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Modeling of Integrated Environmental Control Systems for Coal-Fired Power Plants

Quarterly Progress Report

To

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U.S. Department of Energy
Pittsburgh, Pennsylvania 15236

From

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1. Introduction

This is the fourth quarterly report of DOE Contract No. DE-AC22-87PC79864, entitled "Modeling of Integrated Environmental Control Systems for Coal-Fired Power Plants." This report summarizes accomplishments during the period July 1, 1988 to September 30, 1988. Our efforts during the last quarter focused primarily on the completion, testing and documentation of the NOXSO process model. The sections below present the details of these developments.

2. Deterministic Analysis for the NOXSO Process

This chapter presents a deterministic analysis of the NOXSO process and compares it to a conventional plant design with wet Flue Gas Desulfurization (FGD) and Selective Catalytic Reduction (SCR) and to the advanced Copper Oxide process design. Details of the performance and economic algorithms for the NOXSO process were presented in the July quarterly report [1].¹

The NOXSO process diagram is shown in Figure 2-1. This case study assumed a new 522 MW power plant burning an Illinois #6 coal. The properties of this coal are shown in Table 2-1, while key input assumptions for the NOXSO process are shown in Table 2-2. The financial assumptions and emission standards are shown in Table 2-3. Current federal New Source Performance Standards (NSPS) for SO₂ and particulates are assumed to apply. The assumed NO_x emission constraint, however, is more stringent than the current NSPS, and requires an emission reduction of 85 to 90 percent below current U.S. requirements. This is representative of current NO_x control requirements in Japan, West Germany, and California.

2.1 Cost Results

Table 2-4 shows the calculated capital and levelized total revenue requirement for the assumptions shown in Tables 2-1 to 2-3. The assumptions for the wet FGD and Copper Oxide process are shown in Reference [2]. As can be seen, the NOXSO process has the lowest capital and revenue requirements for the unwashed coal. The revenue requirement is 8% and 15% lower than the conventional and Copper Oxide designs, respectively. For the washed coal, its capital cost is still 13% less than the Copper Oxide process, while the total revenue requirement is 5% higher.

Table 2-5 shows the cost breakdown for the capital and operating cost for the NOXSO process. For the unwashed coal, the direct capital cost is divided relatively equally between the equipment associated with the sorbent, flue gas, and acid plant. Most of the indirect costs,

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¹ A typographical error was found in the July quarterly report, where the parameter λ was defined as the ratio of available sodium to sulfur removed. Rather, it is the stoichiometric ratio of sulfur removed to available sodium.

Figure 2-1: NOXSO Process Diagram

**Table 2-1: Properties of Illinois No. 6 Coal Used for Case Studies
(As-Fired Basis)**

Coal Property	Run-of-Mine Coal	Washed Coal ^a
Btu/lb	10,190	10,330
% Sulfur	4.4	3.1
% Carbon	57.0	57.7
% Hydrogen	3.7	4.0
% Oxygen	7.2	8.4
% Nitrogen	1.1	1.1
% Moisture	12.3	17.5
\$/ton (at mine) ^b	28.10	32.70
\$/ton (transport) ^b	5.90	5.90

^a Model results for a 30% reduction on lbs/MBtu basis using conventional coal cleaning (Level 3 plant design).

^b 1985 dollars, assuming a midwestern plant and mine location.

Table 2-2: NOXSO Performance and Economic Assumptions

Parameter	Value	Parameter	Value
Air Preheater Enlargement	no	Regenerator Utilization of:	
Bed Height (in)	36	Methane (%)	90
Bed Pressure Drop (in water)	22	Steam (%)	30
Bed Temperature (°F)	250	Residence Time of Sorbent in:	
Flue Gas Bed Velocity (ft/s)	2.95	Cooler (min)	18
Gas Burner for Regenerator Gas	yes	Heater (min)	28
K _a , Rate Constant (atm ⁻¹ s ⁻¹)	3.7	Regenerator (min)	35
Methane Cost (\$/MSCF)	4.50	Sodium Loading (% of weight)	3.8
NO _x "destruction" efficiency (%)	65	Sorbent Characteristics:	
Oil Cost (\$/MBtu)	4.50	Attrition Rate (% of sorbent flow)	0.07
Oil Sulfur Correction Factor	0.6	Cost (\$/lb)	2.57
Regenerator Gas Sulfur Composition:		Density (lbs/ft ³)	30
H ₂ S (% of S regenerated)	30	Sorbent Regeneration Efficiency by:	
S ₂ (% of S regenerated)	5	Methane (%)	81.4
SO ₂ (% of S regenerated)	65	Steam (%)	90.7
		Overall Calc. (%)	98.3

Table 2-3: Financial and Emission Assumptions for Case Studies

Model Parameter	Nominal Value
<u>Emission Constraints</u>	
Nitrogen Oxides	0.08 lbs/MBtu
Sulfur Dioxide	90% Removal
Particulates	0.03 lbs/MBtu
<u>Power Plant Parameters</u>	
Gross Capacity	522 MW
Gross Heat Rate	9500 Btu/kWh
Capacity Factor	65%
Excess Air (boiler / total)	20% / 39%
Ash to Flue Gas	80%
Sulfur to Flue Gas	97.5%
Preheater Inlet Temperature	700 °F
Preheater Outlet Temperature	300 °F
<u>Financial Parameters</u>	
Inflation Rate	0%
Debt Fraction / Real Return	50% / 4.6%
Common Stock Fraction / Real Return	35% / 8.7%
Preferred Stock Fraction / Real Return	15% / 5.2%
Federal Tax Rate	36.7%
State Tax Rate	2.0%
Ad Valorem Rate	2.0%
Investment Tax Credit	0%
Book Life	30 years
Real Escalation Rates	0%
Real Discount Rate (calc)	6.13%
Fixed Charge Factor (calc)	10.34%
Levelization Factors	1.000

TABLE 2-4: Summary of Total Pollution Control Costs
(4.4% S Illinois No. 6 Coal)
All Costs in Constant 1985 Dollars

Process Configuration	No Coal Cleaning		Washed Coal ^a	
	Capital (\$/net kW)	Level Rev Req (Mills/kWh)	Capital (\$/net kW)	Level Rev Req (Mills/kWh)
<u>Conventional Plant Design</u>				
Coal Cleaning	-	-	(6) ^b	1.9
SCR System	69	3.7	70	3.7
Cold-Side ESP	43	0.9	40	0.8
Wet FGD System	197	9.0	192	8.3
Solid Waste Disposal	56	2.0	43	1.6
Total	365	15.6	339	16.3
<u>Copper Oxide Design</u>				
Coal Cleaning	-	-	(3) ^b	1.9
Copper Oxide Process	244	14.4	180	8.5
Fabric Filter	62	1.4	62	1.4
Solid Waste Disposal	25	1.1	21	1.0
Total	332	16.9	260	12.8
<u>NOXSO Design</u>				
Coal Cleaning	-	-	(4) ^b	1.9
NOXSO Process	178	11.9	149	9.2
Fabric Filter	60	1.4	59	1.4
Solid Waste Disposal	25	1.1	21	0.9
Total	263	14.4	225	13.4

^a 30% sulfur reduction on lbs/MBtu basis.

^b These costs (or credits) result from changes in the base plant cost.

TABLE 2-5: Breakdown of Cost for NOXSO Process

Component	Capital Cost (85 M\$)
Feed Equipment	1.8
Flue Gas Equipment	15.6
Sorbent Equipment	15.4
General Support Equipment	0.2
Acid Plant	<u>18.0</u>
Total Direct Capital	51.0
General Facilities	3.3
Engineering Fees	4.1
Project Contingency	7.0
Process Contingency	<u>4.4</u>
Total Plant Cost	69.9
AFUDC	<u>6.6</u>
Total Plant Investment	76.5
Royalties	0.2
Startup Costs	4.3
Working Capital	7.1
Initial Catalyst Cost	<u>7.2</u>
Total Capital Cost	95.2

Component	Operating and Maintenance Cost (85 M\$/yr)
Fixed Operating	4.0
Methane	6.0
Oil	8.6
Makeup Sorbent	13.6
Miscellaneous	0.6
Acid Plant O&M	1.5
Sulfuric Acid Sales	-8.3
Coal Credit	<u>-2.5</u>
Total Variable Cost	23.4

including startup costs, working and inventory capital are attributed to the sorbent flow rate. Approximately 40% of the total variable cost, excluding sulfuric acid sales and coal energy credits, is for sorbent makeup, while oil and gas consumption account for 25% and 18%, respectively. The fixed operating costs (12%), acid plant variable cost (4%), and miscellaneous cost (1%) account for the rest.

Sulfuric acid sales reduce the total variable cost by 24%, while the coal energy credit reduces the cost by 7%. For the washed coal the same general trends appear; however, the capital cost associated with the flue gas remains constant, while the direct capital costs associated with the sorbent and acid plant decrease by approximately 20%. The variable operating cost and sales of sulfuric acid decrease by 30%. The sulfuric acid plant capital cost is 37 \$/kW, while producing 8.3 M\$/yr from selling sulfuric acid and 0.9 M\$/yr as a coal credit from steam production. The variable and utility cost is 2.1 M\$/yr. This translates into a savings of 1.7 mills/kWh for the NOXSO process or a profit of 31 \$/ton for the sulfuric acid generated. For the washed coal, the capital cost is 30 \$/kW and the sulfuric acid sales are 5.8 M\$/yr. The net results is a savings of 1.1 mills/kWh or 28 \$/ton.

2.2 NO_x Removal Efficiency

Originally, NOXSO believed that the NO_x removal efficiency was a function of the SO₂ concentration and the bed temperature[3]. In Reference [4], it was estimated that at a bed temperature of 248°F the NO_x and SO₂ removal efficiencies would be equal. However, in the process developmental unit test at the Pittsburgh Energy Technology Center (PETC), the experimental data did not show any clear correlation between NO_x removal efficiency and bed temperature [5]. However, there was clear evidence that the NO_x removal efficiency was depended upon the regeneration gas. Methane was found to improve the NO_x removal efficiency compared to hydrogen. Four experimental test were conducted using methane at a bed temperature of approximately 245°F. The SO₂ and NO_x concentrations were held at approximately 2300 and 630 ppm, respectively. Three test were conducted at an 18 inch bed height; the other was at 42 inches. For three of the tests with removal efficiencies were greater than 84%, the SO₂ and NO_x removal efficiencies were within ± 2. For the other test, the removal efficiencies were much lower (SO₂ equalled 69%, while NO_x equalled 78%). For the reasons shown above the SO₂ and NO_x removal efficiencies are assumed to be equalled.

In Reference [5], NOXSO also showed that the NO_x removal efficiency was correlated to the ratio of SO₂ to NO_x by a factor of 0.80. However, all the data came from experiments which used hydrogen as a regeneration gas and it is not clear if the same trend would hold if methane is used as a regeneration gas. NOXSO also showed a significant transient effect of increasing the ratio of SO₂ to NO_x from 3.1 to 4.1; however, the steady-state effect was small, approximately a

3.5 removal points. Since this is a steady-state model, the transient change is not useful; although, it could be very useful to a power plant operator. These test were also conducted using hydrogen, so the steady-state effect may not be applicable using methane; especially is removal efficiencies of 90% or greater are required.

2.3 Effects of Stoichiometry

Table 2-6 gives a breakdown of the important performance parameters for the washed and unwashed coals. For both cases the required NO_x removal efficiency, (92.8%) is higher than the SO_2 efficiency, (91.1%). This is caused by the inefficient destruction of NO_x which is recycled to the boiler. Since the NO_x and SO_2 efficiencies are assumed to be equal, the molar stoichiometry is determined by the NO_x removal efficiency. As can be seen, the sorbent flow rate decreases by 30% for the washed coal. However, if the molar stoichiometry were driven by the SO_2 removal efficiency, its value for the unwashed and washed coals would be 2.8 and 2.2 respectively, instead of 3.3. These stoichiometries would produce sorbent flow rates of 1.1×10^6 lbs/hr and 0.57×10^6 lbs/hr, respectively. The NOXSO process capital cost would decrease from 178 to 166 \$/kW for the unwashed coal and from 149 to 129 \$/kW for the washed, and the total NOXSO process revenue requirement would fall to 10.4 mills/kWh and 6.8 mills/kWh, respectively. Thus, if the NO_x "destruction" efficiency were improved substantially, or the NO_x emission constraint were relaxed, the cost of the NOXSO process would decrease more substantially with coal cleaning.

Figure 2-2 shows the SO_2 removal efficiency versus molar stoichiometry for three different bed heights with the flue gas velocity and rate constant, K_a , held constant at 2.95 ft/s and 3.7 respectively. The molar stoichiometry is defined as the moles of sodium required to remove one mole of SO_2 . This value is equal to the inverse of the average conversion factor of sodium, X , times the effective molar ratio of sodium to sulfur removed [1]. According to NOXSO [6,7], the effective molar ratio of sodium to sulfur removed is one. Therefore, the equation shown below is the molar stoichiometry of sodium to sulfur removed.

$$\text{msr} = \frac{1}{\frac{\ln(1 - \eta)V}{K_a H} + 1} \quad (2.1)$$

where H = Bed height (inches)

K_a = Reaction rate constant ($\text{atm}^{-1} \text{s}^{-1}$)

msr = molar stoichiometry (Na to S removed)

V = Flue gas velocity (inches s^{-1})

η = SO_2 removal efficiency (fraction)

For a fixed value of K_a , increasing the bed height or decreases the flue gas velocity will decrease the molar stoichiometric ratio. While this equation is very useful for understanding and

predicting the performance of the absorber, it is only an approximation. A limitation of this equation is that the molar stoichiometry goes to infinity if the following conditions occur.

$$\eta = 1 - e^{-\frac{KaH}{V}} \quad (2.2)$$

For $Ka = 3.7 \text{ atm}^{-1}\text{sec}^{-1}$, $H = 36 \text{ in}$, and $V = 2.95 \text{ ft/s}$, the molar stoichiometry goes to infinity at $\eta = 97.7\%$. At 18 and 42 inches and the same values of Ka and V , the asymptotic values are 84.8% and 98.8%, respectively. It is important to know this asymptotic value since the molar stoichiometry increase rapidly as this value is approached. For the 36 inch bed height, the molar stoichiometry increases rapidly for values above 90% removal, since it is approaching its asymptotic value. For the unwashed coal, the molar stoichiometry is 3.31. If the bed height were increased to 42 inches, the molar stoichiometry would decrease by 25%. It was not possible to determine if this would be economical since pressure drop data does not exist for the 42 inch bed height.

2.4 Process Energy Requirements

The NOXSO process consumes and produces large quantities of energy as can be seen from Table 2-6. About half of the energy from the fuel oil consumption is recovered by recycling the heater gas to the boiler. Also, 35% of the energy from the methane and steam consumed in the regenerator is converted to steam in the acid plant. Approximately 50% of the electricity consumption for the NOXSO process is used to overcome the pressure drop across the absorber. The sorbent cooler compressor and the pneumatic conveying blowers consume about 35% and 13%, respectively. The acid plant consumes some additional electricity (2MW), while producing steam which can be used to regenerate the sorbent in the steam treatment vessel or sent to the steam cycle to produce electricity.

Table 2-6: NOXSO Performance Parameters

Parameter	Unwashed Coal	Washed Coal
Concentration of Pollutants entering Absorber (ppm)		
NO _x	806	811
SO ₂	3300	2350
Energy Consumption (-) or Production (+)		
Acid Plant		
Electricity (MW)	-2.0	-1.4
Steam (10 ⁶ Btu/hr)	+84.4	+59.0
NOXSO Process		
Electricity (MW)	-12.1	-10
Methane Consumption (10 ⁶ Btu/hr)	-234	-164
Oil Consumption (10 ⁶ Btu/hr)	-334	-230
Recycled Gas to Boiler (10 ⁶ Btu/hr)	+172	+132
Steam (10 ⁶ Btu/hr)	-7.9	-5.5
Total Electricity Consumption (MW)	-14.2	-11.4
Coal Credit (tons/hr) ^a	12.7	9.3
Flue Gas Flow Rate (10 ⁶ acfm)	1.78	1.74
Makeup Sorbent Flow Rate (lbs/hr)	929	641
Molar Stoichiometry (Na/S removed)	3.31	3.27
Net Capacity (MW)	533	529
Number of Operating Trains	4	4
Required Removal Efficiency of Absorber		
NO _x	92.8	92.7
SO ₂	91.1	87.0
Actual (larger value of SO ₂ and NO _x) ^b	92.8	92.7
Sorbent Flow Rate (10 ⁶ lbs/hr)	1.33	0.92

^a This value is the sum of the steam requirements of the NOXSO process and acid plant divided by the boiler efficiency (88%), plus the energy sent to the boiler in the recycled gas. This net energy requirement is divided by the higher heating value of the coal and converted to tons/hr.

^b The SO₂ and NO_x removal efficiency are assumed to be equal, so the larger of these two values is used to determine the molar stoichiometry.

Figure 2-2: Sulfur Removal Efficiency versus Molar Stoichiometry

3. Sensitivity Analysis for the NOXSO Process

Beginning with the results shown above, a sensitivity analysis was performed to determine the effects of important parameters affecting the NOXSO process. The process parameters which were varied are the molar stoichiometry, sorbent attrition rate, regenerator efficiency, NO_x "destruction" efficiency, sorbent cost, methane cost, oil cost, and sulfuric acid cost. System effects of coal cleaning, air preheater size, and the use of a gas burner for the acid plant inlet gas also were examined. Since pressure drop data are not available for different bed height and flue gas velocities, these parameters were held constant.

3.1 Combustion of Acid Plant Gases

The default assumption for the NOXSO process is to use a gas burner for the regenerator, since it appeared in the EPRI report [8]. However, it was not clear if it was needed since it was never mentioned in previous reports by NOXSO. Using a gas burner for the regenerator gas is useful if there are sufficient combustible gases in the gas stream entering the acid plant. This is because:

- the energy in the combustible gases is released before the boiler and superheater so it can be captured and turned into steam
- any hydrogen or gases containing hydrogen are oxidized to water before the dryers so there is less moisture in the converter.
- the gas flow rate entering the converter is lower, so the capital and power costs are lower for the converter and any downstream equipment.

The disadvantages are additional capital and power costs for the gas burner equipment and a larger gas flow through the boiler, superheater, and gas humidification and cooling tower. This tends to increase the cost of these components, so the overall economics depends upon the composition of the gas. Table 3-1 summarizes the important performance and cost changes of a gas burner. If a gas burner is not used the acid plant capital cost doubles and the acid plant uses more than three times as much electricity while producing no steam. The total revenue requirement for the NOXSO process increases by 1.8 mills/kWh and the net capacity of the power plant decreases 15 MW. Thus, the advantages of using a gas burner for the gas entering the acid plant are substantial. If the washed coal is used, the magnitude of the effects are smaller, but it is still advantages to use a gas burner. Another feature of a gas burner is that it minimizes the effects of varying gas composition on the acid plant.

3.2 Air Preheater Effects

The NOXSO process recycles NO_x to the boiler via the gas stream used to heat the sorbent. This stream contains about 18% oxygen, which reduces by 37% the required combustion air entering the air preheater. Since the combustion air stream entering the air preheater is smaller, the

**Table 3-1: Effects of Gas Burner for Sulfuric
Acid Plant with the Unwashed Coal**

Parameter	With Gas Burner	No Gas Burner
Acid Plant		
Capital Cost (\$/kW)	37	72
Total Variable Cost (M\$/yr)	1.52	1.63
Utility Cost (M\$/yr)	0.43	1.43
Levelized Rev. Requirement (mills/kWh)	-1.72	-0.46
Acid Profit (\$/ton of acid)	31	8
Electricity Consumption (kW)	2030	6770
Steam Production (kW)	0	10,020
Total Process		
Capital Cost (\$/kW)	263	305
Levelized Rev. Requirement (mills/kWh)	14.4	16.1
Net Capacity (MW)	533	519

Table 3-2: Effects of the Air Preheater for the Unwashed Coal

Parameter	w/o NOXSO	Original Size	Larger Size
Flue Gas Flow Rate (acfm)			
Inlet Temperature (°F)	700	700	700
Exit Temperature (°F)	300	382	370
Combustion Air (10 ³ acfm)	1041	655	655
Inlet Temperature (°F)	80	80	80
Exit Temperature (°F)	515	634	660
Energy Flow Rate (10 ⁶ Btu/hr)			
Across Air Preheater	497	400	419
From Recycled Air	0	269	269
Total to Boiler	497	669	688
Coal Energy Credit (tons/hr) ^a	0	8.4	9.4
NOXSO Process			
Additional Capital Cost (\$/kW)	--	0	8.4
Coal Credit (M\$/yr)	--	1.63	1.82
Total Process			
Capital Cost (\$/kW)	--	263	271
Levelized Rev. Requ. (mills/kWh)	--	14.4	14.5
Net Capacity (MW)	--	533	535

^a This is the difference between the total energy flow to the boiler minus the energy flow to the boiler without the NOXSO process divided by the higher heating value of the coal and converted to tons/hr

air preheater cannot transfer as much heat, resulting in an exit temperature of the flue gas higher than the nominal value of 300°F. This results in an energy penalty for the NOXSO process. Building a larger air preheater would reduce the energy penalty; however, it is not possible to eliminate the energy penalty, since the combustion air stream is so much smaller. Table 3-2 summarizes the important performance and cost changes of building a larger air preheater, which raises the temperature of the combustion air to 660°F. Increasing the air preheater size increases the energy credit by 19 MBtu/hr or 1 ton/hr of coal. The larger air preheater costs 8.4 \$/kW, while increasing the coal credit by 0.2 M\$/yr. The effect on the total process including the cost of the fabric filter is to increase the total revenue requirement by 0.3 mills/kWh. For the washed coal the energy credit increases by 54 MBtu/hr or 2.5 tons/hr of coal. The capital cost increases to 12 \$/kW, while the total revenue requirement still increases by 0.3 mills/kWh. This value is within the error of the model, so it is uncertain whether a larger air preheater is economical. Air preheaters of different sizes also were examined, which resulted in the same conclusion: the change in cost was within the error of the model. Other options, such as a second economizer after the air preheater, for capturing more energy from the flue gas were not explored. However, if the flue gas temperature could be reduced to 300°F, an additional coal credit of 6.6 tons/hr could be achieved, which would save an additional 1.3 M\$/yr.

3.3 Other Process Parameters

The next sensitivity test was to multiply the nominal values (shown in Table 2-2) of the molar stoichiometry, attrition rate, sorbent cost, methane cost, oil cost, and acid price by 0.5, 0.75, 1, 1.5, and 2. Each parameter was varied individually while all other parameters were held constant. Figure 3-1 and 3-2 show the effects on the total capital and levelized revenue requirement of these parameters for the unwashed coal. Also plotted are the effects of a larger air preheater and not using a burner to show how they compare to variations in other parameters.

It is readily apparent that the most important parameter is the molar stoichiometry, since the both the capital and total revenue requirement change drastically with minor changes in the stoichiometry. The capital cost and revenue requirement increase by 13% and 32%, respectively, for a 50% increase in the molar stoichiometry. There are several reasons for this:

- The sorbent flow rate is directly proportional to the molar stoichiometry. Higher sorbent flow increases the electricity used for sorbent transportation and the size of all the equipment, except the absorber and acid plant.
- The gas quantity returned to the boiler is also proportional to the sorbent flow rate. This stream increases the flue gas flow rate entering the absorber and the fabric filter. The capital cost for the absorber increases slightly; doubling the molar stoichiometry increases the capital cost of the fabric filter by 7%.

- Most of the indirect costs, working capital, startup costs, and inventory capital are proportional to either the total process capital, which is proportional to the sorbent flow rate, or to the sorbent flow rate directly.
- Most of the variable costs are proportional to the sorbent flow rate either directly or indirectly. The only exceptions are the variable cost for methane and sulfuric acid sales. The oil cost increases since the heat required for the sorbent heater is proportional to the sorbent flow rate. Most of the fixed operating cost are related to the total process capital.

Figure 3-1: Capital Cost Sensitivity Analysis for Unwashed Coal

Figure 3-2: Total Revenue Requirement Sensitivity Analysis

The next most important parameters are the sorbent cost and the attrition rate. These two are closely related, since they affect the makeup sorbent cost, working capital, and startup cost in exactly the same manner. The only difference is that the sorbent cost affects the inventory capital while the attrition rate does not. This difference is shown most distinctly in Figure 3-1.

Of the remaining parameters (the price of methane, fuel oil, and sulfuric acid) the total revenue requirement is most sensitive to the price of oil and least sensitive to the price of methane. A 50% increase in the price of oil increases the total revenue requirement by 10%, while the same increase in methane cost, results in a 7% increase. These two parameters also affect the capital cost since they affect the startup and working capital. The price of sulfuric acid has no effect on the capital cost, while having a significant effect on the total revenue requirement. A 50% increase in the price of sulfuric acid decreases the total revenue requirement by 8.5%.

3.4 NO_x Reduction Efficiency

Figure 3-3 shows the capital and revenue requirements for different values of the NO_x being destroyed in the boiler. The NO_x "destruction" efficiency determines the NO_x removal efficiency required in the adsorber to meet the emission standard. As this value decreases the removal efficiency increases since there is a more NO_x entering the adsorber. Between destruction efficiencies of 50% and 80%, the NO_x removal efficiency in the adsorber is higher than the required SO₂ removal efficiency, so any change in this parameter changes the molar stoichiometry, which has a significant effect on the overall cost. Above a NO_x "destruction" efficiency of 80% the SO₂ removal efficiency is greater, so the molar stoichiometry is not affected. For the washed coal, the NO_x removal efficiency is always greater so the cost always decreases with improvements in the destruction efficiency.

3.5 Sorbent Regenerator Efficiency

Figure 3-4 shows the capital and revenue requirements versus the overall regeneration efficiency. The overall regeneration efficiency is a function of the efficiencies of both the regenerator and the steam treatment vessel:

$$\eta_{R\&S} = \eta_R + (1 - \eta_R)\eta_{STV} \quad (3.1)$$

where η_R = the efficiency of removing sulfur from the sorbent with methane in the regenerator vessel

η_{STV} = the efficiency of removing sulfur from the sorbent with steam in the steam treatment vessel

$\eta_{R\&S}$ = the overall regeneration efficiency of both vessels

To analyze the sensitivity of the NOXSO process to the overall regeneration efficiency, η_R and η_{STV} were given the following values: 55%, 60%, 70%, 80%, 90%, 95%. This produced the

overall efficiencies of 79.75%, 84%, 91%, 96%, 99%, and 99.75% shown plotted in Figure 3.4. The nominal value is 98.27%, which was estimated from an EPRI report [8]. The utilization of methane and steam were assumed to remain constant at their nominal values of 90% and 30%, respectively. The efficiency of the regenerator using methane was not set to 50%, because this would have created a gas composition which the acid plant was not designed to handle.² As can be seen, the capital and revenue requirement cannot be improved significantly, since the nominal value for the regeneration efficiency is quite high. However, if the regeneration efficiency drops to 91%, then the capital and revenue requirement increase by 2.7% and 6.7%, respectively.

4. Conclusion

From this analysis it is obvious that the most significant parameter affecting the design and cost of the NOXSO process is the sorbent flow rate needed to achieve a given removal efficiency. Therefore, the relationship between NO_x and SO_2 removal efficiencies, the molar stoichiometry and regeneration process should be studied further, since they affect the sorbent flow rate directly and have a high degree of uncertainty. Data on NO_x removal and destruction in the boiler is the most critical, since it is the most uncertain of any of the parameters. More experimental data is needed to confirm the analytical model for SO_2 removal efficiency; however, the results of the performance model are consistent with previous studies.

Experiments using a continuous regenerator should be performed since only batch test have been performed. These experiments should be designed to develop a performance curve relating methane and steam consumption to regeneration efficiency, and predict the regenerated gas composition. If a performance curve existed, the IEC model could be used to optimize the NOXSO process, since the tradeoffs between the sorbent flow rate, oil consumption, methane consumption, and acid plant are modeled explicitly.

The analysis of the air preheater showed that an additional that an energy credit worth 1.3 M\$/yr can be achieved if the temperature in the flue gas is reduced to 300°F. Other options for capturing this energy should be explored. A second economizer after the air preheater may be economical.

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² The inlet sulfur dioxide concentration was too low for this design, so the capital and operating cost would not be valid.

Figure 3-3: Effect of NO_x Destruction Efficiency in Boiler

Figure 3-4: Effect of Overall Regenerator Efficiency

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