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

Technical Progress Report

To

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From

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Table of Contents

1 Introduction	3
2 Sulfuric Acid Plant Model	4
3 Air Preheater	12
3.1 Base Case Air Preheater	13
3.2 New Air Preheater	15
3.3 Calculation of the Energy Credit	16
4 Effect of Air Preheater Size on Performance and Cost	17
5 Process Comparisons	21
6 Conclusion	21
I. IEC Conference Paper	22

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List of Figures

Figure 1: Monsanto Enviro-Chem Sulfuric Acid Plant	5
Figure 2: Regenerator Off-Gas Cooling Heat Exchangers	7
Figure 3: Air Preheater Model	14
Figure 4: Effect of Air Preheater Size on 36 Inch Bed Copper Oxide Process	18
Figure 5: Effect of Air Preheater Size on 46 Inch Bed Copper Oxide Process	19
Figure 6: Effect of Air Preheater Size on Downstream Fabric Filter	20

List of Tables

Table 1: Enthalpies of Species in Catalytic Converter
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12

NOMENCLATURE

English Letter Symbols

A_j	= area of device j (ft^2)
A_i	= regression constant for molar enthalpy (Btu/lbmole)
B_i	= regression coefficient for molar enthalpy ($\text{Btu/lbmole-}^\circ\text{F}$)
$c_{p,i}$	= specific heat for species i ($\text{Btu/lb-}^\circ\text{F}$)
C_i	= cost of species i (\$)
C_{AE}	= architect/engineer contractor cost (\$)
C_{CE}	= construction expense (\$)
C_{CF}	= construction contractor fee
C_{CT}	= process and project contingency (\$)
C_E	= engineering, design, and supervision (\$)
C_{IC}	= inventory capital cost (\$)
C_{IS}	= initial sorbent cost (\$)
C_{Int}	= interest during construction (\$)
C_L	= land cost (\$)
C_P	= incremental cost of air preheater (\$)
C_{PPC}	= pre-production cost (\$)
DCC_j	= direct capital cost for device j (\$)
$E_{j,k}$	= energy inlet or outlet k to device j (Btu/hr)
E_l	= Inlet energy to the absorber lower region (Btu/hr)
E_{Int}	= Inlet energy to the absorber upper region (Btu/hr)
FOC	= fixed operating cost (\$)
h_i	= enthalpy of species i (Btu/lb)
$\hat{h}_i$	= Molar enthalpy of species i (Btu/lbmole)
H	= fluidized bed expanded height (inches)
HR	= gross cycle heat rate (Btu/kWh)
$LMTD_j$	= log mean temperature difference of device j ($^\circ\text{F}$)
$m_{i,j,k}$	= mass flow of species i at the inlet or outlet k of device j (lb/hr)
m_1	= total gas flow into stage 1 of solids heater (lb/hr)
m_2	= total gas flow into stage 2 of solids heater (lb/hr)
m_{air}	= mass flow of air into stage 2 of the solids heater (Btu/hr)
m_{g1}	= mass flow of combustion air into stage 1 of solids heater (Btu/hr)
m_{g2}	= mass flow of combustion air into stage 2 of solids heater (Btu/hr)
m_s	= mass flow of sorbent through solids heater (lb/hr)
$m_{s,m}$	= required sorbent make-up (lb/hr)
$M_{i,j,k}$	= mass flow of species i at the inlet or outlet k of device j (lbmole/hr)
$M_{i,Int}$	= mass flow of species i from the lower to upper absorber regions (lbmole/hr)
$M_{CS,p}$	= copper sulfate (CuSO_4) produced in absorber (lbmole/hr)
$M_{1,i}$	= mass flow of species i out of stage 1 of solids heater (lbmole/hr)
$M_{FG,i}$	= flue gas flow entering absorber (lbmole/hr)
$M_{FG,o}$	= flue gas flow exiting absorber (lbmole/hr)
$M_{N,app}$	= approximate nitrogen flow (lbmole/hr)
$M_{st,SA}$	= steam generated in sulfuric acid plant (lbmole/hr)
$M_{TR,i}$	= air flow into regenerator to absorber transport line (lbmole/hr)
$M_{TR,o}$	= air flow out of regenerator to absorber transport line (lbmole/hr)
$M_{W,inj}$	= water injected as steam into the absorber with ammonia (lbmole/hr)
MW_i	= molecular weight of species i (lb/lbmole)

Q_j	= heat load across heat exchanger device j (Btu/hr)
R_{cs}	= absorber inlet copper to sulfur mole ratio (fraction)
R_{an}	= absorber inlet ammonia to NO_x mole ratio (fraction)
$T_{FG,i}$	= temperature of flue gas entering absorber ($^{\circ}F$)
$T_{j,k}$	= temperature of all species at inlet or outlet k of device j ($^{\circ}F$)
$T_{i,j,k}$	= temperature of species i at the inlet or outlet k of device j ($^{\circ}F$)
T_a	= temperature of sorbent entering solids heater ($^{\circ}F$)
T_c	= temperature of combustion gases entering solids heater ($^{\circ}F$)
T_{int}	= intermediate temperature between stages 1 and 2 of solids heater ($^{\circ}F$)
T_r	= temperature of sorbent entering regenerator ($^{\circ}F$)
TCC	= total capital cost (\$)
TDC	= total direct cost (\$)
TDV	= total plant investment (\$)
TIV	= total indirect investment (\$)
TPI	= total plant investment (\$)
U_j	= universal heat transfer coefficient for device j (Btu/hr-ft ² - $^{\circ}F$)
V_s	= superficial velocity of flue gas in fluidized bed (ft/s)
VOC	= variable operating cost, total (\$/yr)
VOC_i	= variable operating charge for species i (\$/yr)
VOC_{Coal}	= credit to copper oxide process for reduced plant coal consumption (\$/yr)
W_C	= weight percent of copper oxide in sorbent (fraction)

Greek Letter Symbols

ΔH_r^j	= heat of reactions in device j (Btu/hr)
ΔP	= pressure drop in fluidized bed (inches water)
η_{NO_x}	= NO_x removal efficiency, fraction
η_r	= regeneration efficiency, fraction
η_s	= sulfuric acid plant efficiency converting SO_2 to H_2SO_4 , fraction
η_{SO_2}	= SO_2 removal efficiency, fraction
σ_A	= standard deviation of regression constant A
σ_B	= standard deviation of regression constant B
Σ_m	= summation series m

Subscripts

Species:

1	= all gas entering stage 1 of solids heater
2	= all gas entering stage 2 of solids heater
air	= air
A	= ammonia, NH_3
Al	= aluminum oxide, Al_2O_3
BA	= boiler air
C	= copper, Cu
CD	= carbon dioxide, CO_2
CO	= copper oxide, CuO
CM	= carbon monoxide, CO
CS	= copper sulfate, $CuSO_4$
e	= leakage air in air preheater

FG	= flue gas
g	= flue gas in air preheater
g1	= combustion gas entering stage 1 of solids heater
g2	= combustion gas entering stage 2 of solids heater
H	= hydrogen, H_2
HI	= hydrochloric acid, HCl
M	= methane, CH_4
N	= nitrogen, N_2
NO	= nitrous oxide, NO
ND	= nitrogen dioxide, NO_2
O	= oxygen, O_2
s	= sorbent
st	= steam
SD	= sulfur dioxide, SO_2
ST	= sulfur trioxide, SO_3
SU	= sulfuric acid, H_2SO_4
W	= water, H_2O
wa	= weak acid

Equipment:

A	= fluidized bed absorber
AI	= ammonia injection
B	= boiler of sulfuric acid plant
BF	= blower and filter for sulfuric acid plant dilution air
CS	= catalytic converter contact section of sulfuric acid plant
D	= drying tower of sulfuric acid plant
ES	= effluent stripper for sulfuric acid plant
F	= general facilities for copper oxide plant
FG	= flue gas before entering fluidized bed
GC	= gas humidification and cooling tower of sulfuric acid plant
H	= heat recovery preheater of sulfuric acid plant
P	= air preheater
R	= regenerator
S	= superheater of sulfuric acid plant
SA	= sulfuric acid plant
SS	= strong acid handling system for sulfuric acid plant
SH	= solids heater
T	= interpass and final towers and associated equipment for sulfuric acid plant
TA	= pneumatic transport line, fluidized bed absorber to solids heater
TR	= pneumatic transport line, regenerator to fluidized bed absorber
TS	= pneumatic transport system
WA	= weak acid system of sulfuric acid plant

1 Introduction

This is the second 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 January 1, 1988 to March 31, 1988.

The general goal of this research project is to enhance and transfer to DOE a new computer simulation model for analyzing the performance and cost of integrated environmental control (IEC) systems for coal-fired power plants. A unique capability of this model is the probabilistic representation of uncertainty

in model parameters. This capability allows performance and cost to be quantified stochastically in comparing conventional technologies with advanced systems offering improved cost and/or effectiveness for SO_2 and NO_x removal. Several pre-combustion and post-combustion processes of interest to DOE have been selected for detailed modeling and analysis as part of this project.

During the past quarter, our efforts focused on improvements to previously developed analytical models of the DOE/PETC fluidized bed copper oxide process and the NOXSO process for combined SO_2/NO_x removal. Specifically, we have developed more detailed models of the sulfuric acid recovery system and plant air preheater since we have found that both of these components are important to overall system economics. Neither of these components had been treated rigorously in previous studies of the copper oxide or NOXSO process, however.

The sections below summarize progress to date on the development of a sulfuric acid plant model and a boiler air preheater model. An additional accomplishment during the past quarter was the conduct of new case studies employing these models in comparisons between the copper oxide process and a conventional power plant with wet flue gas desulfurization (FGD) and selective catalytic reduction (SCR). Results are reported in the final section of this report.

2 Sulfuric Acid Plant Model

Development of a sulfuric acid plant performance and economic model was initiated based on work performed by Monsanto for SMC Corporation under a previous DOE project.¹ The Monsanto plant was designed based on a regenerator off-gas composition specified by SMC. Based on the mass flows presented in the report, a performance model was derived.

Generally, the sulfuric plant must cool and remove water from the off-gas. Dilution air must be added to the off-gas to insure that the temperature in the catalytic converter, which catalyzes the reaction in which sulfur dioxide is converted to sulfur trioxide, does not exceed $1,158^\circ\text{F}$. The reactions in the catalytic converter are exothermic. The sulfur trioxide combines with water to form sulfuric acid.

The sulfuric acid plant configuration is shown in Figure 1. The off-gas from the regenerator must be cooled to about 500°F . First, the off-gas passes through a preheater to cool a converter section gas stream. Then the off-gas is further cooled by heating water into steam through the use of a boiler and a superheater.

A weak acid stream circulating through the humidification and cooling tower brings the off-gas temperature down to about 170°F and results in the removal of some water vapor. The off-gas is mixed with dilution air and the new gas stream enters the drying tower, where all the water vapor is condensed. The drying tower includes a cooling loop in which weak acid is circulated through the tower.

¹Ratcliffe-Brown, J., R. Hershey. *Economic Evaluation of the Fluidized Bed-Copper Oxide Flue Gas Treatment Process and Integrated Sulfuric Acid Plant*. Science Management Corporation, April, 1983.

Figure 1: Monsanto Enviro-Chem Sulfuric Acid Plant

Stream	1	2	4	5	6	7	8	9	11	14	16	17	18	19	20
Temp. F	824	609	506	169	120	85	115	226	450	163	120	171	603	779	1158
" Water	97			67	51	0	67	100	100		51	167		155	

Stream	21	22	23	24	25	26	27	28	29	30	31	3a	6a	8a	9a
Temp. F	1158	828	918	570	180	256	667	850	877	380	180	120	180	180	180
" Water				97				37			1				

The dilution air is required to prevent the temperature in the catalytic converter from exceeding its temperature rating, which is 1,158°F. Any combustible gases in the converter inlet gas stream will react in the first stage of the converter. Because combustible gas reactions are exothermic, the presence of combustible gases is a penalty to the sulfuric acid plant.

The converter section involves a number of interpass heat exchangers and cooling towers. The interpass and final cooling towers each have their respective acid cooling systems. The strong acid system includes pumps, tanks, and coolers.

The sulfuric acid plant model developed for the copper oxide process is meant to respond to a number of the factors that affect sulfuric acid plant cost. These include the volume flow of the off-gas stream, the amount of water vapor in the off-gas, the amount of combustible gases in the off-gas, and the temperature of the off-gas.

Because the converter section of the plant is complicated and all the component sizes are interrelated, it is desirable to model this part of the plant based on the volume flow rate of the gas stream entering the converter. This is a reasonable approximation assuming that the inlet gas is dry and contains some combustible gas, which will in general be true for the regenerator reductant gases under consideration for the copper oxide process. The intent of the model then is to determine the performance required to cool the off-gas, add the dilution air, and remove the water vapor from the combined off-gas and dilution air stream, and to estimate the gas flow rate entering the converter section.

The gas flow rate in the converter section is calculated based on the inlet off-gas, the required dilution air flow, and corrections for the removal of sulfur dioxide and oxygen due to the creation and removal of sulfur dioxide. The gas flow in the converter section is calculated to approximate the mass flow that enters the heat recovery preheater. The inlet flow to the catalytic converter is corrected for the loss of sulfur dioxide and oxygen due to reactions in the converter's first and second stages.

$$M_H = M_{D,o} - 0.95 M_{SD,R,o} - 0.475 M_{SD,R,o} \quad (1)$$

where:

$$M_{D,o} = M_{M,R,o} + M_{H,R,o} + M_{CM,R,o} + M_{SD,R,o} + M_{CD,R,o} + M_{air,SA} \quad (2)$$

The gas stream leaving the dryer contains no water vapor.

Schematic diagrams for the heat recovery preheater, boiler, and superheater are shown in Figure 2. The heat load in the heat recovery preheater is proportional to the converter side gas flow, which in the SMC Monsanto design is 4,323 lbmole/hr. For the Monsanto design, the heat load is 2,400,000 btu/hr. The heat load in the heat recovery preheater, adjusted for the converter side mass flow rate, is given by:

$$Q_H = 2,400,000 \left(\frac{M_H}{4,323} \right) \quad (3)$$

Figure 2: Regenerator Off-Gas Cooling Heat Exchangers

a) Heat Recovery Preheater

b) Superheater

b) Boiler

The exit temperature of the off-gas from the heat recovery preheater is given by a simple energy balance.

$$\Sigma_A = \sum_{k=1}^6 M_{k,R,o} \hat{h}_k \quad (4)$$

$$\Sigma_B = \sum_{k=1}^6 M_{k,R,o} \hat{\lambda}_k \quad (5)$$

$$\Sigma_C = \sum_{k=1}^6 M_{k,R,o} \hat{b}_k \quad (6)$$

$$T_{H,o} = \left(\frac{\Sigma_A - Q_H - \Sigma_B}{\Sigma_C} \right) \quad (7)$$

where k represents the species in the off-gas stream, M, H, CM, CD, W, and SD.

The factor which will most directly affect the cost of the heat exchanger is the required heat transfer surface area. This is calculated based on knowledge of the heat transfer load, the heat transfer coefficient U, and the log mean temperature difference (LMTD) for the heat exchanger. The heat transfer coefficient is taken from the Monsanto estimate. It is assumed that the converter side gas flow enters at 180°F and leaves at 256°F as in the Monsanto estimate.

$$U_H = 4.5 \frac{Btu}{(hr-ft^2-^{\circ}F)} \quad (8)$$

$$LMTD_H = \frac{(T_{R,out} - 256^{\circ}F) - (T_{H,o} - 180^{\circ}F)}{\ln\{(T_{R,out} - 256^{\circ}F)/(T_{H,o} - 180^{\circ}F)\}} \quad (9)$$

$$A_H = \frac{Q_H}{U_H LMTD_H} \quad (10)$$

The calculations for the boiler and superheater are similar, except that first the amount of water that can be heated to steam must be determined, and then the calculations for the LMTD and heat transfer areas of the boiler and superheater can be made. In order to parallel the design work performed by Monsanto, the off-gas temperature leaving the boiler is specified as 500°F. It is assumed, as in the Monsanto study, that water is available at 226°F to be heated to steam at 450°F and 100 psia.

A simple energy balance determines the amount of water that can be converted to steam under these conditions. Then the boiler and superheater energy loads can be determined, and after an energy balance calculation, the boiler inlet temperature can be determined. From this, the LMTD for both the superheater and the boiler can be calculated. Finally, the required heat transfer areas for both heat

exchangers are calculated from knowledge of the heat transfer coefficient, heat load, and LMTD.

The water mass flow rate required for conversion to steam is calculated by determining how much energy is available in the hot regenerator off-gas to heat water from 226°F to superheated steam at 450°F and 100 psia.

$$\Sigma_D = \sum_{k=1}^6 M_{k,R,o} (h_{k,T_{H,o}} - h_{k,500}) \quad (11)$$

$$M_{st} = \frac{\Sigma_D}{19,068 \text{ BTU/lbmole}} \quad (12)$$

The heat loads in the boiler and superheater can then be calculated, assuming, as in the Monsanto design, that the boiler will heat the water to steam at a temperature of 338°F and a pressure of 100 psia. The heat loads for these two heat exchangers are calculated based on the water mass flow and difference in outlet and inlet enthalpies of the water streams.

$$Q_{B,SA} = M_{st,SA} (21,380 - 3,497) \quad (13)$$

$$Q_{S,SA} = M_{st,SA} (22,565 - 21,380) \quad (14)$$

The inlet temperature of the regenerator off-gas stream to the boiler is calculated by an energy balance on the boiler, assuming the off-gas exit temperature to be 500°F.

$$\Sigma_E = \sum_{k=1}^6 M_{k,R,o} (h_{k,500} - \hat{A}_k) \quad (15)$$

$$\Sigma_F = \sum_{k=1}^6 M_{k,R,o} \hat{B}_k \quad (16)$$

$$T_{B,I,SA} = \frac{Q_{B,SA} + \Sigma_E}{\Sigma_F} \quad (17)$$

The log mean temperature differences for the boiler and superheater are based on the inlet and outlet temperatures of the two heat exchangers. The required heat transfer surface area for each is calculated based on the heat load, the log mean temperature difference, and the universal heat transfer coefficient. The values for the heat transfer coefficients are adopted from the Monsanto estimate.

$$U_B = 15 \frac{\text{Btu}}{(\text{hr-ft}^2\text{-}^\circ\text{F})} \quad (18)$$

$$\text{LMTD}_B = \frac{(T_{B,I,SA} - 338^\circ\text{F}) - (500^\circ\text{F} - 226^\circ\text{F})}{\ln\{(T_{B,I,SA} - 338^\circ\text{F})/(500^\circ\text{F} - 226^\circ\text{F})\}} \quad (19)$$

$$U_B = 6 \frac{Btu}{(hr-ft^2-^{\circ}F)} \quad (20)$$

$$LMTD_S = \frac{(T_{H,o} - 450^{\circ}F) - (T_{B,I,SA} - 338^{\circ}F)}{\ln\{(T_{H,o} - 450^{\circ}F)/(T_{B,I,SA} - 338^{\circ}F)\}} \quad (21)$$

$$A_{B,SA} = \frac{Q_{B,SA}}{U_B LMTD_B} \quad (22)$$

$$A_{S,SA} = \frac{Q_{S,SA}}{U_S LMTD_S} \quad (23)$$

where k represents the species in the off-gas stream, CH₄, H₂, SO₂, H₂O, CO, and CO₂

The off-gas enters the cooling and humidification tower at a temperature of 500°F and leaves at 170°F. It is assumed that about 30 percent of the water vapor is removed based on the Monsanto design. Cooling is effected by the circulation of a weak acid stream through the cooling tower. The mass flow of this stream is calculated by an energy balance.

$$\Sigma_G = \sum_{k=1}^6 M_{k,R,o} h_{k,500} \quad (24)$$

$$\Sigma_H = \sum_{j=1}^5 M_{j,R,o} h_{j,170} \quad (25)$$

$$\Sigma_I = M_{W,R,o} \{(0.3(17,932) - 0.3(1,654) - 0.7(716))\} \quad (26)$$

$$M_{wa,GC} = \frac{\Sigma_G - \Sigma_H + \Sigma_I}{324.6} \quad (27)$$

where:

k represents the species in the off-gas stream, CH₄, H₂, SO₂, H₂O, CO, and CO₂

j represents the species CH₄, H₂, SO₂, CO, and CO₂

The dilution air stream is determined by an energy balance on the first stage of the catalytic converter. Seventy percent of the inlet sulfur dioxide is assumed converted to sulfur trioxide based on the typical performance of a three stage converter. In addition, all combustible gases present, such as methane, hydrogen, and carbon monoxide, are assumed to react completely. Therefore, the energy balance takes into account the exothermic reactions due to the conversion of sulfur dioxide and combustible gases. The

gas stream enters the converter at 779°F, and leaves at the maximum allowable temperature of 1,158°F.

The inlet gas flow into the first stage of the catalytic converter includes all the methane, hydrogen, carbon monoxide, carbon dioxide, and sulfur dioxide present in the regenerator off-gas and the oxygen and nitrogen added as part of the dilution air. Because the off-gas and dilution air pass through a drying tower before entering the converter, no water vapor is present.

The chemical reactions that occur in the first stage of the catalytic converter are as follows:

From the stoichiometry of the chemical reactions occurring in the first stage of the catalytic converter, the outlet mass flow from the first stage can be calculated.

$$M_{M,CS,o} = 0 \quad (32)$$

$$M_{H,CS,o} = 0 \quad (33)$$

$$M_{CM,CS,o} = 0 \quad (34)$$

$$M_{SD,CS,o} = 0.3 M_{SD,R,o} \quad (35)$$

$$M_{ST,CS,o} = 0.7 M_{SD,R,o} \quad (36)$$

$$M_{CD,CS,o} = M_{CD,R,o} + M_{M,R,o} + M_{CM,R,o} \quad (37)$$

$$M_{W,CS,o} = 2 M_{M,R,o} + M_{H,R,o} \quad (38)$$

$$M_{O,CS,o} = 0.21 M_{air,SA} - 2 M_{M,R,o} - 0.5 M_{CM,R,o} - 0.5 M_{H,R,o} - 0.35 M_{SD,R,o} \quad (39)$$

$$M_{N,CS,o} = 0.79 M_{air,SA} \quad (40)$$

An energy balance for the first stage of the converter is used to determine the required flow of dilution air on a dry basis. All species enter the first stage at 779°F and exit at 1,158°F. The converter is assumed to be adiabatic for this calculation. The increase in enthalpy for air over this temperature range is 2,875 Btu/lbmole. The enthalpies used for all species are shown in Table 1.

The heat of reaction for the first stage of the converter is calculated from the heats of reaction for Equations (28), (29), (30), and (31).

$$\Sigma_J = M_{SD,CS,o}(12,323) + M_{ST,CS,o}(16,800) + M_{CD,CS,o}(11,660) + M_{W,CS,o}(9,227) \quad (41)$$

Table 1: Enthalpies of Species in Catalytic Converter

Compound	779°F Btu/lbmole	1,158°F Btu/lbmole
Carbon Dioxide	7,484	11,660
Carbon Monoxide	5,045	--
Hydrogen	4,912	--
Methane	7,733	--
Nitrogen		
Oxygen		8,132
Sulfur Dioxide	7,699	12,323
Sulfur Trioxide	--	16,800
Water Vapor	--	9,227

$$\Sigma_K = [-2 M_{M,R,o} - 0.5 M_{CM,R,o} - 0.5 M_{H,R,o} - 0.35 M_{SD,R,o}] 8,132 \quad (42)$$

$$\Sigma_L = M_{M,R,o}(7,733) + M_{H,R,o}(4,912) + M_{CM,R,o}(5,045) + M_{SD,R,o}(7,699) + M_{CD,R,o}(7,484) \quad (43)$$

$$\Delta H_r^{CS} = -42,600(0.7)M_{SD,R,o} - 345,700M_{M,R,o} - 104,300M_{H,R,o} - 121,700M_{CM,R,o} \quad (44)$$

$$M_{air,SA} = \frac{\Sigma_J + \Sigma_K - \Delta H_r^{CS} - \Sigma_L}{-2,875} \quad (45)$$

The gas flow into the drying tower is given by the sum of the dilution air and the off-gas. Assuming the air is about 3.6 percent water vapor, this mass flow is given by:

$$M_{D,i} = M_{SD,R,o} + M_{CD,R,o} + M_{M,R,o} + M_{H,R,o} + M_{CM,R,o} + 0.7 M_{W,R,o} + 1.036 M_{air,SA} \quad (46)$$

These equations provide sufficient information to estimate the sulfuric acid plant cost. Completion of the sulfuric acid plant model, including its extension to the NOXSO process and a detailed numerical example, is planned for the next calendar quarter of this project.

3 Air Preheater

The copper oxide process alters the flue gas flow in two ways. First, it increases the temperature of the flue gas because of the exothermic reactions occurring in the fluidized bed, and, second, it slightly increases the volume flow rate of the gas because of the addition of the solids heater and pneumatic transport gases to the fluidized bed. Of these two, the increase in flue gas temperature is the most important; for a high sulfur coal, the flue gas temperature may increase over 100°F. This will result in an

energy credit to the boiler through the preheater.

The air preheater preheats the air entering the boiler. Generally in older power plants this is a shell and tube type heat exchanger. Typically there is a large amount of leakage between the inlet boiler air and the outlet flue gas. The model presented here assumes that the mixing due to the leakage air occurs independently of the heat transfer across the heat exchanger. This allows separate consideration of the effect of the leakage air and of the characteristics of the heat exchanger. The separation of the mixing and heat transfer is shown in Figure 3

Because the copper oxide performance and economic model is a module of the overall power plant system, it is necessary to look at the incremental costs associated with changing the size of the preheater. These costs are charged to the copper oxide process. Two cases are considered: zero incremental cost due to using the same size preheater designed as if the copper oxide process were not present, and the incremental cost associated with using a bigger preheater to take full advantage of the energy credit offered by the copper oxide process.

For both cases, similar definitions of variables apply. All variables are shown in Figure 3. Three mass flows and five temperatures are defined. The heat exchanger is assumed to be a counterflow type.

The air flow required at the boiler, the leakage in the preheater, and the air flow leaving the copper oxide process are inputs to the air preheater models, and these values are known based on the copper oxide and power plant models. The specific heats of air and flue gas are taken to have average values for simplicity of calculation. The air is assumed to enter at 80°F, and the flue gas enters the preheater at the same temperature as the exit temperature from the copper oxide fluidized bed absorber.

For the base case air preheater, the original exit temperature of the flue gas is 300°F, and the original exit temperature of the boiler air is 515°F. The new preheater is sized to maintain the flue gas exit at 300°F, causing an increase in the boiler air temperature. In both cases, there will be an increase in thermal energy supplied to the boiler.

For the base case air preheater, the flue gas exit and boiler inlet temperatures will increase because of the higher temperature of the flue gas due to the copper oxide process. For the new air preheater, the flue gas exit will be the same as for the base case without the copper oxide process, but the air temperature entering the boiler will be significantly higher than for the base case preheater without the copper oxide process.

3.1 Base Case Air Preheater

The equations in this section describe the base case air preheater, which is sized as if the copper oxide process were not present, as in the case of an existing plant.

First, an exponential factor D is calculated, using Equation (47), based on the mass flows in the heat exchanger. Then the uncorrected outlet temperature of the flue gas, $T_{FG,P,unc}$, the outlet temperature of

Figure 3: Air Preheater Model

the boiler air, $T_{BA,P,o}$, and the corrected outlet temperature of the flue gas, $T_{FG,P,o}$, can be calculated using Equations (49), (50), and (51), respectively. Finally, the heat credit to the boiler, q_c , can be calculated, using Equation (52), based on the incremental change in the amount of heat transported by the air stream to the boiler. For this configuration, there is no change in the cost of the air preheater.

$$D = UA \left(\frac{1}{w_g C_{p,g}} - \frac{1}{w_a C_{p,a}} \right) \quad (47)$$

$$E = \frac{w_g C_{p,g}}{w_a C_{p,a}} \quad (48)$$

$$T_{FG,P,o,unc} = T_{FG,o} - (T_{FG,o} - T_{BA,P,i}) \frac{1 - e^D}{E - e^D} \quad (49)$$

$$T_{BA,P,o} = \frac{w_g C_{p,g} (T_{FG,o} - T_{FG,P,o,unc})}{w_a C_{p,a}} + T_{BA,P,i} \quad (50)$$

$$T_{FG,P,o} = \frac{w_a C_{p,a} T_{BA,P,i} + w_g C_{p,g} T_{FG,P,o,unc}}{w_g C_{p,g} + w_a C_{p,a}} \quad (51)$$

$$q_c = \frac{w_a C_{p,a} (T_{BA,P,o} - 515^\circ F)}{HR} \quad (52)$$

3.2 New Air Preheater

The new air preheater is sized to accommodate the new flue gas flow and temperature due to the copper oxide process, while maintaining a flue gas exit temperature of 300°F. The cost of the heat exchanger is driven primarily by the product UA.

The uncorrected flue gas exit temperature is calculated based on an energy balance with the leakage air. In this case, the corrected outlet flue gas temperature $T_{FG,P,o}$ is specified.

$$T_{FG,P,o,unc} = \frac{w_a C_{p,a} (T_{FG,P,o} - T_{BA,P,i})}{w_g C_{p,g}} + T_{FG,P,o} \quad (53)$$

The boiler air exit temperature from the preheater is calculated based on an energy balance with the flue gas. The flue gas air preheater inlet temperature is known because it is the same as the flue gas fluidized bed absorber outlet temperature, and the uncorrected flue gas air preheater exit temperature is calculated using Equation (53).

$$T_{BA,P,o} = \frac{w_g C_{p,g} (T_{FG,P,i} - T_{FG,P,o,unc})}{w_a C_{p,a}} + T_{BA,P,i} \quad (54)$$

The log mean temperature difference is calculated from the heat exchanger inlet and outlet temperatures.

$$\Delta T_{lm} = \frac{(T_{FG,P,i} - T_{BA,P,o}) - (T_{FG,P,o,unc} - T_{BA,P,i})}{\ln \{(T_{FG,P,i} - T_{BA,P,o}) / (T_{FG,P,o,unc} - T_{BA,P,i})\}} \quad (55)$$

The heat load across the heat exchanger must be calculated using Equation (56) in addition to the log mean temperature difference in order to estimate the required product of heat transfer coefficient and surface area required for the heat exchanger, calculated using Equation (57).

$$q = w_g c_{p,g} (T_{FG,P,i} - T_{FG,P,o,unc}) \quad (56)$$

$$UA = \frac{q}{\Delta T_{lm}} \quad (57)$$

Finally, the energy credit associated with adding thermal energy to the boiler inlet air is calculated by comparing the energy added to the boiler air for the new air preheater with that for the base case air preheater.

$$q_c = \frac{w_a c_{p,a} (T_{BA,P,o} - 515^\circ F)}{HR} \quad (58)$$

3.3 Calculation of the Energy Credit

The energy credit recoverable through the air preheater due to the copper oxide process ranges from about 1 percent to 5 percent of gross plant output based on a coal sulfur content range from about 2.2 percent sulfur to 4 percent sulfur, assuming a 90 percent sulfur dioxide removal efficiency. The energy credit varies with removal efficiency and bed height; the credit increases with an increase in sulfur dioxide removal efficiency or a decrease in absorber bed height. A decrease in bed height requires an increase in sorbent flow rate for a given removal efficiency, in turn requiring increased sorbent heating. The energy credit from the copper oxide process offsets some of the power plant coal requirement compared to a plant of the same gross size without the copper oxide process. In this work, the power plant capital and operating costs are calculated for a fixed gross output. The copper oxide process is charged with the incremental capital cost of a new air preheater, but is credited with a reduction in plant coal consumption on an annual basis.²

²Alternatively, the energy credit can be viewed as an increment to gross plant capacity. In this case the copper oxide process is charged with the full cost of the incremental capacity (e.g., the difference between, say, a 500 MW plant and 522 MW plant), then credited with the sales revenue from the incremental electricity generation (at the marginal or other specified rate). This method will lead to higher capital costs but lower annualized costs relative to credits based on fuel savings alone.

4 Effect of Air Preheater Size on Performance and Cost

The energy recoverable to the boiler from the exothermic reactions in the absorber through the air preheater affects power plant operating cost by reducing the amount of coal the plant must burn to maintain the same gross output. Though all other performance parameters are calculated based on the actual system configuration, the coal consumption of the power plant is calculated as if the energy credit from the copper oxide process were not available. An annual credit is then calculated based on subtracting from the plant coal consumption the actual consumption, taking into consideration the copper oxide energy credit. This annual coal consumption credit is attributed to the copper oxide process variable cost. Therefore, for a larger recovery of energy through the air preheater, the credit to the copper oxide process for reduced coal consumption increases.

Figures 4 and 5 show the electricity credit, capital cost, and annual cost for the copper oxide process for 36 and 46 inch bed heights respectively. For each bed height, the parameters are shown for both 90% and 70% sulfur dioxide removal with both a "new" and "old" air preheater. The "old" preheater is a preheater sized as if the copper oxide process were not present; the temperature of the flue gas exiting the "old" preheater is roughly 380°F for a high sulfur coal. A "new" preheater is sized to take full advantage of the energy credit from the copper oxide process; the temperature of the flue gas exiting the "new" preheater is fixed at 300°F. For all cases, the "new" air preheater is more expensive than the old, entailing increased capital costs. However, for both bed heights, the larger air preheater allows a substantially larger recovery of energy than the old preheater. The differences are slightly more pronounced for the 36 inch bed height, because the sorbent flow rate and the flue gas temperature exiting the absorber are slightly higher than for the 46 inch bed height. The differences in energy credit between the 70% and 90% removal efficiencies for the same bed heights are attributable to the same reasons. Because of the substantial difference in energy credit for the two air preheaters, the revenue requirements for all cases are lower for the larger preheater. The increased energy credit for the larger air preheater results in a decrease in the amount of coal that must be consumed to produce the same gross plant output, which is a variable operating cost credit to the copper oxide process. Lower coal consumption in turn means that there is less sulfur dioxide in the flue gas, requiring a lower sorbent flow rate and a correspondingly less expensive copper oxide process plant.

An additional consideration is the effect of the air preheater size on downstream equipment, most notably particulate control devices. Figure 6 shows the effect of new and old preheater size on the capital cost of a fabric filter device. The fabric filter is less expensive for the larger preheater because the exit temperature of the flue gas from the preheater is lower, resulting in a lower volume flow rate. The flue gas volume flow rate is a key physical parameter that governs the cost of a fabric filter. From a system perspective, it is apparent that the larger air preheater, though representing an increase in capital cost of the copper oxide process, will reduce the capital cost of downstream equipment and will reduce the annual cost for the copper oxide process. On a plant level, the capital cost of the power plant and pollution control equipment is slightly higher for the new preheater configuration by about \$3/kw, but the revenue requirement is about 0.7 mills/kwhr less for the case considered in Figure 6.

Figure 4: Effect of Air Preheater Size on 36 inch Bed Copper Oxide Process

Figure 5: Effect of Air Preheater Size on 46 inch Bed Copper Oxide Process

Figure 6: Effect of Air Preheater Size on Downstream Fabric Filter

5 Process Comparisons

A series of case studies comparing the fluidized bed copper oxide to a conventional plant with FGD plus SCR was carried out to illustrate the application of the IEC computer model using the newly developed plant component models. Details of the case studies are reported in a technical paper^{*} invited for presentation at the 4th Symposium on Integrated Environmental Control (co-sponsored by EPRI, ASME and APCA), and attached to this report as Appendix A. Briefly, the results showed that for a high sulfur (4%S) coal, the economic benefits of the copper oxide process were greatest when combined with physical coal cleaning. This analysis underscores the importance of considering integrated system designs to identify the most cost-effective applications of new clean coal technology.

6 Conclusion

Efforts during the coming calendar quarter will focus on completion and detailed documentation of the process models described in this report, including extension of the sulfuric acid plant model to the NOXSO process gas stream. Progress on documentation of the model computer code (user's manual) also will be reported along with other developments related to model transfer.

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