader, 1993; Stevens, et al., 1991) is provided below:

η_{SO_2}	Cl ppm
86%	10000
80%	80000

This is represented by a term which represents the loss in removal efficiency as the chloride concentration in slurry increases above 25,000 ppm.

$$\eta_{SO_2} = 1 - \exp \left\{ \begin{array}{l} -[0.725 + 0.0175 \times L / G + (10.0 \times \phi - 10.3)] \\ - 2.5 \times 10^{-4} (SO_2 - 2000) + 5.14 \times 10^{-6} (Cl - 25000) \end{array} \right\} \quad (7)$$

Addition of organic acids such as adipic, glutiric, and succinct acids to the scrubber slurry improves SO₂ removal efficiency. These acids are usually stronger than sulfurous acid and weaker than carbonic acid. This provides a buffering effect. First it assists limestone dissolution by lowering carbonate ion backpressure and hence more alkalinity in form of bicarbonate enters the scrubber in soluble form. Moreover, since sulfurous acid is a stronger acid, the organic acid ion acts as a base in SO₂ sorption. Dibasic acid (a mixture of) has been found to be the most effective organic acid for use as an additive in commercial scale scrubbers. This improves SO₂ removal and typically every 500 ppm addition of organic acids in the slurry improves SO₂

removal that is comparable to increasing the L/G by 30 gpm/Kacfm in a normal pH operating range of 5.3-5.8 (Blythe, et al., 1993; Rader & Bakke, 1991; Smolenski, et al., 1993; Stevens, et al., 1993). This simple approximation can be represented as follows:

$$\eta_{SO_2} = 1 - \exp \left\{ \begin{array}{l} -[0.725 + 0.0175 \times L / G + (10.0 \times \phi - 10.3) \\ - 2.5 \times 10^{-4} (SO_2 - 2000) + 5.14 \times 10^{-6} (Cl - 25000) \\ + 0.00042 \times DBA] \end{array} \right\} \quad (8)$$

where DBA represents the concentration (in ppm) of dibasic acid in the slurry.

Gas Film Limited

Systems with low inlet SO₂ concentrations are gas film limited, since the liquid phase alkalinity is sufficient to neutralize the dissolved SO₂. For gas film limited case the removal efficiency is largely determined by the contactor design and the L/G ratio and is not affected by the liquid phase alkalinity. The SO₂ concentration at which the liquid film resistance becomes large enough to be comparable to the gas film resistance depends on the liquid phase alkalinity and L/G ratio. For typical limestone systems with L/G in the 90-130 gpm/Kacfm interval and pH between 5.3-5.8 the transition is in the 500-1000 ppm range. Therefore for inlet SO₂ concentrations below 1000 ppm the SO₂ removal can be treated as gas film limited and a relation without the effects of liquid phase alkalinity as given in equation 1 can be used for determining SO₂ removal.

2.2 Wet Lime FGD System

Magnesium lime systems are also gas film limited. Mag-lime systems have abundant liquid phase alkalinity since lime has a higher solubility than limestone and magnesium species (usually about 2-4% by weight) provides additional liquid phase alkalinity. Hence the availability of liquid phase alkalinity does not impose any resistance in the absorption of SO₂ and the mass transfer coefficient K_g is based solely on the gas film resistance for inlet SO₂ concentrations from 500-5000 ppm. Typical values for η_{SO_2} and L/G for mag-lime systems are provided in the table below (Benson, 1993; Benson, et al., 1991; Johnson, 1993).

η_{SO_2}	L/G(gpm/Kacfm)
95%	40
98%	56

Based on these values an approximate relation between sulfur removal efficiency and L/G is:

$$\eta_{SO_2} = 1 - \exp\{-(0.2 + 0.07 \times L / G)\} \quad (9)$$

where L/G is measured in gpm/1000 acfm.

2.3 Lime Spray Dryer System

Spray dryer scrubber are sometimes used as an alternative to wet scrubbing since it provides simpler waste disposal and can be installed with lower capital costs. In a typical system, hot flue gas is contacted with finely atomized aqueous solution in the spray dryer. Dry products leave the spray dryer and are removed from the gas in a baghouse. The liquid phase consists of an aqueous slurry of slaked lime. The sulfur dioxide in the flue gas is absorbed into the droplets which simultaneously evaporate on contact with the hot flue gas. Although some moisture still remains in the flue gas solids which later promotes further removal of SO_2 in the fabric filter. The total SO_2 removal is the sum of the removal in the dryer and the baghouse. In this subsection, we develop empirical relations to describe the total SO_2 removal (Blythe, et al., 1985; Blythe, et al., 1991; Brown & Felsvang, 1991).

The main process parameters which affect the SO_2 removal efficiency are: inlet SO_2 concentration, reagent ratio², (ϕ) calcium chloride concentrations in the liquid slurry (Cl), the inlet flue gas temperature (T_{in}), and the approach temperature to the adiabatic saturation temperature of the flue gas at outlet (ΔT) (Brown & Felsvang, 1991; Jozewicz & Rochelle, 1984). Increase in the reagent ratio increases the alkalinity in the liquid phase which improves the SO_2 removal. Effects of inlet SO_2 concentrations are similar to those in wet FGD systems. Low inlet SO_2 concentrations (< 1000 ppm) lead to gas film limited mass transfer and medium to high inlet SO_2 concentration leads to comparable contributions from both gas and liquid film resistances. As a result, we develop different response surfaces to describe the removal efficiency for low and high inlet SO_2 concentrations. Increasing the inlet temperature allows more water to be added to the spray dryer to achieve the same approach temperature. This in turn leads to a higher recycle ratio and increased mass transfer area thus improving the removal efficiency.

In order to understand the effect of the other parameters one has to consider the heat transfer characteristics of the spray dryer. The evaporation time, t , for a given droplet is also the time available for a droplet to absorb SO_2 . As the approach to saturation temperature increases, the

² For spray dryers the reagent ratio is measured as moles of reagent supplied/mole of SO_2 entering the scrubber. Note that for wet FGD systems the reagent ratio is measured as moles of reagent/mole of SO_2 removed.

evaporation time decreases thereby decreasing removal efficiency. The addition of chlorides in the slurry improves the SO₂ removal efficiency of the spray dryer. This is due to the deliquescent properties of chlorides which helps the droplet to retain water and hence increase the evaporation time. Hence chlorides are often used as additives to reduce the lime consumption. We now develop response surfaces to describe the overall SO₂ removal efficiency in spray dryers using data from pilot studies conducted by EPRI (1985, 1991) and Joy/Niro (Svend, et al. 1983). We use the log-linear form shown in Equation 3 to develop these response surfaces.

The reagent ratio is a key process variable and it affects the removal efficiency significantly. Typical value of SO₂ removal for different stoichiometries are presented in the table below:

η_{SO_2}	ϕ
80%	0.9
90%	1.1

These values are representative of both low and high inlet SO₂ concentrations. Based on these numbers the log-linear approximation for the effect of reagent ratio is as follows:

$$\eta_{SO_2} = 1 - \exp\{-(-1.495 + 3.45 \times \phi)\} \quad (10)$$

This relation can be used for reagent ratios in the range of 0.9-1.8. Now we develop the response surface for low and high SO₂ concentrations separately.

Low Inlet SO₂ Concentrations (≤ 1000 ppm)

The data presented in this section was collected against baseline conditions of 1000 ppm inlet SO₂, approach temperature of 20°F, and inlet temperature of 280°F. For low inlet SO₂ concentrations, the inlet temperatures in the range of 280 – 325°F has been observed to have no effect on the removal efficiency. However increasing the approach temperature to saturation from 20 to 40°F results in the requirement of higher stoichiometries for the same removal efficiencies as shown in the table below:

η_{SO_2}	ϕ
80%	1.1
85%	1.3

Averaging the effects of these values an additional term to describe this effect is added to equation 6 as follows:

$$\eta_{SO_2} = 1 - \exp\{-[-1.495 + 3.45 \times \phi - 0.05 \times (\Delta T - 20)]\} \quad (11)$$

As discussed earlier the deliquescent properties of calcium chloride improves the removal efficiency. A presence in the slurry resulting in 0.6% calcium chloride by weight in solids collected in the fabric filter results in higher removal efficiencies for the same stoichiometry as shown below:

η_{SO_2}	ϕ
90%	1.0
95%	1.2

Averaging the effects based on these values an additional term is added in equation 7 to describe the effect of calcium chlorides as shown below:

$$\eta_{SO_2} = 1 - \exp\{-[-1.495 + 3.45 \times \phi - 0.05 \times (\Delta T - 20) + 0.58 \times Cl]\} \quad (12)$$

This is the final form of the response surface for low SO_2 inlet concentrations.

Medium to High SO_2 Concentrations (> 1000 ppm)

The data presented in this section was collected against baseline conditions of 2000 ppm inlet SO_2 , approach temperature of $20^\circ F$, and inlet temperature of $325^\circ F$. For inlet SO_2 concentrations greater than 1000 ppm, the removal efficiency increases with increases in inlet temperatures from $280 - 325^\circ F$. Decreasing the inlet temperature to $280^\circ F$ decreases the removal efficiency to 80% for a given stoichiometry of 1.1. This effect is represented by an additional term to Equations 10 as follows:

$$\eta_{SO_2} = 1 - \exp\{-[-1.495 + 3.45 \times \phi + 0.015 \times (T - 325)]\} \quad (13)$$

The presence of 0.4% by weight of calcium chloride in the fabric filter solids provides an improved efficiency of 90% for a reagent ratio of 1.0. This is described in by an additional term as follows:

$$\eta_{SO_2} = 1 - \exp\{-[-1.495 + 3.45 \times \phi + 0.015 \times (T - 325) + 0.86 \times Cl]\} \quad (14)$$

Increasing the approach temperature to $30^\circ F$ reduces the removal efficiency to 75% for a reagent ratio of 1.05. On the other hand increasing the inlet SO_2 concentration to 3000 ppm

reduces removal efficiency to 80% at a reagent ratio of 1.1. These effects are included as shown below:

$$\eta_{SO_2} = 1 - \exp \left\{ \begin{aligned} & -[-1.495 + 3.45 \times \phi + 0.015 \times (T - 325) + 0.86 \times Cl] \\ & - 0.0007 \times (SO_2 - 2000) + 0.08475 \times (\Delta T - 20) \end{aligned} \right\} \quad (15)$$

This is the final form of the response surface for high SO₂ inlet concentrations.

3.0 FGD Cost Models

The FGD cost models originally developed for the IECM were derived from a detailed FGD computer program developed in the late 1970s and early 1980s by the Tennessee Valley Authority (TVA). The TVA model was subjected to an extensive set of parametric runs, and the results were used to derive statistical "response surface" models which accurately represented the results of the TVA model in terms of a smaller number of key site-specific plant and design parameters. We follow a similar procedure in developing the new models reported here.

Changes in the design, performance and cost of FGD systems in the mid-to-late 1980s left the TVA model outdated as a basis for performance and cost estimation. The TVA modeling team also was disbanded during this period, and no further updates to the model were made. Instead, the most reliable and widely used estimates of FGD cost came from a new modeling effort sponsored by the Electric Power Research Institute (EPRI). This effort led to the development of a computer model called FGDCOST, which has become the new industry standard in the area. FGDCOST, however, does not contain any FGD performance models (all performance parameters must be supplied by the user); hence, the development of the new performance models described earlier.

The new cost models that we developed for the IECM are based on results from FGDCOST, a spreadsheet program model developed for EPRI which provides the most up-to-date cost models for fifteen FGD processes (Keeth, et al., 1991). The cost development and breakdown follow the EPRI economic premises. We have augmented FGDCOST to allow for systematic sensitivity analysis with large numbers of input variables. In order to allow the simultaneous variation of many input parameters, the values of the input parameters are read from a lookup table. The values for input variables are generated from sampling probability distributions. These values are imported into the tables. On each execution, the values of all the inputs are read from each row of the table, the corresponding performance parameters are calculated, and finally the capital and O&M costs are written along the same row. The output columns can be now

exported to other programs for sensitivity analysis and regression modeling. In the following paragraphs we provide a brief outline of the criteria used for designing costs models for the IECM.

As in the current IECM algorithms, cost is broken down in terms of, (1) capital costs, and (2) operating and maintenance costs. These two elements then are combined to calculate the annual revenue requirement and cost of electricity due to the FGD system.

Process Capital Cost. Models of capital cost focus first on the process capital requirements of each section of the FGD system. The bases and items included for each component of the capital estimate is provided in Table 1. The process capital estimate is broken down on an area-by-area basis for each process area of a given FGD system. The costs areas along with the equipment used in each area are provided in Table 2. The equipment size is based on a mass balance performed by FGDCOST, which is similar to the mass balance calculations in IECM. The sulfur content and flue gas flow rate are key variables in costing various FGD processes. Based on the mass balance the size of the various equipment required for each process areas is determined, which is then used to determine costs. The capital costs are also affected by the sparing philosophy for the absorber towers. The sum of the capital cost for all process areas provides the total process capital cost (ctot).

Other Capital Costs. Besides the process capital costs, a series of additional indirect costs is included in the IECM cost models. Following the EPRI format, these are represented as simple parameters whose values are user-specified.

General Facilities: The general facilities costs include roads, office buildings, shops, laboratories etc. A cost factor (f_{gf}) (typically about 5 - 20% of the process capital cost) is used to evaluate this cost.

Engineering & Home Office Fees: An estimate of engineering and home office overhead and fees is included in capital costs estimates. A cost factor (f_{eho}) (typically about 10 -15%) is used to estimate this expense.

Project Contingency: Project contingency costs covers additional equipment and other costs that would arise from a more detailed design. Project contingency factors (f_{cproj}) typically range from 10 - 20% of process capital costs.

Table 1. Nomenclature for EPRI Cost Estimates

ITEMS INCLUDED IN PROCESS CAPITAL COST ESTIMATE

Earthwork
Concrete
Buildings and Structures
Process Equipment
Piping
Electrical
Painting
Instruments and Controls
Insulation
Direct Field Labor
Indirect Field Costs
Payroll Taxes
Insurance, Bonds
Construction Supplies
Temporary Facilities
Construction Equipment
Vendor Fees

FGD SYSTEM CAPITAL COST ESTIMATE COMPONENTS

<u>Capital Investment</u>	<u>\$/kw</u>
Process Capital (includes sales tax)	A
General Facilities	B
Engineering and Home Office Fees	C
Project Contingency	D
Process Contingency	E
Total Plant Cost, TPC	= A+B+C+D+E
Total Cash Expended, TCE	TPC x Adjust factor*
AFDC (Allowance for Funds During Construction)	F
Total Plant Investment, TPI	= TCE + F
Royalty Allowance	G
Preproduction Costs	H
Inventory Capital	I
Initial Catalyst and Chemicals	J
Land	K
Total Capital Requirement, TCR (Capital Investment Jan., 1990, Startup Jan., 1990)	<u>TPI+G+H+I+J+K</u>

*Adjustment Factor for TCE per table 3-4 of EPRI TAG (P6587-L)

Table 2. Cost Areas for Process Capital Breakdown (Source: EPRI)

<u>Area</u>	<u>Description</u>
10	Reagent Feed System
20	SO ₂ Removal System
30	Flue Gas System
40	Regeneration System
50	By-product System
60	Solids Handling System
70	General Support Area
80	Miscellaneous Equipment

- 10: Reagent Feed System - all equipment required for storage, handling and preparation of raw materials, reagents, and additives used in each process.
- 20: SO₂ Removal System - equipment required for SO₂ scrubbing, such as the absorption tower, recirculation pumps and other associated equipment.
- 30: Flue Gas System - duct work and fans required for flue gas distribution to the SO₂ scrubbing system, plus gas reheat as required.
- 40: Regeneration System - specific to regenerable reagent systems, equipment used to regenerate spent absorbent for return to the process, plus any preconditioning system for SO₂ or H₂S off gas.
- 50: By-product System - production equipment for salable process by-products and storage facilities for the final products.
- 60: Solids Handling System - equipment required for fixation, treatment, and transportation of all sludge/dry solids materials produced by each scrubbing process.
- 70: General Support Area - additional equipment required to support FGD system operation such as makeup water and instrument air.
- 80: Miscellaneous Equipment - This area will include plant modifications necessitated by the addition of the FGD system. Also included are costs for electrical equipment tie-ins and other associated systems.

Process Contingency: A process contingency is applied to a new technology in an attempt to quantify the design uncertainty and the cost of a commercial scale system. The contingency factor (f_{cproc}) can range from 2-50% of the process capital costs.

Total Plant Cost: The total plant cost (tpc) is the sum of process capital, general facilities, engineering and home office fees, and contingencies.

Total Cash Expended: The total cash expended is an estimate of the cash expected to be spent during the construction duration of the FGD system. This estimate accounts for the escalation of costs up to the date of expenditure using standard adjustment factors.

Allowance for Funds During Construction (AFUDC): The schedule for engineering procurement and construction is assumed to be two years for installation. For an escalation of 5% a year, the total plant costs is multiplied by 0.0548 to calculate the allowance for interest expenses during a two year construction period. Other AFUDC costs are calculated by the model.

Total Plant Investment: The total plant investment is the sum of the total cash expended and the allowance for interest during construction.

Royalty Allowance: Royalties paid are nominally 0.5% of the process capital.

Preproduction Costs: Preproduction costs are intended to cover operator training, equipment checkout, major changes in plant equipment, extra maintenance, and inefficient use of materials during plant startup. It is a sum of one-month fixed operating costs, one-month variable operating costs at full capacity and 2% of the total plant investment.

Inventory Capital: The inventory costs includes the expense of raw materials and other consumables based on a 100% capacity operation for 60 days.

Land: FGD system land requirements include the plant site area and disposal area. Land has not been included as a line item.

Operating & Maintenance Costs

Operating costs for FGD systems are separated into fixed and variable operating costs. Fixed costs include operating and maintenance labor, maintenance materials and administrative/support labor. Various factors based on EPRI premises are used to estimate these costs based on capital cost estimate. Variable operating costs include consumables such as fuel, water, power, chemicals, and solids disposal. Table 3 provides the operating cost criteria used. The fixed and variable O&M costs depend on the feed rate of reagent, sludge disposal costs, power for pumping, labor and other operations. Table 4 provides the default process design criteria for FGDCOST. The default used in our models are outlined in Table 5.

3.1 The Methodological Approach

Response surface construction is the central objective of the sensitivity analysis techniques presented in this section. Sensitivity analysis seeks to identify the set of input variables that have significant effect on the model output and to rank order them according to the magnitude of their effects. Once the relevant variables are identified regression methods can be used to develop adequate response surface replacements for the models. Subsequently probability distributions on the input variables can be used in conjunction with the response surface to derive the probability distributions of the model outputs. In this section, we briefly describe the use of Latin hypercube sampling (LHS) (Iman, et al., 1980) along with partial rank correlation coefficients (PRCC) to examine the influence of individual variables on the model output (Iman, et al., 1981a; Iman, et al., 1981b). Finally the use of regression analysis with log-linear transformations for developing response surfaces is described (Ang & Tang, 1975; Draper & Smith, 1966; Neter, et al., 1983).

3.1.1 Sampling and Ranking Methods

In order to perform sensitivity analysis it is necessary to obtain model output for various values of input variables. Latin hypercube sampling is used to select values for input variables. The LHS technique to select n different values from k different input variables operates in the following manner. For each variable, the range is divided into n nonoverlapping intervals of equal probability. One value from each interval is selected at random with respect to the probability density in the interval. The n values for variable x_1 are paired at random with the n values of x_2 and these n pairs are paired at random with n values of x_3 and so on to generate n k -tuples. This is the Latin hypercube sample used as input to the model. The model is then exercised at each of these n input values to generate n l -tuples for l output variables of interest. This data set of n k -tuples of input values and n l -tuple of output values forms the basis for the rest of the analysis.

Table 3

<u>Fixed Operating Costs</u>	<u>Unit</u>	<u>Rate</u>
Operating Labor	M ² ft-hrs	\$20.00 (Jan. 1990)
Maintenance Labor - Slurry Handling	\$/yr	3.2% of Process Capital
- Liquid Handling	\$/yr	1.2% of Process Capital
Maintenance Material - Slurry Handling	\$/yr	4.8% of Process Capital
- Liquid Handling	\$/yr	1.8% of Process Capital
Administrative & Support Labor	\$/yr	30% of Operating and Maintenance Labor

Variable Operating Costs	Unit	Jan. 1990 Cost/Unit	Additional Freight \$/t	30-Yr Level Factor	30-Yr Levelized Cost/Unit
Fuel Oil (No. 6)	gal	\$ 0.41	Incl	1.613	\$ 0.66
Raw water	1000 gal	\$ 0.60	Incl	1.613	\$ 0.97
Cooling water	1000 gal	\$ 0.16	Incl	1.613	\$ 0.26
Power	kwh	50 mills	Incl	1.668	83 mills
Methane	1000 ft ³	\$ 3.00	Incl	1.668	\$ 5.00
Lime	ton	\$ 55.00	Incl	1.613	\$ 88.72
Limestone	ton	\$ 15.00	Incl	1.613	\$ 24.20
Soda Ash	ton	\$ 93.00	\$ 43.00	1.613	\$ 219.37
Magnesia	ton	\$ 232.00	\$135.00	1.613	\$ 591.97
Ammonia	ton	\$ 145.00	\$ 5.50	1.613	\$ 242.76
Sulfur Emulsion	ton	\$ 220.00	Incl	1.613	\$ 354.86
Dibasic Acid	ton	\$ 360.00	Incl	1.613	\$ 580.68
Formic Acid	ton	\$ 800.00	\$20 (est.)	1.613	\$1322.66
Allied Catalyst	ton	\$2500.00	\$45.00	1.613	\$4105.09
Claus Catalyst	ton	\$1000.00	Incl	1.613	\$1613.00
Disposal Charges,					
Dry Solids (lined)	ton (dry)	\$ 9.29	Incl	1.613	\$ 14.98
Fly ash (unlined)	ton (dry)	\$ 8.00	Incl	1.613	\$ 12.90
Fly ash with Nahcolite	ton (dry)	\$ 11.14	Incl	1.613	\$ 17.97
Sludge (trucked to lined landfill)	ton (dry)	\$ 9.25	Incl	1.613	\$ 14.92
Sludge (trucked to unlined landfill)	ton (dry)	\$ 8.15	Incl	1.613	\$ 13.15
Sludge (ponded)	ton (dry)	\$ 6.00	Incl	1.613	\$ 9.68
Gypsum (pumped & stacked)	ton (dry)	\$ 4.75	Incl	1.613	\$ 7.66
Condensate	1000 lb.	\$ 0.77	Incl	1.613	\$ 1.24
Steam					
0-70 psia	1000 lb.	\$ 2.85	Incl	1.668	\$ 4.75
70-250 psia	1000 lb.	\$ 3.50	Incl	1.668	\$ 5.84
250-400 psia	1000 lb.	\$ 5.30	Incl	1.668	\$ 8.84
By-product Credit					
Sulfur	Long ton	\$ 90.00	Incl	1.613	\$ 145.17
Sulfuric Acid	ton	\$ 50.00	Incl	1.613	\$ 80.65
Liquid Sulfur Dioxide	ton	\$ 230.00	Incl	1.613	\$ 371.00
Gypsum	ton	\$ 2.00	Incl	1.613	\$ 3.23

Note: 1000 gal = 3.785 m³, t (short) = 0.9072 t (SI),
1000 ft³ = 28.32 m³, 1000 lb = 453.6 Kg
Assumed 100 mile shipping distance for all chemical reagents except where noted.

Table 4
EPRI Design Basis

**LIMESTONE SYSTEM WITH FORCED OXIDATION
PROCESS DESIGN CRITERIA***

Flue Gas Handling Area Criteria

Flue gas flow rate	1,020,000 acfm @ 277°F
Absorber pressure drop	6 in. H ₂ O
Total system pressure drop	10 in. H ₂ O
ESP fly ash removal efficiency	99.5%
Spray tower outlet gas temperature	127°F

SO₂ Removal Area Criteria

SO ₂ removal (2.6% S Coal)	90%
Scrubber modules	3 @ 50% capacity each (2 operating, 1 spare)
Scrubber design	Spray tower
L/G ratio	96 gpm/1000 saturated acfm
Scrubber slurry solids concentration	15%
CaCO ₃ reagent feed ratio	$\frac{1.10 \text{ lb mole CaCO}_3}{1 \text{ lb mole SO}_2 \text{ removed}}$
Absorber tower superficial velocity	10 fps
Tower gas residence time	10 seconds
Reaction mix tank retention time	6 min
Oxygen feed rate	$\frac{2 \text{ lb mole O}_2}{1 \text{ lb mole SO}_2 \text{ removed}}$
Percent oxidation	100%
Scrubber slurry pH	5.3

Reagent Feed Area Criteria

Total limestone storage	60 days
Limestone day bin storage	30 hrs
Limestone slurry tank storage	8 hrs

Solids Handling Area Criteria

Thickener underflow % solids	45%
Thickener overflow tank capacity	30 min
Thickener size	2 ft ² /tpd dry solids
Vacuum filter product % solids	85%
Vacuum filter size	150 lb/hr/ft ²

General Support Area Criteria

Makeup water tank capacity	30 min
Seal water tank - capacity	30 min
- water requirement	10 gpm/spray pump

*For one 300 MW unit.

Table 5: Description of Input Variables

Variable name	Units	Description	Default Value
gas	Kacfm	flue gas flow rate	971
MW	10 ⁶ W	Net plant capacity	300
L/G	gpm/Kacfm	liquid to gas ratio	90/55/30 ^a
%S	%	% sulfur in coal by weight	2.6% 1750 ppm
stoichiometry (ϕ)		moles of Ca per mole of SO ₂ removed	1.03/1.02/ 1.1 ^b
labor	\$/hr	FGD operating labor rate	20
reag	\$/ton	reagent (CaCO ₃ /CaO) cost	15/55 ^d
solid	\$/ton	sludge disposal cost	8.15
stack	\$/ton	sludge disposal - stacking cost	4.75
power	mills/kWh	power cost	50
steam	\$/1000 lbs	steam cost	3.5
waterf	\$/1000 gallons	fresh water cost	0.6
waterc	\$/1000 gallons	cooling water cost	0.16
DBA _{ppm} [*]	ppm	scrubber slurry DBA concentration	1500
DBA _{feed} [*]	lbs DBA/ton SO ₂ removed	DBA feed rate	20
DBA _{cost} ^c	\$/ton	DBA additive cost	360
ΔP	inches of H ₂ O	total system pressure drop	10
ΔT	°F	temperature increase of scrubbed flue gas due to reheat	25

^a the default values are: 90 gpm/Kacfm for LSFO and 55 gpm/Kacfm for LSDBA and 30 for Mg-lime System for a 90% removal efficiency.

^b the stoichiometry for limestone systems is 1.03, for lime systems is 1.02, and for lime spray dryer is 1.1 (note that for lime spray dryer the reported stoichiometry is moles of Ca per mole of inlet SO₂).

^c these input variables for LS/DBA only

^d limestone cost is 15 and lime cost is 55

The importance of an input variable is derived by assessing its influence on the output variable. For linear relationships this can be assessed by the use of partial correlation coefficients (PCC) which measure the correlation between an input and output variable when the effects of all the other variables is removed. For nonlinear relations it becomes more difficult to assess the importance of individual variables. However, if the model output is a monotonic function of the input then it is possible to linearize the relationship by using rank transformations on the input and output values. A rank transformation involves replacing each value of a variable by its rank. Now the partial correlation coefficient on the rank transformed variables (PRCC) can be used to assess the importance of the input variables.

3.1.2 Regression Analysis

Once the important input variables are chosen, multivariate linear regression approaches can be used with the data set to construct response surfaces. If k input variables are chosen as relevant, a linear regression model is written as:

$$y_i = a + b_1x_1 + b_2x_2 + \dots + b_kx_k + \varepsilon \quad (16)$$

where ε is an error term which represents the variance in y_i unexplained by the model. The linear fit is obtained by selecting a, b_i so as to minimize the sum of squares between the model predictions and the data for y_i . Often the relationships between input and output variables is nonlinear. In this case it is often possible to use some data transformation to linearize the relationship. A commonly used transformation for exponential nonlinearities is the logarithm transform. Therefore an exponential relationship such as:

$$y_i = a \times x_1^{b_1} \times x_2^{b_2} \dots \times x_k^{b_k} \quad (17)$$

is transformed to:

$$\ln(y_i) = \ln(a) + b_1 \ln(x_1) + \dots + b_k \ln(x_k) \quad (18)$$

which has a linear relationship.

Common statistical tests are used to determine the adequacy of the regression model. The most common measure is the coefficient of multiple correlation (R^2) which is a ratio of the variation explained by the regression to the total variation in y_i . Values of R^2 near 1 represent good fits.

R^2 however is not a sufficient measure of the goodness of fit of a model. The t -statistic is used to test if the partial regression coefficients a, b_i are significantly different from zero. Moreover an F -statistic can be used to check if any of the partial correlation coefficients are different from zero. This statistic provides an idea of the lack of fit of a regression model as a whole. However, for large R^2 values this test is not informative. Standard error of a regression model provide an estimate of the variance in the residual errors for y_i . This can be used to compare across regression models (in terms of functional forms) and to conduct the Kolmogorov-Smirnov test to evaluate if the errors ϵ are normally distributed. The normality assumption underlies the most of the statistical tests used. Most statistical packages provide utilities for regression analysis and the statistical tests outlined in this section. The regression analysis for this work was done using *Splus* on the Andrew network (Becker, et al., 1988).

3.1.3 Software Employed

In the following sections we present the results of our sensitivity analysis and provide response surface algorithms for capital costs and O&M costs based on results from FGDCOST. A set of input variables is chosen and Latin hypercube sampling is used to generate a sample of 100 values for each variable using a uniform distribution for the given range. A Fortran77 package was used to generate the samples (Iman 1981a). These samples are then imported into the sensitivity analysis module in FGDCOST and the model exercised to generate corresponding output values for capital and O&M costs. The data set consisting of the input samples and the corresponding output samples is used for response surface generation. PRCC, a Fortran77 package, is used to compute the partial rank correlation coefficients between the input and output variables. Note that linear relations are monotonic, hence a rank transformation does not affect the order of importance of the input variables. Using these coefficients the relevant variables for regression modeling are gleaned. *Splus* is used to generate regression models for capital and O&M costs. Table 5 provides a list of the input variables used for various FGD systems.

3.2 Wet Limestone with Forced Oxidation

A PRCC analysis of the data set for LSFO was performed. Based on this analysis the input variables are ranked in order of importance in the table below.

Table 6: PRCC Analysis for Limestone Forced Oxidation

Variables	Range	Process	Capital	Fixed	Costs	Variable Costs	
		PRCC	Rank	PRCC	Rank	PRCC	Rank
flue gas(Kacfm) (MW equivalent)	625-3350 (300-1000)	-0.992	1	-0.989	1	-0.02	11

L/G(gal/Kacfm)	90-130	0.336	4	0.233	5	-0.036	9
SO ₂ (inlet SO ₂ ppm)	0.5-5.0	0.806 [†]	3	0.588	4	0.974	1
stoichiometry	1.05-1.15	0.856	2	0.787	3	0.542	3
labor rate(\$/hr)	15-25	0.018	12	0.868	2	-0.012	12
reagent cost(\$/ton)	10-20	-0.066	9	-0.088	7	0.58	2
solid disposal	5-10	-0.161	6	-0.21	6	0.47	5
waste(\$/ton) stack	3-8	0.121	7	-0.022	10	0.482	4
powercost(mills/kwh)	40-60	-0.03	11	0.004	12	0.1	7
steam cost(\$/Klbs)	3-4	0.10	8	-0.059	8	0.268	6
water fresh(\$/Klbs)	0.5-0.75	-0.059	10	-0.007	11	0.084	8
water cool(\$/Klbs)	0.1-0.2	-0.182	5	-0.042	9	-0.027	10

Now we present the development of regression models using the variables identified from the PRCC analysis.

3.2.1 Capital Costs

The capital cost components for a FGD system is calculated based on the EPRI economic premises for electric power generating plants. The total plant cost is the sum of the process capital, general facilities capital, engineering and home office overhead and project and process contingencies (refer to table 1 for details). The process capital cost is the total constructed cost of all on-site processing and generating units including direct and indirect constructions costs. In the paragraphs below, the process capital cost models are developed on an area by area basis. The capital costs related to the other terms above are calculated by using cost factors on the process capital cost. This is also demonstrated below.

The main variables which effect the process capital costs are flue gas flow rate (Kacfm), SO₂ the inlet SO₂ concentration in flue gas (ppm), L/G ratio (gpm/Kacfm), and stoichiometry. The capital

costs are normalized by plant size and expressed in \$/kW units for all process areas¹. It is known from prior engineering experience that the relation between the capital costs and the input variables is exponential, hence we use logarithmic transformations on the variables and conduct a linear regression of the transformed variables. The results of a regression analysis using *Splus* is provided in the Appendix. The highlights are presented here.

Reagent Feed System (Process Area 10): The main input variables are flue gas flow rate (gas), SO₂ (inlet SO₂ in ppm), the stoichiometric ratio (ϕ), and L/G. The regression³ model is:

$$\ln(C_{10}) = 3.050657 - 0.7902577 \times \ln(gas) + 0.2203442 \times \ln(SO_2) + 1.664678 \times \ln(\phi) \quad (19)$$

$$R^2 = 0.9953, (n=93).$$

SO₂ Removal System (Process Area 20): The main input variables are flue gas flow rate (gas), SO₂, L/G and the stoichiometric ratio (ϕ). The regression model is :

$$\ln(C_{20}) = 2.670582 - 0.4402019 \times \ln(gas) + 0.01914421 \times \ln(SO_2) + 0.1375112 \times \ln(L / G) + 1.362315 \times \ln(\phi) \quad (20)$$

$$\text{with } R^2 = 0.9975 \text{ and } n=93.$$

Flue Gas System (Process Area 30): The main input variables are flue gas flow rate (gas), SO₂, and the stoichiometric ratio (ϕ). The regression model is:

$$\ln(C_{30}) = 2.422378 - 0.3985948 \times \ln(gas) + 1.372434 \times \ln(\phi) \quad (21)$$

$$\text{with } R^2 = 0.9889, (n=93).$$

Solids Handling System (Process Area 60): The main variables are flue gas flow rate (gas), SO₂, the stoichiometric ratio (ϕ), and L/G. The regression model is:

¹ The capital costs, fixed O&M costs and the variable operating costs are normalized in terms of \$/kWh where the normalization is done against the net power rating in MW. In FGDCOST the net power rating is calculated as a fixed factor (~0.9585) of the gross power rating.

³ Note that all logarithms in the cost model are to the base 10.

$$\ln(C_{60}) = 1.32039 - 0.6194293 \times \ln(gas) + 0.3790324 \times \ln(SO_2) + 1.655228 \times \ln(\phi) \quad (22)$$

with $R^2 = 0.9824$ (n=93).

General Support Area (Process Area 70): The main variables are flue gas flow rate (gas), L/G, and the stoichiometric ratio (ϕ). The regression model is:

$$\ln(C_{70}) = 2.717534 - 0.8992044 \times \ln(gas) + 1.433288 \times \ln(\phi) + 0.02249781 \times \ln(SO_2) \quad (23)$$

with $R^2 = 0.9974$, (n=93)

Miscellaneous Equipment (Process Area 80): The main variables are flue gas flow rate (gas), SO_2 in coal, L/G, and the stoichiometric ratio (ϕ). The regression model is:

$$\ln(C_{80}) = 2.523896 - 0.6969073 \times \ln(gas) + 0.01583244 \times \ln(SO_2) - 0.01161708 \times \ln(L / G) + 0.9585911 \times \ln(\phi) \quad (24)$$

with $R^2 = 0.9978$ (n=93).

Total Capital Cost:

The process capital cost is the sum total of the capital costs for each area:

$$C_i = C_{10} + C_{20} + C_{30} + C_{60} + C_{70} + C_{80} \quad (25)$$

The total plant capital cost is:

$$T_{pc} = (1 + f_{gf} + f_{eho}) \times C_i \times \{1 + f_{cproc} + f_{cproj} \times (1 + f_{cproc})\}$$

where

$f_{gf} \equiv$ general facilities factor

$f_{eho} \equiv$ engineering home and office factor

$f_{cproc} \equiv$ process contingency factor

$f_{cproj} \equiv$ project contingency factor

(26)

3.2.2 Fixed O&M Costs

The main variables that affect fixed O&M cost are flue gas flow rate (air), %S, L/G, stoichiometry (ϕ) and labor rate (*labor*). The regression analysis for this case is interesting since a straight linear regression provides a poor fit. Once again engineering knowledge is factored so as to construct a linear regression about two input variables, one of which is a composite variable, i.e. the total plant capital cost (T_{pc}) and the other is labor rate. The regression analysis using these inputs provides a good fit. The regression model for fixed cost is given by:

$$F = -0.2681854 + 0.011157 \times T_{pc} + 0.02668693 \times labor \quad (27)$$

where F is the total fixed O&M cost (mills/kWh). The $R^2 = 0.997$ ($n=95$) indicates a good fit as do the t -statistics for the regression coefficients. The reader is referred to the Appendix for details regarding the statistical analysis.

3.2.3 Variable Operating Costs

The PRCC analysis indicates that the performance variables, %S and stoichiometry (ϕ), and the cost variables, reagent, disposal, steam and power costs are important. However, in order to perform a linear regression for variable costs, composite variables which represent the costs of different process areas need to be constructed. These composite variables are nonlinear (as is shown later) and the rank transform used in PRCC is not effective in linearizing this relationship. Therefore the effect of performance variables such as L/G ratio and size (MW) are not reflected in the PRCC analysis. Engineering experience is used to construct the form of the composite variables. Based on results from FGDCOST we identify four variable costs: (1) reagent use cost (*rcost*), (2) solid waste disposal cost (*scost*), (3) power cost (*pcost*), and (4) steam costs (*stcost*). We then develop regression models for each of these variable costs. The variable costs for all areas are expressed in mills/kWh. Finally the total variable cost is written as the sum of these area costs.

Reagent Cost

The total amount of reagent usage costs (*rcosts*, mills/kWh) depends on the cost of reagent (*reag* in \$/ton), and the amount of reagent used. The amount of reagent used depends on the total sulfur in flue gas, sulfur removal efficiency, and the stoichiometric ratio. Therefore we construct a composite variable ($C_{reag} = reag \times \eta_{SO2} \times SO2 \times \phi$) to represent the total cost of reagent used and regress this variable against the cost from the model (*rcost*). The regression model is:

$$rcost = 1.821221 \times 10^{-5} \times C_{reag} \quad (28)$$

with $R^2 = 0.9999$, and $n=96$.

Solid Waste Disposal Costs

The solid waste disposal cost ($scost$, mills/kWh) depends on the amount of solid waste and the unit cost of disposal. The solid waste can either be disposed ($disposal$, in \$/ton) or stacked ($stack$, in \$/ton). The amount of solid waste produced depends on the %S in coal and the removal efficiency. We construct two composite variables, cost of disposal $C_{solid} = solid \times SO2 \times \eta_{SO2}$, and cost of stacking $C_{stacking} = stack \times SO2 \times \eta_{SO2}$. A regression of $scost$ against these two composite variables yields the following response surface.

$$scost = 3.20172 \times 10^{-5} \times C_{solid} \quad (29)$$

with $R^2 = 0.9975$, and $n=95$

Power Costs

The power costs ($pcost$, mills/kWh) depends on the unit cost of power ($power$, mills/kWh), consumption of power in the reagent handling and flue gas handling systems. Slurry is injected into the absorber towers at four levels to remove SO_2 and each level is served by a pump. The power consumption for these pumps depends on the amount of slurry being pumped and the power cost. The first composite variable $C_{slurry} = power \times L / G \times gas$ is used to represent this cost. The power consumed by I.D. fans to overcome the flow resistance (or equivalent pressure drop ΔP) of the absorber towers and ducts to the chimney is represented by $C_{gas} = power \times gas \times \Delta P$. The response surface is constructed for the total power cost ($fyfuel$, \$/yr) since dividing the composite variables by MW , the plant capacity, leads to a regression of a nonlinear form. Therefore, the response surface for $pcost$ is provided in terms of $fyfuel$ as follows:

$$\begin{aligned} fyfuel &= 0.161388 \times C_{SO2} + 1.467771 \times C_{gas} \\ \therefore pcost &= fyfuel / (cf \times 8760 \times MW) \end{aligned} \quad (30)$$

where cf is the plant capacity factor (an input variable fixed at 65%) and the number 8760 expresses the hrs/yr and is used as a conversion factor. The fit of the regression is $R^2 = 0.9931$, $n=95$.

Steam Costs

The saturated gas exiting the absorbers at $127^{\circ}F$ is reheated to $152^{\circ}F$ using steam heat. The steam cost ($stcost$, mills/kWh) of reheating depends on cost of steam ($steam$, in \$/Klbs), flue gas flow rate ($Kacfm$) and the $\Delta T = 25^{\circ}F$ of reheat. We define a composite variable $C_{steam} = steam \times gas \times \Delta T$ and generate a response surface for $stcost$ in terms of the total steam costs ($fysteam$, \$/yr):

$$\begin{aligned} fysteam &= 7.901601 \times C_{steam} \\ \therefore stcost &= fysteam / (cf \times 8760 \times MW) \end{aligned} \quad (31)$$

where $R^2 = 0.999$ and $n=95$.

Total Variable Operating Costs

The total variable costs (mills/kWh) are derived as the sum of the reagent, solid waste disposal, power and steam costs as follows:

$$varcost = rcost + scost + pcost + stcost \quad (32)$$

Notice that the variables, *power*, *steam* and *water*, appear in the composite variables with L/G and hence L/G cannot be dropped in spite of its low PRCC values. The Appendix presents the regression analysis used to identify a regression model for different variable costs. Notice that initially the regression analysis for some of the variable costs leads to a high standard error for the intercept. A regression by forcing the intercept to be zero provides an excellent fit

3.3 Wet Limestone with Additives

The PRCC analysis for the LS/DBA system was performed. Based on this analysis the input variables are rank ordered in terms of the magnitude of their influence on the costs in Table 7 below:

Table 7 PRCC analysis for Wet Limestone with Dibasic Acid

Variables	Range	Capital	Costs	Fixed	Costs	Variable	Costs
		PRCC	Rank	PRCC	Rank	PRCC	Rank
flue gas(Kacfm)	625-3350	-0.991	1	-0.988	1	0.093	9
size (MW)	(300-1000)						
L/G(gpm/Kacfm)	50-90	0.293	4	0.307	5	0.576	4
%S(weight)	0.5-5.0	0.803	3	0.625	4	0.985	1

stoichiometry	1.05-1.15	0.852	2	0.764	3	0.379	6
labor rate(\$/hr)	15-25	0.071	11	0.867	2	-0.095	8
reagent cost(\$/ton)	10-20	-0.031	13	-0.07	8	0.659	2
solid disposal	5-10	-0.173	6	-0.216	6	0.534	5
waste(\$/ton) stack	3-8	0.075	9	-0.004	12	0.606	3
powercost(mil's/kwh)	40-60	0.056	12	-0.015	10	0.068	12
steam cost(\$/Klbs)	3-4	0.192	5	0.004	13	0.026	7
water fresh(\$/Klbs)	0.5-0.75	-0.073	10	-0.029	9	0.085	10
water cool(\$/Klbs)	0.1-0.2	-0.161	7	-0.013	11	-0.071	11
DBA ppm	1-2K	0		0		0	
DBA feed (lbs/ton of SO ₂)	15-25	-0.147	8	-0.08	7	0.007	13
DBA cost(\$/ton)	300-400	0		0		0	

Now we present the development of regression models using the variables identified by the PRCC analysis. The regression analysis is parallel to the one presented for LSFO with a data set that is different, i.e. generated using a cost model for LS/DBA. Therefore our description of the analysis is brief and the reader is referred to the Appendices for details.

3.3.1 Capital Costs

The main variables which affect the process capital costs (expressed in \$/kW) are flue gas flow rate (Kacfm), %S in coal, L/G ratio (gpm/Kacfm), and stoichiometry. Note that the range of L/G for LS/DBA is between 60-90. Once again using logarithmic transformations, linear regressions for each process area are presented in the Appendix. We provide highlights for each process area.

Reagent Feed System (Process Area 10)

The main input variables are flue gas flow rate, stoichiometry, and SO₂. The regression model is:

$$\ln(C_{10}) = 3.037518 - 0.7841248 \times \ln(gas) + 0.2199026 \times \ln(SO_2) + 1.76738 \times \ln(\phi) \quad (33)$$

with $R^2 = 0.995$ (n=94).

SO₂ Removal System (Process Area 20)

The main input variables are flue gas flow rate(gas), L/G, SO₂, and stoichiometry. The model is:

$$\ln(C_{20}) = 2.898487 - 0.4501866 \times \ln(gas) + 0.03762383 \times \ln(SO_2) + 1.416616 \times \ln(\phi) \quad (34)$$

with $R^2 = 0.9977$ (n=93).

Flue gas System (Process Area 30)

The main variables are flue gas flow rate(gas), SO₂, L/G, and stoichiometry.

$$\ln(C_{30}) = 2.359523 - 0.3990599 \times \ln(gas) + 0.02008773 \times \ln(SO_2) + 1.361409 \times \ln(\phi) \quad (35)$$

with $R^2 = 0.9916$ (n=93).

Solids Handling System (Process Area 60)

The main variables are flue gas flow rate (gas), %S, and stoichiometry The model is:

$$\ln(C_{60}) = 1.295472 - 0.6127475 \times \ln(gas) + 0.3761581 \times \ln(SO_2) + 1.987594 \times \ln(\phi) \quad (36)$$

where $R^2 = 0.981$, n=94.

General Support Area (Process Area 70)

The main variables are flue gas flow rate (gas), %S, and stoichiometry. The model is:

$$\ln(C_{70}) = 2.712624 - 0.9025382 \times \ln(gas) + 0.02667431 \times \ln(SO_2) + 1.452731 \times \ln(\phi) \quad (37)$$

where $R^2 = 0.9974$ (n=94).

Miscellaneous Equipment (Process Area 80)

The main variables is size(MW). The model is:

$$\ln(C_{80}) = 2.499042 - 0.6970531 \times \ln(gas) + 0.0163421 \times \ln(SO_2) + 0.9591331 \times \ln(\phi) \quad (38)$$

where $R^2 = 0.9977$ (n=94).

Total Capital Cost:

The process capital cost is the sum total of the capital costs for each area:

$$C_i = C_{10} + C_{20} + C_{30} + C_{60} + C_{70} + C_{80} \quad (39)$$

The total plant capital cost is:

$$T_{pc} = (1 + f_{gf} + f_{eho}) \times C_i \times \{1 + f_{cproc} + f_{cproj} \times (1 + f_{cproc})\}$$

where

$$\begin{aligned} f_{gf} &\equiv \text{general facilities factor} \\ f_{eho} &\equiv \text{engineering home and office factor} \\ f_{cproc} &\equiv \text{process contingency factor} \\ f_{cproj} &\equiv \text{project contingency factor} \end{aligned} \quad (40)$$

3.3.2. Fixed O&M Costs

The main variables for the fixed costs (in mills/kWh) are once again flue gas flow rate (*gas*), %S, L/G, stoichiometry(ϕ), and labor rate (*labor*). The response surface is constructed using a composite variable, the process capital cost C_t , and *labor*. The regression model is provided below.

$$F = -0.2646564 + 0.0111555 \times T_{pc} + 0.0266771 \times labor \quad (41)$$

where F is the fixed O&M cost. The $R^2 = 0.9969$ (n=93) indicates good fit as do the t-statistics. The reader is referred to the Appendix for details.

3.3.3 Variable Operating Costs

The analysis for variable costs has once again to be conducted in a manner similar to the case of LSFO. Composite cost variables for different process areas are constructed based on engineering experience. For LS/DBA an additional composite cost to represent the cost corresponding to DBA feed is required. This is discussed in greater detail below.

Reagent Costs

The cost of reagent use ($rcost$) is regressed against the composite variable $C_{reag} = reag \times \eta_{SO2} \times SO2 \times \phi$, which represents the cost of reagent use. The regression model is presented below:

$$rcost = 1.784539 \times 10^{-5} \times C_{reag} \quad (42)$$

with $R^2 = 0.9981$, $n=96$.

Solid Waste Disposal Costs

The solid waste disposal cost ($scost$) is regressed against two composite variables, $C_{solid} = solid \times SO2 \times \eta_{SO2}$, the cost of disposal and $C_{stack} = stack \times SO2 \times \eta_{SO2}$ the cost of stacking solid waste. The regression model is presented below.

$$scost = 3.141843 \times 10^{-5} \times C_{solid} \quad (43)$$

with $R^2 = 0.9927$ $n=95$.

Power Costs

The power costs ($pcost$) is regressed against two composite variables, the power consumed for pumping slurry $C_{slurry} = power \times gas \times L / G$, and power consumed by I.D. fans $C_{gas} = power \times gas \times \Delta P$. The response is constructed for the total power consumed ($fyfuel$, \$/yr) which is related to $pcost$ as follows $fyfuel = pcost \times MW \times 8760 \times cf$, where MW is the plant capacity and cf is the capacity factor fixed at 65%. The regression model is as follows:

$$\begin{aligned} fyfuel &= 0.1071768 \times C_{slurry} + 1.843284 \times C_{gas} \\ \therefore pcost &= fyfuel / (cf \times 8760 \times MW) \end{aligned} \quad (44)$$

where $R^2 = 0.991$, $n=95$.

Steam Costs

The steam cost ($stcost$) for reheating is written in terms of the composite variable $C_{steam} = steam \times gas \times \Delta T$. The total steam cost $fysteam$ (\$/yr) is related to $stcost$ as follows $fysteam = pcost \times cf \times 8760 \times MW$ and is used for the regression models.

$$\begin{aligned} fysteam &= 7.883372 \times C_{steam} \\ \therefore stcost &= fysteam / (cf \times 8760 \times MW) \end{aligned} \quad (45)$$

with $R^2 = 0.9999$ n=96.

DBA Costs

For systems that use organic acid additives, another composite variable $C_{DBA} = DBA_{cost} \times DBA_{feed} \times SO_2 \times \eta_{SO_2}$ is used to describe the cost associated with the use of DBA. Here DBA_{cost} is the cost of DBA (\$/ton) and DBA_{feed} is feed rate of DBA (lbs/ton of SO_2 removed). The variable costs (*dbacost*) associated with the use of DBA is related to the amount of DBA which decomposes in the slurry during the absorption and oxidation reactions. The regression model is provided below.

$$dbacost = 5.487265 \times 10^{-9} \times C_{DBA} \quad (46)$$

with $R^2 = 0.9951$, n=96.

Total Variable Costs

The total variable cost (*varcost*) is the sum of the individual costs of the different areas provided above.

$$varcost = rcost + scost + pcost + stcost + dbacost \quad (47)$$

3.4 Magnesium-Enhanced Lime System

A PRCC analysis of the data set for Mg-Lime system was performed. Based on this analysis the input variables are ranked in order of importance in the table below.

Table 8: PRCC Analysis for Mg-Lime

Variables	Range	Process	Capital	Fixed	Costs	Variable	Costs
		PRCC	Rank	PRCC	Rank	PRCC	Rank
flue gas(Kacfm) (MW equivalent)	625-3350 (300-1000)	-0.986	1	-0.983	1	-0.174	7
L/G(gal/Kacfm)	30-56	0.149	4	0.006	12	-0.092	8
SO ₂ (inlet SO ₂ ppm)	0.5-5.0	0.890	2	0.784	3	0.986	1
stoichiometry	1.05-1.15	0.801	3	0.722	4	0.385	4
labor rate(\$/hr)	15-25	0.13	7	0.874	2	-0.04	11

reagent cost(\$/ton)	50-60	-0.069	10	-0.022	10	0.675	9
solid disposal	5-10	-0.133	5	-0.186	5	0.432	2
waste(\$/ton) stack	3-8	0.113	8	-0.074	8	0.0	3
powercost(mills/kwh)	40-60	-0.051	12	0.058	9	0.222	12
steam cost(\$/Klbs)	3-4	0.064	11	-0.078	7	0.222	5
water fresh(\$/Klbs)	0.5-0.75	-0.097	9	-0.016	11	0.175	6
water cool(\$/Klbs)	0.1-0.2	-0.131	5	-0.083	6	-0.062	10

Now we present the development of regression models using the variables identified from the PRCC analysis.

3.4.1 Capital Costs

The main variables which effect the process capital costs are flue gas flow rate (Kacfm), SO₂ the inlet SO₂ concentration in flue gas (ppm), L/G ratio (gpm/Kacfm), and stoichiometry. The capital costs are normalized by plant size and expressed in \$/kW units for all process areas¹. It is known from prior engineering experience that the relation between the capital costs and the input variables is exponential, hence we use logarithmic transformations on the variables and conduct a linear regression of the transformed variables. The results of a regression analysis using *Splus* is provided in the Appendix. The highlights are presented here.

Reagent Feed System (Process Area 10): The main input variables are flue gas flow rate (gas), SO₂ (inlet SO₂ in ppm), the stoichiometric ratio (ϕ), and L/G. The regression⁴ model is:

$$\ln(C_{10}) = 1.560119 - 0.5687046 \times \ln(gas) + 0.4206398 \times \ln(SO_2) + 1.620048 \times \ln(\phi) \quad (48)$$

$$R^2 = 0.9953, (n=93).$$

SO₂ Removal System (Process Area 20): The main input variables are flue gas flow rate (gas), SO₂, L/G and the stoichiometric ratio (ϕ). The regression model is :

¹ The fixed O&M costs and the variable operating costs are normalized in terms of \$/kWh where the normalization is done against the net power rating in MW.

⁴ Note that all logarithms in the cost models are to the base 10.

$$\ln(C_{20}) = 2.728355 - 0.438832 \times \ln(gas) + 0.02021906 \times \ln(SO_2) + 0.08423401 \times \ln(L / G) + 1.377543 \times \ln(\phi) \quad (49)$$

with $R^2 = 0.9975$ and $n=93$.

Flue Gas System (Process Area 30): The main input variables are flue gas flow rate (gas), SO_2 , and the stoichiometric ratio (ϕ). The regression model is:

$$\ln(C_{30}) = 2.366505 - 0.3991868 \times \ln(gas) + 1.360233 \times \ln(\phi) \quad (50)$$

with $R^2 = 0.9889$, ($n=93$).

Solids Handling System (Process Area 60): The main variables are flue gas flow rate (gas), SO_2 , the stoichiometric ratio (ϕ), and L/G. The regression model is:

$$\ln(C_{60}) = 1.185349 - 0.5551964 \times \ln(gas) + 0.4301355 \times \ln(SO_2) + 1.406037 \times \ln(\phi) \quad (51)$$

with $R^2 = 0.9824$ ($n=93$).

General Support Area (Process Area 70): The main variables are flue gas flow rate (gas), L/G, and the stoichiometric ratio (ϕ). The regression model is:

$$\ln(C_{70}) = 2.802868 - 0.9103186 \times \ln(gas) + 1.448908 \times \ln(\phi) + 0.01913444 \times \ln(SO_2) \quad (52)$$

with $R^2 = 0.9974$, ($n=93$)

Miscellaneous Equipment (Process Area 80): The main variables are flue gas flow rate (gas), SO_2 concentration in flue gas, L/G, and the stoichiometric ratio (ϕ). The regression model is:

$$\ln(C_{80}) = 2.501685 - 0.6971426 \times \ln(gas) + 0.01634989 \times \ln(SO_2) + 0.9604236 \times \ln(\phi) \quad (53)$$

with $R^2 = 0.9978$ ($n=93$).

Total Capital Cost:

The process capital cost is the sum total of the capital costs for each area:

$$C_i = C_{10} + C_{20} + C_{30} + C_{60} + C_{70} + C_{80} \quad (54)$$

The total plant capital cost is:

$$T_{pc} = (1 + f_{gf} + f_{eho}) \times C_i \times \{1 + f_{cproc} + f_{cproj} \times (1 + f_{cproc})\}$$

where

$$\begin{aligned} f_{gf} &\equiv \text{general facilities factor} \\ f_{eho} &\equiv \text{engineering home and office factor} \\ f_{cproc} &\equiv \text{process contingency factor} \\ f_{cproj} &\equiv \text{project contingency factor} \end{aligned} \quad (55)$$

3.4.2 Fixed O&M Costs

The main variables that affect fixed O&M cost are flue gas flow rate (air), %S, L/G, stoichiometry (ϕ) and labor rate (*labor*). The regression analysis for this case is interesting since a straight linear regression provides a poor fit. Once again engineering knowledge is factored so as to construct a linear regression about two input variables, one of which is a composite variable, i.e. the total plant capital cost (T_{pc}) and the other is labor rate. The regression analysis using these inputs provides a good fit. The regression model for fixed cost is given by:

$$F = -0.3644789 + 0.0133118 \times T_{pc} + 0.02620393 \times \text{labor} \quad (56)$$

where F is the total fixed O&M cost (mills/kWh). The $R^2 = 0.997$ ($n=95$) indicates a good fit as do the t -statistics for the regression coefficients. The reader is referred to the Appendix for details regarding the statistical analysis.

3.4.3 Variable Operating Costs

The PRCC analysis indicates that the performance variables, %S and stoichiometry (ϕ), and the cost variables, reagent, disposal, steam and power costs are important. However, in order to perform a linear regression for variable costs, composite variables which represent the costs of different process areas need to be constructed. These composite variables are nonlinear (as is shown later) and the rank transform used in PRCC is not effective in linearizing this relationship. Therefore the effect of performance variables such as L/G ratio and size (MW) are not reflected

in the PRCC analysis. Engineering experience is used to construct the form of the composite variables. Based on results of FGDCOST we identify four variable costs: (1) reagent use cost (*rcost*), (2) solid waste disposal cost (*scost*), (3) power cost (*pcost*), and (4) steam costs (*stcost*). We then develop regression models for each of these variable costs. The variable costs for all areas are expressed in mills/kWh. Finally the total variable cost is written as the sum of these area costs.

Reagent Cost

The total amount of reagent usage costs (*rcosts*, mills/kWh) depends on the cost of reagent (*reag* in \$/ton), and the amount of reagent used. The amount of reagent used depends on the total sulfur in flue gas, sulfur removal efficiency, and the stoichiometric ratio. Therefore we construct a composite variable ($C_{reag} = reag \times \eta_{SO2} \times SO2 \times \phi$) to represent the total cost of reagent used and regress this variable against the cost from the model (*rcost*). The regression model is:

$$rcost = 1.121617 \times 10^{-5} \times C_{reag} \quad (57)$$

with $R^2 = 0.9999$, and $n=96$.

Solid Waste Disposal Costs

The solid waste disposal cost (*scost*, mills/kWh) depends on the amount of solid waste and the unit cost of disposal. The solid waste can either be disposed (*disposal*, in \$/ton) or stacked (*stack*, in \$/ton). The amount of solid waste produced depends on the %S in coal and the removal efficiency. We construct two composite variables, cost of disposal $C_{solid} = solid \times SO2 \times \eta_{SO2}$, and cost of stacking $C_{stacking} = stack \times SO2 \times \eta_{SO2}$. A regression of *scost* against these two composite variables yields the following response surface.

$$scost = 3.916061 \times 10^{-5} \times C_{solid} \quad (58)$$

with $R^2 = 0.9975$, and $n=95$

Power Costs

The power costs (*pcost*, mills/kWh) depends on the unit cost of power (*power*, mills/kWh), consumption of power in the reagent handling and flue gas handling systems. Slurry is injected into the absorber towers at four levels to remove SO₂ and each level is served by a pump. The power consumption for these pumps depends on the amount of slurry being pumped and the power cost. The first composite variable $C_{slurry} = power \times L / G \times gas$ is used to represent this

cost. The power consumed by I.D. fans to overcome the flow resistance (or equivalent pressure drop ΔP) of the absorber towers and ducts to the chimney is represented by $C_{gas} = power \times gas \times \Delta P$. The response surface is constructed for the total power cost ($fyfuel$, \$/yr) since dividing the composite variables by MW , the plant capacity, leads to a regression of a nonlinear form. Therefore, the response surface for $pcost$ is provided in terms of $fyfuel$ as follows:

$$fyfuel = 59464.7 + 0.1650975 \times C_{SO_2} + 1.001896 \times C_{gas} \quad (59)$$

$$\therefore pcost = fyfuel / (cf \times 8760 \times MW)$$

where cf is the plant capacity factor (an input variable fixed at 65%) and the number 8760 expresses the hrs/yr and is used as a conversion factor. The fit of the regression is $R^2 = 0.9931$, $n=95$.

Steam Costs

The saturated gas exiting the absorbers at $127^\circ F$ is reheated to $152^\circ F$ using steam heat. The steam cost ($stcost$, mills/kWh) of reheating depends on cost of steam ($steam$, in \$/Klbs), flue gas flow rate (Kacfm) and the $\Delta T = 25^\circ F$ of reheat. We define a composite variable $C_{steam} = steam \times gas \times \Delta T$ and generate a response surface for $stcost$ in terms of the total steam costs ($fysteam$, \$/yr):

$$fysteam = 7.901601 \times C_{steam} \quad (60)$$

$$\therefore stcost = fysteam / (cf \times 8760 \times MW)$$

where $R^2 = 0.999$ and $n=95$.

Water Costs

Water from the scrubbing slurry is lost due to evaporation in the scrubber. The water costs pertain to the supply needed to replenish this loss. The water cost ($wcost$, mills/kWh) depends on the cost of fresh water ($waterf$, \$/1000 gallons), and the amount sulfur scrubbed defined by a composite variable $C_{water} = SO_2 \times \eta_{SO_2} \times \phi$. The water costs are provided by:

$$wcost = 1.432031 \times 10^{-5} \times C_{water} \quad (61)$$

where $R^2 = 0.9986$ and $n=97$.

Total Variable Operating Costs

The total variable costs (mills/kWh) are derived as the sum of the reagent, solid waste disposal, power and steam costs as follows:

$$\text{var cost} = r \text{ cost} + s \text{ cost} + p \text{ cost} + st \text{ cost} + w \text{ cost} \quad (62)$$

Notice that the variables, *power*, *steam* and *water*, appear in the composite variables with L/G and hence L/G cannot be dropped in spite of its low PRCC values. The Appendix presents the regression analysis used to identify a regression model for different variable costs. Notice that initially the regression analysis for some of the variable costs leads to a high standard error for the intercept. A regression by forcing the intercept to be zero provides an excellent fit

3.5 Lime Spray Dryer

A PRCC analysis of the data set for the lime spray dryer system was performed. Based on this analysis the input variables are ranked in order of importance in the table below.

Table 9: PRCC Analysis for Lime Spray Dryer

Variables	Range	Process	Capital	Fixed	Costs	Variable	Costs
		PRCC	Rank	PRCC	Rank	PRCC	Rank
flue gas(Kacfm) (MW equivalent)	625-3350 (300-1000)	-0.602	2	-0.694	3	-0.145	6
stoichiometry(ϕ)	1.1-1.8	0.325	3	0.305	4	0.789	2
SO ₂ (inlet SO ₂ ppm)	500-3500	0.827	1	0.737	2	0.977	1
gpm	70-400	0.151	5	0.169	5	0.104	8
labor rate(\$/hr)	15-25	0.061	10	0.867	1	0.063	9
reagent cost(\$/ton)	50-60	0.121	6	0.122	6	0.488	4
solid disposal	5-10	0.043	11	-0.1	8	0.645	3
waste(\$/ton) stack	3-8	0.102	8	-0.005	11	0.27	5
powercost(mills/kwh)	40-60	-0.165	4	0.086	10	-0.125	7
steam cost(\$/Klbs)	3-4	0.084	9	-0.088	9	0.046	10
water fresh(\$/Klbs)	0.5-0.75	0.106	7	0.114	7	0.001	11

Now we present the development of regression models using the variables identified from the PRCC analysis.

3.5.1 Capital Costs

Reagent Feed System (Process Area 10): The main input variables are flue gas flow rate (gas), SO₂ (inlet SO₂ in ppm), the stoichiometric ratio (ϕ), and L/G. The regression⁵ model is:

$$\ln(C_{10}) = 1.424123 - 0.550577 \times \ln(gas) + 0.4499927 \times \ln(SO_2) + 0.4494373 \times \ln(\phi) \quad (63)$$

$R^2 = 0.973$, (n=95).

SO₂ Removal System (Process Area 20): The main input variables are flue gas flow rate (gas) and slurry flow rate (gpm). The effect of flue gas flow rate is log-linear, however the effect of slurry flow rate is linear. The regression model is :

$$\ln(C_{20}) = 2.259638 - 0.1879267 \times \ln(gas) - 8.669591 \times 10^{-4} \times gpm \quad (64)$$

with $R^2 = 0.973$ and n=96. Notice that the slurry flow is not an input variable but is based on the recycle ratio and the inlet SO₂ concentrations. We provide a regression which allows the calculation of gpm from recycle ratio (rr) and flue gas flow rate (gas).

$$\ln(gpm) = -0.9195194 + 0.9950315 \times \ln(gas) - 0.03195727 \times \ln(rr) \quad (65)$$

$R^2 = 0.9938$ (n=96).

Flue Gas System (Process Area 30): The main input variable is flue gas flow rate (gas). A log-log plot of capital costs with flue gas flow rate indicates that the response is piecewise linear in the two regions: $gas < 2266$ Kacfm, and $gas > 2266$ Kacfm. Therefore two linear regression models have been constructed for the two regions. The regression model for this has to be constructed in two stages: The regression model is:

$$\ln(C_{30}) = \begin{cases} 2.837432 - 0.5291256 \times \ln(gas) & \text{if } gas < 2266 \\ 2.39163 - 0.4001656 \times \ln(gas) & \text{if } gas \geq 2266 \end{cases} \quad \begin{matrix} R^2 = 0.9673(n = 35) \\ R^2 = 0.9878(n = 60) \end{matrix} \quad (66)$$

Solids Handling System (Process Area 60): The main variables are flue gas flow rate (gas), and inlet SO₂ concentrations. The regression model is:

⁵ Note that all logarithms in the cost model are to the base 10.

$$\ln(C_{60}) = -0.3610955 \times \ln(gas) + 0.5069208 \times \ln(SO_2) \quad (67)$$

with $R^2 = 0.9866$ (n=97).

General Support Area (Process Area 70): The main variables are flue gas flow rate (gas), SO₂ concentration, and the stoichiometric ratio (ϕ). The regression model is:

$$\ln(C_{70}) = 2.934742 - 0.9277054 \times \ln(gas) - 0.01605435 \times \ln(\phi) + 0.01347692 \times \ln(SO_2) \quad (68)$$

with $R^2 = 0.9974$ (n=96)

Miscellaneous Equipment (Process Area 80): The main variables are flue gas flow rate (gas), SO₂ in coal, and the stoichiometric ratio (ϕ). The regression model is:

$$\begin{aligned} \ln(C_{80}) = & 2.568472 - 0.7007278 \times \ln(gas) + 0.0118181 \times \ln(SO_2) \\ & - 0.0018824 \times \ln(\phi) \end{aligned} \quad (69)$$

with $R^2 = 0.9995$ (n=96).

Total Capital Cost:

The process capital cost is the sum total of the capital costs for each area:

$$C_t = C_{10} + C_{20} + C_{30} + C_{60} + C_{70} + C_{80} \quad (70)$$

The total plant capital cost is:

$$T_{pc} = (1 + f_{gf} + f_{eho}) \times C_t \times \{1 + f_{cproc} + f_{cproj} \times (1 + f_{cproc})\}$$

where

$f_{gf} \equiv$ general facilities factor

$f_{eho} \equiv$ engineering home and office factor

$f_{cproc} \equiv$ process contingency factor

$f_{cproj} \equiv$ project contingency factor

(71)

3.5.2 Fixed O&M Costs

The main variables that affect fixed O&M cost are flue gas flow rate (air), SO₂, gpm, stoichiometry (ϕ) and labor rate (*labor*). The regression analysis for this case is interesting since a straight linear regression provides a poor fit. Once again engineering knowledge is factored so as to construct a linear regression about two input variables, one of which is a composite variable, i.e. the total plant capital cost (T_{pc}) and the other is labor rate. The regression analysis using these inputs provides a good fit. The regression model for fixed cost is given by:

$$F = 0.01676287 \times T_{pc} + 0.005205668 \times labor \quad (72)$$

where F is the total fixed O&M cost (mills/kWh). The $R^2 = 0.9976$ ($n=96$) indicates a good fit as do the t -statistics for the regression coefficients. The reader is referred to the Appendix for details regarding the statistical analysis.

3.5.3 Variable Operating Costs

The PRCC analysis indicates that the performance variables, %S and stoichiometry (ϕ), and the cost variables, reagent, disposal, steam and power costs are important. However, in order to perform a linear regression for variable costs, composite variables which represent the costs of different process areas need to be constructed. These composite variables are nonlinear (as is shown later) and the rank transform used in PRCC is not effective in linearizing this relationship. Therefore the effect of performance variables such as L/G ratio and size (MW) are not reflected in the PRCC analysis. Engineering experience is used to construct the form of the composite variables. Based on results from FGDCOST we identify four variable costs: (1) reagent use cost (*rcost*), (2) solid waste disposal cost (*scost*), (3) power cost (*pcost*), and (4) steam costs (*stcost*). We then develop regression models for each of these variable costs. The variable costs for all areas are expressed in mills/kWh. Finally the total variable cost is written as the sum of these area costs.

Reagent Cost

The total amount of reagent usage costs (*rcosts*, mills/kWh) depends on the unit cost of reagent (*reag* in \$/ton), and the amount of reagent used. The amount of reagent used depends on the total sulfur in flue gas, and the stoichiometric ratio. Therefore we construct a composite variable ($C_{reag} = reag \times SO_2 \times \phi$) to represent the total cost of reagent used and regress this variable against the cost from the model (*rcost*). The regression model is:

$$rcost = 9.088184 \times 10^{-6} \times C_{reag} \quad (73)$$

with $R^2 = 0.9999$, and $n=96$.

Solid Waste Disposal Costs

The solid waste disposal cost (s_{cost} , mills/kWh) depends on the amount of solid waste and the unit cost of disposal. The solid waste can either be disposed ($disposal$, in \$/ton) or stacked ($stack$, in \$/ton). The amount of solid waste produced depends on the %S in coal and the removal efficiency. We construct a composite variable, cost of disposal $C_{solid} = solid \times SO_2 \times \eta_{SO_2}$. A regression of s_{cost} against the composite variable and the recycle ratio (rr) yields the following response surface:

$$s_{cost} = 3.858347 \times 10^{-5} \times C_{solid} + 0.01181688 \times rr \quad (74)$$

with $R^2 = 0.99$, and $n=95$.

Power Costs

The power costs (p_{cost} , mills/kWh) depends on the unit cost of power ($power$, mills/kWh), consumption of power in the reagent handling and flue gas handling systems. Slurry is injected into the spray dryer using an atomizer and pumps. The power consumption for these pumps depends on the amount of slurry being pumped and the power cost. The first composite variable $C_{slurry} = power \times gpm$ is used to represent this cost. The power consumed by I.D. fans to overcome the flow resistance (or equivalent pressure drop ΔP) of the absorber towers and ducts to the chimney is represented by $C_{gas} = power \times gas \times \Delta P$. The response surface is constructed for the total power cost ($fyfuel$, \$/yr) since dividing the composite variables by MW , the plant capacity, leads to a regression of a nonlinear form. Therefore, the response surface for p_{cost} is provided in terms of $fyfuel$ as follows:

$$fyfuel = 412602.5 + 54.29614 \times C_{SO_2} - 0.01351677 \times C_{gas} \quad (75)$$

$$\therefore p_{cost} = fyfuel / (cf \times 8760 \times MW)$$

where cf is the plant capacity factor (an input variable fixed at 65%) and the number 8760 expresses the hrs/yr and is used as a conversion factor. The fit of the regression is $R^2 = 0.9461$, $n=95$.

Water Costs

Water from the scrubbing slurry is lost due to evaporation in the scrubber. The water costs pertain to the supply needed to replenish this loss. The water cost (w_{cost} , mills/kWh) depends

on the cost of fresh water (*waterf*, \$/1000 gallons), and the amount sulfur scrubbed defined by a composite variable $C_{water} = SO_2 \times \phi$. The water costs are provided by:

$$wcost = 4.43957 \times 10^{-6} \times C_{water} \quad (76)$$

where $R^2 = 0.9882$ and $n=97$.

Total Variable Operating Costs

The total variable costs (mills/kWh) are derived as the sum of the reagent, solid waste disposal, power and steam costs as follows:

$$varcost = rcost + scost + pcost + wcost \quad (77)$$

3.6 Sparing Philosophy

The models developed in the previous subsections are based on the assumption of two operating absorber towers (each rated at 50% of the plant capacity, with a maximum of 650 MW) and one spare absorber tower. With the recent development of advanced absorber towers of near 100% reliability it is possible to eliminate the spare tower and reduce capital costs significantly (no significant changes in variable costs). Another approach to reducing capital cost exploits the economy of scale by using single absorber towers of large capacity. In this subsection we develop scaling factors to estimate the reduction in capital costs for different FGD systems by changing the sparing philosophy. Note that in all the factors developed below, the scaling factors are derived by regressing the capital cost of various sparing philosophies against the capital cost of the default system, i.e. an FGD with 2 operating towers and 1 spare.

3.6.1 Zero Sparing

In order to estimate the reduction in capital costs by eliminating the spare tower, we set the number of spare towers in FGDCOST to zero and exercise the model for various values of input for the relevant variables such as flue gas flow rate (*gas*), %S, L/G, and stoichiometry. The capital costs for system with no spares (C') is then regressed against the capital cost for the system with one spare tower (C) and correlation coefficient of this regression provides us with the scaling factor we desire. The capital cost of the SO₂ removal system (includes absorber towers) and the flue gas system (includes piping to absorber towers) is affected by the sparing philosophy. We present results only for those process areas whose capital cost is affected.

Limestone Forced Oxidation

Denoting the capital cost of LSFO with no spares by C' and regressing this against C we have:

$$\text{SO}_2 \text{ Removal System: } C'_{20} = 0.6790025 + 0.7192759 \times C_{20} \quad R^2 = 0.9994, n = 97 \quad (78a)$$

$$\text{Flue Gas System: } C'_{30} = -0.2228743 + 0.9569927 \times C_{30} \quad R^2 = 1.0, n = 97 \quad (78b)$$

This implies a reduction of about 25-30% in the capital costs of the SO₂ removal system and 5% reduction in the flue gas system.. Note that both values of C and C' are about 100, therefore the influence of the intercept is less than 1%.

Limestone with Dibasic Acid

Similarly the capital cost of LS/DBA with no spares is given by:

$$\text{SO}_2 \text{ Removal System: } C'_{20} = 0.216904 + 0.718509 \times C_{20} \quad R^2 = 0.9993, n = 97 \quad (79a)$$

$$\text{Flue Gas System: } C'_{30} = -0.2325696 + 0.9577534 \times C_{30} \quad R^2 = 1.0, n = 97 \quad (79b)$$

Once again the SO₂ removal system shows a reduction of over 25-30% in capital costs and the flue gas system shows 5% decrease.

Magnesium-Enhanced Lime

Denoting the capital cost of Mg-Lime with no spares by C' and regressing this against C we have:

$$\text{SO}_2 \text{ Removal System: } C'_{20} = 0.6926448 \times C_{20} \quad R^2 = 1, n = 97 \quad (80a)$$

$$\text{Flue Gas System: } C'_{30} = 0.9437713 \times C_{30} \quad R^2 = 1.0, n = 97 \quad (80b)$$

This implies a reduction of about 30% in the capital costs of the SO₂ removal system and 5-10% reduction in the flue gas system. Note that both values of C and C' are about 100, therefore the influence of the intercept is less than 1%.

Lime Spray Dryer

The cost with no spares is given by:

$$\text{SO}_2 \text{ Removal System: } C'_{20} = 0.7114379 \times C_{20} \quad R^2 \sim 1, (n = 98) \quad (81a)$$

$$\text{Flue Gas System: } C'_{30} = 0.8002883 \times C_{30} \quad R^2 \sim 1, (n = 98) \quad (81b)$$

3.6.2 One Absorber Only

The reduction in capital cost by reducing the number of absorbers from two (rated @50% each) to one (rated @100% of capacity) is examined here. Once again the capital cost of a system with one absorber (rated @100%) and no spares (C') is compared against the default system with 2 operating towers with 1 spare (C).

Limestone Forced Oxidation

Once again the capital costs of the SO₂ removal system and flue gas system show a decrease as shown by the regression models below:

$$\text{SO}_2 \text{ Removal System: } C'_{20} = 2.010769 + 0.5561327 \times C_{20} \quad R^2 = 0.9973, n = 97 \quad (82a)$$

$$\text{Flue Gas System: } C'_{30} = -0.402627 + 0.9708304 \times C_{30} \quad R^2 = 0.9999, n = 97 \quad (82b)$$

Using a single tower @100% capacity results in the economy of scale compared to the case with 2 towers @50%. The capital cost for the SO₂ removal system shows a reduction of over 40-45% and the flue gas system shows a reduction of around 5%.

Limestone with Dibasic Acid

Similarly the capital cost of LS/DBA with 1 absorber @ 100% capacity is given by:

$$\text{SO}_2 \text{ Removal System: } C'_{20} = 0.9438254 + 0.5571021 \times C_{20} \quad R^2 = 0.9973, n = 97 \quad (83a)$$

$$\text{Flue Gas System: } C'_{30} = -0.3750865 + 0.9543123 \times C_{30} \quad R^2 = 0.9999, n = 97 \quad (83b)$$

Once again the capital cost of the SO₂ removal system shows substantial reduction of over 40-45%.

Magnesium-Enhanced Lime

Once again the capital costs of the SO₂ removal system and flue gas system show a decrease as shown by the regression models below:

$$\text{SO}_2 \text{ Removal System: } C'_{20} = 0.525391 \times C_{20} \quad R^2 = 0.9998, n = 97 \quad (84a)$$

$$\text{Flue Gas System: } C'_{30} = 0.9227762 \times C_{30} \quad R^2 = 1, n = 97 \quad (84b)$$

Using a single tower @100% capacity results in the economy of scale compared to the case with 2 towers @50%. The capital cost for the SO₂ removal system shows a reduction of over 45% and the flue gas system shows a reduction of over 5%.

Lime Spray Dryer

The FGDCOST model for spray towers is good up to 650MW. Therefore the comparisons presented here are valid only up to this size. The regressions, show a constant downward shift in the cost:

$$\text{SO}_2 \text{ Removal System: } C'_{20} = -13.94887 + 0.9075569 \times C_{20} \quad R^2 = 0.9955 \quad (n = 55) \quad (85a)$$

$$\text{Flue Gas System: } C'_{30} = 0.5592991 + 0.6021688 \times C_{30} \quad R^2 = 0.9978 \quad (n = 97) \quad (85b)$$

4.0 A Numerical Example

In this section we provide an illustrative example which uses the models developed above to calculate the capital and O&M costs for a given set of inputs. For the example we use the default value for the input parameters provided earlier in Table 5 and compute costs for the limestone forced oxidation (LSFO) FGD system. A plant of gross capacity of 313 MW (300 MW net) with flue gas flow rate of 971 Kacfm, an inlet SO₂ concentration of 2000 ppm, and a chloride concentration in slurry of 32,000 ppm is assumed. We design the FGD system for an SO₂ removal efficiency of 90% with a stoichiometric ratio of 1.03. Equation 7 is first used to evaluate the required L/G ratio:

$$0.9 = 1 - \exp \left\{ \begin{array}{l} -[0.725 + 0.0175 \times L / G - 2.5 \times 10^{-4} \times (1750 - 2000)] \\ + 5.14 \times 10^{-6} \times (32000 - 25000) \end{array} \right\}$$
$$\Rightarrow L / G \approx 90 \text{ gpm} / \text{Kacfm}$$

(a) Process Capital Costs

Reagent Feed System (Process Area 10):

$$\begin{aligned} \ln(C_{10}) &= 3.050657 - 0.7902577 \times \ln(971) + 0.2203442 \times \ln(1750) \\ &\quad + 1.664678 \times \ln(1.03) \\ &= 26.67 \text{ \$} / \text{kW} \end{aligned}$$

SO₂ Removal System (Process Area 20):

$$\begin{aligned} \ln(C_{20}) &= 2.670582 - 0.4402019 \times \ln(971) + 0.01914421 \times \ln(1750) \\ &\quad + 0.1375112 \times \ln(90) + 1.362315 \times \ln(1.03) \\ &= 50.57 \text{ \$} / \text{kW} \end{aligned}$$

Flue Gas System (Process Area 30):

$$\begin{aligned}\ln(C_{30}) &= 2.422378 - 0.3985948 \times \ln(971) + 1.372434 \times \ln(1.03) \\ &= 17.75 \text{ \$ / kW}\end{aligned}$$

Solids Handling System (Process Area 60):

$$\begin{aligned}\ln(C_{60}) &= 1.32039 - 0.6194293 \times \ln(971) + 0.3790324 \times \ln(1750) \\ &\quad + 1.655228 \times \ln(1.03) \\ &= 5.25 \text{ \$ / kW}\end{aligned}$$

General Support Area (Process Area 70):

$$\begin{aligned}\ln(C_{70}) &= 2.717534 - 0.8992044 \times \ln(971) + 1.433288 \times \ln(1.03) + 0.02249781 \times \ln(1750) \\ &= 1.32 \text{ \$ / kW}\end{aligned}$$

Miscellaneous Equipment (Process Area 80):

$$\begin{aligned}\ln(C_{80}) &= 2.523896 - 0.6969073 \times \ln(971) + 0.01583244 \times \ln(1750) \\ &\quad - 0.01161708 \times \ln(90) + 0.9585911 \times \ln(1.03) \\ &= 3.0 \text{ \$ / kW}\end{aligned}$$

Total Process Capital Costs

$$\begin{aligned}T_{pc} &= (26.67 + 50.57 + 17.75 + 5.25 + 1.32 + 3.0) \\ &= 104.56 \text{ \$ / kW}\end{aligned}$$

Capital Cost Summary:

<u>Capital Investment</u>	<u>\$/kw</u>
Process Capital	104.56
General Facilities (10%)	10.46
Engg. & Home Office Fees(10%)	10.46
Project Contingency (15%)	19.20
Process Contingency (2%)	2..51
Total Plant Cost	147.19 ^a

Total Cash Expended	143.66
(Adjust Factor=0.976)	
AFDUC (Adjust Factor=0.0548)	8.07
Total Plant Investment (TPI)	151.73
Royalty Allowance (0.5%)	0.82
Preproduction costs	6.24 ^b
Inventory Capital	0.78 ^c
Total Capital Requirement	159.57

a Total plant cost is given as

$$T_{pc} = (1 + f_{gf} + f_{aho}) \times C_i \times \{1 + f_{cproc} + f_{cproj} \times (1 + f_{cproc})\}$$

where

$f_{gf} \equiv$ general facilities factor (10%)

$f_{aho} \equiv$ engineering and H.&O. factor (10%)

$f_{cproc} \equiv$ process contingency factor (2%)

$f_{cproj} \equiv$ project contingency factor (15%)

b Preproduction costs (one month)= fixed costs + variable costs + 0.02 TPI

$$= (1.91 + 2.07) \times 0.001 \times 24 \times 30 + 3.37$$

$$= 6.24$$

c Inventory Capital = reagent flow tons/hr*24*storage time*limestone cost/mw

$$= 10.86 \times 24 \times 60 \times 15 / 3000000 = 0.78$$

(b) Fixed O&M Costs

$$F = -0.2681854 + 0.011157 \times 147.19 + 0.02668693 \times 20$$

$$= 1.91 \text{ mills / kWh}$$

(c) Variable O&M Costs

Reagent Cost

$$C_{reag} = 15 \times 0.9 \times 1.03 \times 1750 = 24333.8$$

$$r_{cost} = 1.821221 \times 10^{-5} \times 24333.8$$

$$= 0.44 \text{ mills / kWh}$$

Solid Waste Disposal Costs

$$C_{solid} = 8.15 \times 1750 \times 0.9 = 12836.25$$

$$\begin{aligned} scost &= 3.20172 \times 10^{-5} \times 12836.25 \\ &= 0.41 \text{ mills / kWh} \end{aligned}$$

Power Costs

$$C_{slurry} = 50 \times 90 \times 971 = 4.3695 \times 10^5$$

$$C_{gas} = 50 \times 971 \times 10 = 4.855 \times 10^5$$

$$\begin{aligned} f_{fuel} &= 0.161388 \times 4.3695 \times 10^5 + 1.467771 \times 4.855 \times 10^5 \\ &= 1,417787.7 \end{aligned}$$

$$\begin{aligned} \therefore pcost &= 1.417787.7 / (0.65 \times 8760 \times 300) \\ &= 0.83 \text{ mills / kWh} \end{aligned}$$

Steam Costs

$$C_{steam} = 3.5 \times 971 \times 25 = 84962.5$$

$$f_{steam} = 7.901601 \times 84962.5$$

$$\begin{aligned} \therefore stcost &= 84962.5 / (cf \times 8760 \times MW) \\ &= 0.39 \text{ mills / kWh} \end{aligned}$$

Total Variable Operating Costs

The total variable costs (mills/kWh) are derived as the sum of the reagent, solid waste disposal, power and steam costs as follows:

$$varcost = 0.44 + 0.41 + 0.83 + 0.39 = 2.07 \text{ mills / kWh}$$

(d) Sparing Philosophy

Zero Sparing

Denoting the capital cost of LSFO with no spares by C' we have:

$$\text{SO}_2 \text{ Removal System: } C'_{20} = 0.6790025 + 0.7192759 \times 50.57 = 37.05 (\$/kW)$$

$$\text{Flue Gas System: } C'_{30} = -0.2228743 + 0.9569927 \times 17.75 = 16.76 (\$/kW)$$

100% Capacity

Denoting the capital cost of LSFO with 1 absorber of 100% capacity with no spares by C' we have

$$\text{SO}_2 \text{ Removal System: } C'_{20} = 2.010709 + 0.5561327 \times 50.57 = 30.13 \quad (\$/kW)$$

$$\text{Flue Gas System: } C'_{30} = -0.402627 + 0.9708304 \times 17.75 = 16.83 \quad (\$/kW)$$

The new costs for these component areas then would be used in the total cost algorithms described earlier.

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