Near-Zero Emissions Oxy-Combustion Flue Gas Purification – Power Plant Performance

Topical Report

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Abstract

A technical feasibility assessment was performed for retrofitting oxy-fuel technology to an existing power plant burning low sulfur PRB fuel and high sulfur bituminous fuel. The focus of this study was on the boiler/power generation island of a subcritical steam cycle power plant. The power plant performance in air and oxy-firing modes was estimated and modifications required for oxy-firing capabilities were identified.

A 460 MWe (gross) reference subcritical PC power plant was modeled. The reference air-fired plant has a boiler efficiency (PRB/Bituminous) of 86.7%/89.3% and a plant net efficiency of 35.8%/36.7%. Net efficiency for oxy-fuel firing including ASU/CPU duty is 25.6%/26.6% (PRB/Bituminous). The oxy-fuel flue gas recirculation flow to the boiler is 68%/72% (PRB/bituminous) of the flue gas (average O₂ in feed gas is 27.4%/26.4%v (PRB/bituminous)). Maximum increase in tube wall temperature is less than 10°F for oxy-fuel firing.

For oxy-fuel firing, ammonia injected to the SCR was shut-off and the FGD is applied to remove SO_x from the recycled primary gas stream and a portion of the SO_x from the secondary stream for the high sulfur bituminous coal.

Based on CFD simulations it was determined that at the furnace outlet compared to air-firing, SO₃/SO₂ mole ratio is about the same, NO_x ppmv level is about the same for PRB-firing and 2.5 times for bituminous-firing due to shutting off the OFA, and CO mole fraction is approximately double.

A conceptual level cost estimate was performed for the incremental equipment and installation cost of the oxyfuel retrofit in the boiler island and steam system. The cost of the retrofit is estimated to be approximately 81 M\$ for PRB low sulfur fuel and 84 M\$ for bituminous high sulfur fuel.

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Executive Summary

A technical feasibility assessment was carried out for retrofitting oxy-fuel technology to an existing power plant burning low sulfur PRB fuel and high sulfur bituminous fuel. The focus of this study was on the boiler/power generation island of a subcritical steam cycle power plant. The power plant performance in air and oxy-firing modes was estimated and modifications required for oxy-firing capabilities were identified. There appears to be no major obstacles to oxyfuel retrofitting the reference plant presented herein; however, specific constraints and limitations must be evaluated by a detailed design.

A 460 MWe (gross) reference subcritical PC power plant was modeled using Aspen-Plus® to model the cycle and FW-FIRE to model the furnace. An air-fired simulation was run at full load. The reference air-fired plant has a boiler efficiency (PRB/Bituminous) of 86.7%/89.3% and a plant net efficiency of 35.8/36.7%.

The plant was modified for oxyfuel firing with the addition of flue gas recirculation ducts, oxygen distribution piping, steam extractions for ASU/CPU, added flue gas cooler for CPU, increased size of lower economizer in HRA, a low pressure economizer downstream of the air-heater, and a quench tower after the FGD, as well as increased cooling duty. Net efficiency including ASU/CPU duty is 25.6%/26.6% (PRB/Bituminous). CO₂ capture efficiency of the CPU is 99%.

For oxy-fuel firing, ammonia injected to the SCR was shut-off, but the SCR catalyst remained in place for operational flexibility. Since the coal mill requires low moisture nearly SO_x-free gas for proper performance and avoidance of acid corrosion, the FGD is applied to remove SO_x from the recycled primary gas stream. To limit the boiler corrosive SO_x concentration for the high sulfur bituminous fuel, 36% of the recycled secondary gas was also treated by the FGD. SO₂ in the boiler is predicted to be 500 ppm for PRB and 4080 ppmv and 5690 ppmv with and without the secondary gas treatment for bituminous. Gas to the CPU was extracted before the FGD.

The flue gas recirculation flow to the boiler is 68%/72% (PRB/bituminous) of the flue gas (average O₂ in feed gas is 27.4%/26.4%v (PRB/bituminous)). Maximum increase in tube wall temperature is less than 10°F for oxy-fuel firing.

Based on CFD simulations it was determined that at the furnace outlet compared to air-firing, SO₃/SO₂ mole ratio is about the same, NO_x ppmv level is about the same for PRB-firing and 2.5 times for bituminous-firing due to shutting off the OFA, and CO mole fraction is approximately double. Performance and emissions are summarized in Table 1.

A conceptual level cost estimate was performed for the incremental equipment and installation cost of the oxyfuel retrofit in the boiler island and steam system. The cost of the retrofit is estimated to be approximately 81 M\$ for PRB low sulfur fuel and 84 M\$ for bituminous high sulfur fuel.

Table 1 – Summary of Performance and Emissions

		low S PRB		High S Bit	
		Air-fired	O2-fired	Air-fired	O2-fired
Fuel Flow	klb/h	449	452	331	334
Air Flow	klb/h	3585	0	3567	0
Oxygen Flow	klb/h	0	725	0	712
Recirculated. Flue Gas	%	0.0%	67.7%	0.0%	72.5%
Limestone	klb/h	2.14	1.35	27.2	17.1
Boiler efficiency	%	86.7	89.4	89.3	92.4
Gross Power	MWe	461	467	460	465
Aux. Power	MWe	43	38	45	40
ASU/CPU Power	MWe		125.9		119.7
Extra Cooling Water Power	MWe		2.5		2.3
Net Power	MWe	418	301	416	304
Net efficiency	%	35.8	25.6	36.7	26.6

low Sulfur PRB						
	Air-Fired at Stack			O2-Fired to CPU		
	ppmv	lb/h	lb/Mbtu	ppmv	lb/h	lb/Mbtu
CO	128	536	0.13	280	244	0.06
SO2	52	501	0.13	465	928	0.23
SO3	1.3	15.8	0.0040	5.8	14.4	0.0036
NOx	36	247	0.062	156	223	0.056
NH3	0.7	1.5	0.00037	0	0	0
HCl	0.1	0.8	0.00020	18	20.9	0.0052
PM		49	0.012		49	0.012
VOC	1.8	11.9	0.0030	1.3	1.8	0.0004
Hg	ppbv	lb/h	lb/Tbtu	ppbv	lb/h	lb/Tbtu
	1.7	0.051	12.8	10.1	0.063	15.7

	High Sulfur Bit.								
	Air-Fired at Stack			O2-Fired to CPU					
							Reduced SOx		
	ppmv	lb/h	lb/Mbtu	ppmv	lb/h	lb/Mbtu	ppmv	lb/h	lb/Mbtu
CO	133	538	0.14	282	215	0.06	284	205	0.05
SO2	35	328	0.08	5349	9349	2.40	3837	6326	1.62
SO3	2.4	28.0	0.0073	66.7	145.7	0.0374	47.4	97.7	0.0251
NOx	40	267	0.069	388	522	0.134	391	496	0.127
NH3	0.6	1.6	0.00041	0	0	0	0	0	0
HCl	3.7	19.4	0.00503	559.7	556.7	0.143	402.1	377.3	0.097
PM		79	0.020		70	0.018		66	0.017
VOC	1.4	9.1	0.0024	1.1	1.3	0.0003	1.2	1.3	0.0003
	ppbv	lb/h	lb/Tbtu	ppbv	lb/h	lb/Tbtu	ppbv	lb/h	lb/Tbtu
Hg	0.2	0.004	1.1	1.1	0.006	1.6	1.0	0.005	1.3

Introduction

Foster Wheeler (FW) was subcontracted to support Praxair's effort of developing CO₂ purification technology for existing power plants retrofitted with oxy-fuel fired capability. FW's objective is to perform a technical/economic feasibility study to determine retrofit performance, cost, and interface conditions with the CPU and ASU. To accomplish this objective FW developed a computer simulation process model of the air-fired and oxyfuel-fired pulverized coal (PC) boiler system.

FW is to provide Praxair with plant heat and material balances, estimates of concentrations of trace impurities in the flue gas, a list of oxygen-fired plant modifications and constraints, and an economic assessment of the boiler retrofit. Two fuels were simulated: 1) low sulfur PRB and 2) high sulfur bituminous. This topical report covers the results of both the low sulfur PRB and high sulfur bituminous fuel simulations as well as the economic assessment of the boiler retrofit for both fuels.

Approach

A system level model and 3-D CFD model of a reference subcritical PC power plant firing either low-sulfur PRB or high-sulfur bituminous coal was created. The Aspen-Plus platform was used to simulate the power plant and the FW-FIRE CFD code was used to determine furnace performance. Simulations were performed at full load (Boiler Maximum Continuous Rating). The reference plant is an existing 460 MWe (gross) power plant.

Design Basis and Assumptions

Site ambient conditions are listed in Table 2.

Table 2 – Site Conditions

Elevation	ft	7000
Ambient Pressure	psia	11.3
Ambient Dry Bulb	F	87
Ambient Wet Bulb	F	61
Ambient Humidity	%	26
Condenser Pressure	in Hg	2.75
Cooling Water Temp.	F	76

Coal analysis is presented in Table 3.

Table 3 – Coal Analysis: Low Sulfur and High Sulfur Coals

		PRB		Bituminous	
		wet	dry	wet	dry
C	%w	51.47	70.60	63.75	71.73
H	%w	3.43	4.71	4.50	5.06
O	%w	12.73	17.46	6.88	7.74
N	%w	0.66	0.91	1.25	1.41
S	%w	0.20	0.27	2.51	2.82
Ash	%w	4.40	6.04	9.70	10.91
Moisture	%w	27.10	0.00	11.12	0.00
He	%w	0.01	0.01	0.29	0.33
sum	%w	100.00	100.00	100.00	100.00
HHV	Btu/lb	8870	12167	11666	13126
Hg	ppmdw	0.8		0.09	

The reference plant selected for this study uses a subcritical steam cycle. Steam cycle conditions are listed in Table 4.

Table 4 – Steam Conditions

<u>Main Steam</u>		
Pressure	psia	2415
Temperature	F	1000
Flow Rate	klb/h	2990
<u>Reheat Steam</u>		
Pressure	psia	600
Temperature	F	1000
Flow Rate	klb/h	2630

The reference power plant selected employs a natural circulation boiler, selective catalytic reduction (SCR) for reducing NO_x, a wet flue gas desulfurization (FGD) for capturing SO_x, and a baghouse to remove particulates. The furnace has a total of 24 opposed wall-fired low NO_x Vortex series split-flame burners (3 vertical x 4 horizontal x 2 walls) and 10 over-fire air ports. The furnace heat transfer surfaces consist of waterwalls, constructed from MONO-WALL™ finned-tube construction for a gas tight furnace enclosure, radiant superheater partial division walls, and convective finishing superheater and reheater tube banks. After leaving the furnace, flue gas energy is recovered in a parallel-pass convective heat recovery area (HRA) and a regenerative rotary air heater. Final main steam temperature is controlled by spray water attemperation, while reheat steam temperature is controlled by a HRA gas flow proportioning damper, which regulates the flue gas flow to two parallel superheater and reheater passes. Primary air and secondary air are heated in the air heater after which the former is sent to vertical spindle MBF pulverizers to convey the coal and the latter is sent to the furnace windbox. Figure 1 shows a sketch of typical boiler layout.

Figure 1 – Boiler Layout

Key assumptions are listed below:

1. For air-fired operation, both FGD and SCR are operating. NO_x reduction in SCR is 72% and SO_x capture in FGD is 71% for PRB coal and 98% for bituminous coal.
2. For oxy-fired operation, ammonia injection to the SCR unit is shut down; however, any reactions that could be catalyzed by the SCR are assumed to take place such as oxidations of SO₂ to SO₃, CO to CO₂, and Hg(0) to Hg(+ / ++), where the extent of oxidation is determined by the catalyst and conditions. Since no data is available for oxy-firing, the same extent/conversion as for air-firing was assumed for the oxy-firing.
3. For oxy-fired operation, the FGD is used to treat a portion of recirculated flue gas, which is then quenched and used as the primary gas stream and part of the secondary gas stream (bituminous only).
4. Air ingress is 2% (by vol.) of the flue gas exiting the boiler.
5. The purity of oxygen from the ASU is 97% by volume with Ar = 2.9% and N₂ = 0.1%.
6. Boiler outlet oxygen concentration (vol.) is 2.4% for air-firing and O₂-firing of PRB fuel and 3.0% for air-firing and O₂-firing of bituminous fuel
7. Overfire air is 25% for air-firing, 25% for PRB O₂-firing and 0% for bit. O₂-firing
8. A boiler design of 1% margin is applied.
9. Main steam flow was maintained the same for air-firing and O₂-firing.
10. To simplify the analysis, the same air-fired furnace design was used for both PRB and bituminous fuels. Although the slagging potential for the PRB and bituminous coals may affect the base air-fired geometry, it will have negligible influence on the retrofit modifications and cost. Note that all specific combustion characteristics of the two fuels (ultimate analysis, heating value, kinetics, etc.) were included.

Air-Fired

The air-fired power plant model is shown in Figure 2. Heat released from combustion is absorbed in the furnace by waterwalls, division walls, and roof and absorbed by the superheater, reheater, and economizer convective tube bundles in the heat recovery area (HRA). After exiting the economizer, flue gas passes through the selective catalytic reactor (SCR) for NO_x control and then through a regenerative air heater to heat up both primary air (PA) and secondary air (SA). Downstream of the air heater, a baghouse (BHG) removes particulates and a flue gas desulfurizer (FGD) captures SO_x using a sprayed limestone solution/slurry and reduces the gas temperature by sprayed water. An air ingress of 2%v is assumed and lumped into BHG in the model.

Main operational variables are controlled as follows: main steam temperature by two water attenuators, reheat steam temperature by a HRA parallel pass gas damper, pulverizer PA outlet temperature by tempering air, and steam generation by coal firing rate. The same mole fraction of O₂ in the furnace flue gas is maintained for both air-firing and oxy-firing.

Emission predictions in the model include NO_x, SO_x, CO, PM, NH₃, VOC, HCl, and Hg. A wet FGD with limestone feed is applied to remove most of SO_x to form gypsum (CaSO₄*2H₂O). For NO_x control, the power plant is equipped with low NO_x burners, air staging (over-fire air), and SCR. In the model, part of Hg is oxidized and removed by the combination of SCR/BHG/FGD. Fuel-Cl released as HCl in the furnace is effectively removed with SO_x in the FGD to form MgCl₂, which is drained together with gypsum.

Figure 2 – Aspen-Plus Air-Fired Model

O₂-Fired

The O₂-fired power plant model is shown in Figure 3. Modifications to adapt the plant to accommodate oxyfuel firing include the addition of hot/wet and cold/dry flue gas recirculation ducts, oxygen distribution piping, two additional steam extractions for air separation unit (ASU)/CO₂ processing unit (CPU), added flue gas cooler for CPU, increased size of HRA lower economizer, a low pressure economizer downstream of the air-heater, and a quench tower, as well as increased cooling duty and equipment size. In addition for the reduced SO_x bituminous-fired case a primary gas heater is added.

Secondary gas recycle and gas extracted to the CPU are taken before the FGD and primary gas recycle is taken after the FGD and quench tower since the coal mill requires low moisture nearly SO_x-free gas for proper performance and avoidance of acid corrosion. For the reduced SO_x bituminous-fired case, secondary gas recycle is split 64% before the FGD and 36% after the FGD to limit the boiler corrosive SO_x concentration

An additional low pressure economizer is required to cool the gas temperature down in oxy-firing because the air heater functions as a recuperator. The air heater duty is reduced since recycled flue gas flow is less than the air-firing air flow and the secondary gas inlet temperature is significantly higher than the air inlet temperature.

Due to air heater leakage, the O₂ from the ASU is not mixed with the recycled gas before the air heater to avoid O₂ leakage to the flue gas. Oxygen from the ASU is assumed to be supplied at 97%v purity and 120°F. Oxygen is assumed to be mixed with the secondary gas stream just upstream of the windbox. For proper combustion, some portion of the oxygen may be required to be mixed with the primary gas either at the burners or prior to the burners depending on the burner design and piping constraints.

Steam at about 300 psia and 100 psia is extracted and de-superheated by makeup water from steam cycle, and sent to the ASU and CPU for regeneration and other uses. The return condensate from the ASU and CPU is ducted to deaerator or condenser depending upon its pressure. A working fluid circulation loop (hot water at 950 psia and 525°F) is used to preheat the CPU vent gas (external to the FW model) by fluegas downstream of the SCR. The return water is set to 250°F to avoid the acid gas dew point.

Ammonia to the SCR was shut-off, but the SCR catalyst layers remained in place. The FGD is operated at part load to treat only the recycled primary gas stream. Limestone fed to the FGD was reduced to the amount required to remove SO_x from the recycled primary gas stream. For the PRB fuel, SO_x removal in total is 47% (as compared to 71% in the air-fired case). For the bituminous fuel, SO_x removal in total is 43% for the treatment of only the primary gas and 62% for the reduced SO_x case for the treatment of primary gas + 36% secondary gas (as compared to 98% in the air-fired case).

Emissions generated in the boiler were predicted by 3-D CFD simulations: NO_x and CO by FW-FIRE and SO₃ by Fluent. For PRB-firing over-fire gas was set at 25% of the total combustion gas (the same as for air-firing). For bituminous-firing over-fire gas was shut off to increase NO_x so as to shift NO_x control duty to the CPU (as specified by Praxair).

Figure 3 – Aspen-Plus O₂-Fired Model

Results and Discussion

Air-Fired Low Sulfur Fuel

The plant fires 450 klb/hr of 27% moisture PRB coal with 15% excess air to produce 2.99 Mlb/hr 2415 psia 1000°F main steam, 2.63 Mlb/hr 603 psia 1000°F reheat steam at the turbine and discharges to the condenser at 2.75" Hg. Plant net efficiency is 35.8%, net power is 418 MWe, and boiler efficiency is 86.7%. Power plant feed flow rates are presented in Table 5 and steam cycle auxiliary power is presented in Table 6. Flue gas composition and emissions are presented in Table 7.

The Foster Wheeler CFD computer program, FW-FIRE, was used to determine the furnace performance. The FW-FIRE model simulates the furnace, in height from the bottom of the hopper to the roof, in depth from the front wall to the rear wall, and in width from the left side wall to the right side wall. The furnace model is shown in Figure 4. Figure 5 presents a plot of gas temperature in a vertical plane through the second burner column. The heat flux at the furnace water wall is shown in Figure 6. The maximum heat flux is approximately 70,000 Btu/hr-ft² and the total heat absorbed is 1812 Mbtu/hr.

Table 5 – Feed Flow Rates: Low Sulfur Fuel

Flow Rates		Air-fired	O2-fired
Fuel	klb/h	449	452
Air	klb/h	3585	0
Oxygen	klb/h	0	725
Recirculated Flue Gas	klb/h	0	2550
Air Ingress	klb/h	81	61
Air Heater Leakage	klb/h	191	190
NH3	klb/h	0.29	0
FGD Makeup Water	klb/h	186	44
Limestone	klb/h	2.14	1.35
Condenser Cooling Water	k-gpm	147	156
Total Cooling Water Flow	k-gpm	147	232

Table 6 – Steam Cycle Power and Efficiency: Low Sulfur Fuel

		Air-fired	O2-fired
Heat in	Mbtu/h	3980	4012
Boiler efficiency	%	86.7	89.4
Gross Power	MWe	461	467
Aux. Power	MWe	43	38
ASU/CPU Power	MWe		125.9
ASU/CPU Cooling Water Power	MWe		2.5
Net Power	MWe	418	301
Net efficiency	%	35.8	25.6
Auxiliary Power			
Cond. Pump	MWe	0.8	0.8
FW Pump	MWe	12.7	12.7
FGD	MWe	2.9	0.8
PA fan	MWe	2.6	1.9
SA fan	MWe	2.2	1.6
ID fan	MWe	8.3	6.4
Pulverizers	MWe	1.8	1.8
Condenser Cooling	MWe	4.7	5.0
others	MWe	6.9	7.0
total	MWe	42.8	37.9

Table 7 – Flue Gas Composition and Emissions: Air-Fired Low Sulfur Fuel

		Air-fired		O2-fired	
		Furnace	Stack	Furnace	To CPU
Flow	klb/h	3822	4276	3517	1108
FEGT	F	2202	131	2164	325
H2O	%v	12.38	17.65	28.78	27.28
O2	%v	2.37	3.35	2.33	2.75
N2	%v	69.64	65.39	4.08	5.69
Ar	%v	0.83	0.78	2.14	2.14
CO2	%v	14.73	12.80	62.57	62.05

	Air-Fired at Stack			O2-Fired to CPU		
	ppmv	lb/h	lb/Mbtu	ppmv	lb/h	lb/Mbtu
CO	128	536	0.13	280	244	0.06
SO2	52	501	0.13	465	928	0.23
SO3	1.3	15.8	0.0040	5.8	14.4	0.0036
NOx	36	247	0.062	156	223	0.056
NH3	0.7	1.5	0.00037	0	0	0
HCl	0.1	0.8	0.00020	18	20.9	0.0052
PM		49	0.012		49	0.012
VOC	1.8	11.9	0.0030	1.3	1.8	0.0004
	ppbv	lb/h	lb/Tbtu	ppbv	lb/h	lb/Tbtu
Hg	1.7	0.051	12.8	10.1	0.063	15.7

**Figure 4 – Furnace Model
(with right side wall removed)**

Figure 5 – Gas Temperature: Air-Fired Low Sulfur Fuel

Figure 6 – Wall Heat Flux: Air-Fired Low Sulfur Fuel

O₂-Fired Low Sulfur Fuel

Iterations were performed between the Aspen-Plus system model and FW-FIRE CFD furnace model to determine the required amount of flue gas recirculation. For a given flue gas recirculation quantity, the firing rate and attemperation were adjusted to make steam flow and temperature. The minimum quantity of flue gas recirculation was selected to match the air-fired heat transfer profile through the boiler. The flue gas recirculation flow to the boiler is 68% of the flue gas (44% as secondary gas at 325°F after ID fan and 24% as primary gas at 102°F after wet FGD and quenching tower) and is recuperated to 670°F. Required oxidant flow is 725 klb/hr and the average O₂ in feed gas is 27.4%v. The fluegas exit O₂ concentration is the same as the air-fired case at 2.35%v. Furnace exit gas temperature (FEGT) is 2164°F, which is about 38°F less than for the air-fired case.

The plant fires 452 klb/hr of PRB coal to produce 2.99 Mlb/hr 2415 psia 1000°F main steam, 2.63 Mlb/hr 603 psia 1000°F reheat steam at the turbine and discharges to the condenser at 2.75" Hg. Boiler island gross power is 467 MWe, auxiliary power for the boiler is 38 MWe, and boiler efficiency is 89.4%. Heat extracted for the ASU and CPU is 47 Mbtu/hr. The boiler island net efficiency is 36.5% (air-fired boiler net efficiency = 35.8%) without counting the power consumption of the ASU and CPU, but including heat extraction to the ASU and CPU (without heat extraction to the ASU and CPU net power would increase by 4.4 MWe). The lower flue gas flow rate of O₂-firing than air-firing increases boiler efficiency due to less heat discharge and reduces auxiliary power due to less fan power. Net efficiency including ASU/CPU duty (125.9 MWe from Praxair) and quench tower and ASU/CPU cooling water duty (2.5 MWe) is 25.6%. CO₂ capture efficiency of the CPU is 99%. Power plant feed flow rates are presented in Table 5 and steam cycle auxiliary power is presented in Table 6. Flue gas composition and emissions are presented in Table 7. The CO₂ purity of the gas to the CPU is 62.0%v (85.3%dv).

The lower economizer duty is increased by 28% compared to air-firing by increasing surface area by 45%. All other HRA banks are the same size and have nearly the same performance as air-firing. Flue gas exiting the HRA is cooled by 10°F to heat up the vent gas from the CPU through a circulated working fluid (970 psia hot water). A low pressure economizer (LPECO) is added downstream of air heater to cool the fluegas to 310°F (136 Mbtu/hr) by extracted condensate at 223°F. The LPECO fully replaces one feedwater heater just before the deaerator. The addition of the LPECO and added economizer surface is required due the reduced duty of the air-air heat exchanger in the O₂-fired operation.

The water attemperation flow for control of the main steam temperature is 155 klb/hr (5.2%), which is nearly the same as for the air-fired case (6.3%). Due to increased end steam flow of 2097 klb/hr (Air=2011 klb/hr), the condenser cooling duty in the steam cycle rises to 2009 Mbtu/hr (Air=1885 Mbtu/hr) resulting in a cooling water flow increase of 9 k-gpm (4,700 klb/hr) compared to the air-fired case. Additional cooling water is required by the quench tower (6,300 klb/hr) and the ASU/CPU (31,769 klb/hr from Praxair) resulting in additional cooling power load of 2.5 MWe.

Figure 7 – Gas Temperature: O₂-Fired Low Sulfur Fuel

Figure 8 – Wall Heat Flux: O₂-Fired Low Sulfur Fuel

FW-FIRE was used to determine the furnace performance. Figure 7 presents a plot of gas temperature in a vertical plane through the second burner column. The heat flux at the furnace water wall is shown in Figure 8. The maximum heat flux is approximately 71,500 Btu/hr-ft² and the total heat absorbed is 1809 Mbtu/hr. The flame shapes and heat flux patterns are nearly identical to those under air-firing indicating the capability of the furnace to operate effectively under O₂-firing. Table 8 presents the maximum tube wall temperatures and shows that the maximum increase in wall temperature is less than 10°F for oxy-fuel firing.

Table 8 – Maximum Tube Wall Temperatures: O₂-Fired Low Sulfur Fuel

		Air-Fired	O2-Fired
Waterwall	F	791	793
Div Wall	F	861	860
Final SH	F	1045	1049
Final RH	F	1067	1075
Primary SH	F	787	784
Upper Econ	F	580	585
Primary RH	F	870	880
Lower Econ	F	543	552

Since the wet FGD is used to treat only the recycled primary gas (35% of the recycled gas), the SO₂ at furnace exit is about 500 ppmv (assuming 98% FGD efficiency) compared to air-firing at 210 ppmv. The SO₂ leaving the model is 465 ppmv (0.23 lb/Mbtu) to the CPU compared to air-firing at 52 ppmv (0.13 lb/Mbtu) to the stack. For the recycled primary flue gas 98% SO₂ is removed by the wet-FGD and water quenching producing a SO₂ concentration of 12 ppmv.

Fluent was used to determine the SO₃/SO₂ conversion. Small amounts of SO₃ are formed in the boiler by oxidation of SO₂. For air-firing, the ratio of SO₃/SO₂ at the furnace outlet was 0.57%. For oxy-firing, the ratio of SO₃/SO₂ at the furnace outlet was 0.61%. The outlet SO₃ was found to be independent of the amount of SO₃ in the recycle flue gas since SO₃ is destroyed in the high temperature regions of the furnace and only formed in the upper sections of the furnace as temperature reduces. Furthermore, the ratio of SO₃/SO₂ was determined to be independent of the amount of SO₂ in the recycle flue gas. Based on these Fluent simulations, it is concluded that the ratio of SO₃/SO₂ in the flue gas will be only about 7% higher for oxy-firing as air-firing. This can be explained by the fact that the formation of SO₃ occurs in the upper section of the furnace where the oxy-fired furnace O₂ concentration and temperature are very similar to that of the air-fired furnace. The 7% increase for oxy-firing is presumably due to longer residence time of oxy-fired flue gas in the furnace (volumetric flow of oxy-fired flue gas is 25% less than air-fired flue gas). Based on published test results, the SO₃/SO₂ generally peaks at about 1% near the furnace outlet and then decreases to as low as 0.2% due to adsorption and reaction (i.e. to form H₂SO₄) with fly ash. For conservatism, the SO₃ predicted by Fluent at the furnace outlet is applied. Additional SO₃ is created by the SCR catalyst where approximately 0.5% of SO₂ is oxidized to SO₃. Thus, the boiler outlet SO₃ concentration is 6.2 ppm for oxy-firing versus 2.5 ppm for air-firing. Model outlet SO₃ concentration is 5.8 ppm (3.6 lb/Bbtu) for oxy-firing versus 1.3 ppm (4.0 lb/Bbtu) for air-firing. For the recycled primary flue gas 90% SO₃ is removed by the wet-FGD and water quenching producing a SO₃ concentration of 0.7 ppmv.

FW-FIRE was used to determine boiler outlet NO_x. For oxy-firing with no ammonia injection into the SCR (full amount of NO_x was recycled in the flue gas) the NO_x was predicted to be 156 ppm at the boiler outlet (0.056 lb/Mbtu to CPU) compared to the air-fired boiler outlet NO_x of 147 ppm (0.25 lb/Mbtu) and SCR outlet NO_x of 36 ppm (0.062 lb/Mbtu to stack).

FW-FIRE was used to determine boiler outlet CO. For oxy-firing the outlet CO mole fraction was predicted to be about 2.2 times that of air-firing. The outlet CO was determined to be independent of the amount of CO in the recycle flue gas since CO (and likewise VOC) is destroyed in the high temperature high oxygen regions of the near burner region. Thus, the increase in CO can be explained by the greater equilibrium CO concentration from oxy-firing which at the equilibrium temperature is approximately 2.3 times that of air-firing. The CO at the model outlet is predicted to be 280 ppm (0.061 lb/Mbtu to the CPU) compared to the air-fired emission of 128 ppm (0.13 lb/Mbtu to the stack).

Mercury exits the furnace at 18 ppbv and is reduced to 10 ppbv (15.7 lb/Tbtu) by the SCR catalyst (removed mercury remains in the catalyst and ash) before exiting to the CPU compared to the air-fired boiler outlet Hg of 1.4 ppbv (10.5 lb/Tbtu) to the stack. For the recycled primary flue gas all Hg is essentially removed by the wet-FGD due to reduced loading and by water quenching by cooling the fluegas to a low temperature.

HCl exits the furnace and the model (to the CPU) at 20 ppmv (5.2 lb/Bbtu) compared to the air-fired boiler outlet HCl of 0.14 ppmv (0.2 lb/Bbtu). For the recycled primary flue gas nearly all HCl is removed by the wet-FGD and water quenching.

Air-Fired High Sulfur Fuel

The plant fires 331 klb/hr of 11% moisture bituminous coal with 18% excess air to produce 2.99 Mlb/hr 2415 psia 1000°F main steam, 2.63 Mlb/hr 603 psia 1000°F reheat steam at the turbine and discharges to the condenser at 2.75" Hg. Plant net efficiency is 36.7%, net power is 416 MWe, and boiler efficiency is 89.3%. Power plant feed flow rates are presented in Table 9 and steam cycle auxiliary power is presented in Table 10. Flue gas composition and emissions are presented in Table 11.

The Foster Wheeler CFD computer program, FW-FIRE, was used to determine the furnace performance. The furnace model is shown in Figure 4 (same geometry as for PRB-fired). Figure 9 presents a plot of gas temperature in a vertical plane through the second burner column. The heat flux at the furnace water wall is shown in Figure 10. The maximum heat flux is approximately 76,000 Btu/hr-ft² and the total heat absorbed is 1887 Mbtu/hr.

Table 9 – Feed Flow Rates: High Sulfur Fuel

Flow Rates		Air-fired	O2-fired	
				Low SOx
Fuel	klb/h	331	334	334
Air	klb/h	3567	0	0
Oxygen	klb/h	0	712	712
Recirculated Flue Gas	klb/h	0	2754	2829
Air Ingress	klb/h	77	60	60
Air Heater Leakage	klb/h	150	149	149
NH3	klb/h	0.32	0	0
FGD Water Makeup	klb/h	235	61	101
Limestone	klb/h	27.2	12.0	17.1
Condenser Cooling Water	k-gpm	147	152	154
Total Cooling Water Flow	k-gpm	147	226	228

Table 10 – Steam Cycle Power and Efficiency: High Sulfur Fuel

		Air-fired	O2-fired	
				Low SOx
Heat in	Mbtu/h	3861	3893	3894
Boiler efficiency	%	89.3	92.1	92.4
Gross Power	MWe	460	467	465
Aux. Power	MWe	45	38	40
ASU/CPU Power	MWe		119.7	119.7
ASU/CPU Cooling Water Power	MWe		2.4	2.4
Net Power	MWe	416	306	304
Net efficiency	%	36.7	26.9	26.6
Auxiliary Power				
Cond. Pump	MWe	0.8	0.8	0.8
FW Pump	MWe	12.7	12.7	12.7
FGD	MWe	6.1	2.1	3.3
PA fan	MWe	2.0	1.5	1.5
SA fan	MWe	2.4	1.9	1.8
ID fan	MWe	7.8	6.3	6.3
Pulverizers	MWe	1.3	1.3	1.3
Condenser Cooling	MWe	4.7	4.9	5.0
others	MWe	6.9	7.0	7.0
total	MWe	44.6	38.4	39.5

Table 11 – Flue Gas Composition and Emissions: Air-Fired High Sulfur Fuel

		Air-fired		O2-fired			
		Furnace	Stack	Furnace	To CPU	Furnace	To CPU
						Reduced SOx	
Flow	klb/h	3715	4129	3618	1002	3693	971
FEGT	F	2199	127	2112	325	2119	310
H2O	%v	8.93	15.85	23.46	22.45	19.01	18.25
O2	%v	2.99	3.63	2.94	3.32	2.93	3.31
N2	%v	72.97	67.55	4.99	6.54	5.36	6.88
Ar	%v	0.87	0.81	2.41	2.40	2.56	2.54
CO2	%v	13.99	12.14	65.50	64.63	69.62	68.53

Air-Fired at Stack				O2-Fired to CPU					
							Reduced SOx		
	ppmv	lb/h	lb/Mbtu	ppmv	lb/h	lb/Mbtu	ppmv	lb/h	lb/Mbtu
CO	133	538	0.14	282	215	0.06	284	205	0.05
SO2	35	328	0.08	5349	9349	2.40	3837	6326	1.62
SO3	2.4	28.0	0.0073	66.7	145.7	0.0374	47.4	97.7	0.0251
NOx	40	267	0.069	388	522	0.134	391	496	0.127
NH3	0.6	1.6	0.00041	0	0	0	0	0	0
HCl	3.7	19.4	0.00503	559.7	556.7	0.143	402.1	377.3	0.097
PM		79	0.020		70	0.018		66	0.017
VOC	1.4	9.1	0.0024	1.1	1.3	0.0003	1.2	1.3	0.0003
	ppbv	lb/h	lb/Tbtu	ppbv	lb/h	lb/Tbtu	ppbv	lb/h	lb/Tbtu
Hg	0.2	0.004	1.1	1.1	0.006	1.6	1.0	0.005	1.3

Figure 9 – Gas Temperature: Air-Fired High Sulfur Fuel

Figure 10 – Wall Heat Flux: Air-Fired High Sulfur Fuel

O₂-Fired High Sulfur Fuel: High SO_x

Iterations were performed between the Aspen-Plus system model and FW-FIRE CFD furnace model to determine the required amount of flue gas recirculation. The quantity of flue gas recirculation was selected to minimize changes to the boiler by approximately matching heat transfer and maximum metal temperatures. The flue gas recirculation flow to the boiler is 72% of the flue gas (54% as secondary gas at 310°F after ID fan and 18% as primary gas at 102°F after wet FGD and quenching tower) and is recuperated to 675°F. Required oxidant flow is 712 klb/hr and the average O₂ in feed gas is 27.2%v. The fluegas exit O₂ concentration is the same as the air-fired case at 3.0%v. Furnace exit gas temperature (FEGT) is 2112°F, which is about 87°F less than for the air-fired case.

The plant fires 334 klb/hr of bituminous coal to produce 2.99 Mlb/hr 2415 psia 1000°F main steam, 2.63 Mlb/hr 603 psia 1000°F reheat steam at the turbine and discharges to the condenser at 2.75" Hg. Boiler island gross power is 467 MWe, auxiliary power for the boiler is 38 MWe, and boiler efficiency is 92.1%. Heat extracted for the ASU and CPU is 47 Mbtu/hr. The boiler island net efficiency is 37.4% (air-fired boiler net efficiency = 36.7%) without counting the power consumption of the ASU and CPU, but including heat extraction to the ASU and CPU (without heat extraction to the ASU and CPU net power would increase by 4.4 MWe). Net efficiency including ASU/CPU duty (119.7 MWe from Praxair) and quench tower and CPU cooling water duty (2.4 MWe) is 26.9%. CO₂ capture efficiency of the CPU is 99%.

Power plant feed flow rates are presented in Table 9 and steam cycle auxiliary power is presented in Table 10. Flue gas composition and emissions are presented in Table 11. The CO₂ purity of the gas to the CPU is 64.6%v (83.3%dv).

The lower economizer duty is increased by 35% compared to air-firing by increasing surface area by 45%. All other HRA banks are the same size and have nearly the same performance as air-firing. Flue gas exiting the HRA is cooled by 10°F to heat up the vent gas from the CPU through a circulated working fluid (970 psia hot water). A low pressure economizer (LPECO) is added downstream of air heater to cool the fluegas to 310°F (135 Mbtu/hr) by extracted condensate at 223°F. The LPECO fully replaces one feedwater heater just before the deaerator.

The water attemperation flow for control of the main steam temperature is 72 klb/hr (2.6%), which is less than for the air-fired case (3.9%). Due to increased end steam flow of 2096 klb/hr (Air=2010 klb/hr), the condenser cooling duty in the steam cycle rises to 1965 Mbtu/hr (Air=1884 Mbtu/hr) resulting in a cooling water flow increase of 6 k-gpm compared to the air-fired case. Additional cooling water is required by the quench tower (7,200 klb/hr) and the ASU/CPU (29,671 klb/hr from Praxair) resulting in additional cooling power load of 2.4 MWe.

The maximum heat flux is approximately 82,500 Btu/hr-ft² and the total heat absorbed is 1872 Mbtu/hr.

Since the wet FGD is used to treat only the recycled primary gas (23% of the recycled gas), the SO₂ at furnace exit is about 5690 ppmv (assuming 99% FGD efficiency) compared to air-firing at 205 ppmv. The SO₂ leaving the model is 5350 ppmv (2.4 lb/Mbtu) to the CPU compared to air-firing at 35 ppmv (0.08 lb/Mbtu) to the stack. For the recycled primary flue gas 99% SO₂ is removed by the wet-FGD and water quenching producing a SO₂ concentration of 63 ppmv. This SO_x level of 5690 ppm is comparable to an air-fired unit burning 6.9% (7.7% dry) sulfur coal.

FW-FIRE was used to determine boiler outlet NO_x. For oxy-firing with no ammonia injection into the SCR (full amount of NO_x was recycled in the flue gas) the NO_x was predicted to be 388 ppm at the boiler outlet (0.134 lb/Mbtu to CPU) compared to the air-fired boiler outlet NO_x of 166 ppm (0.25 lb/Mbtu) and SCR outlet NO_x of 40 ppm (0.069 lb/Mbtu to stack).

FW-FIRE was used to determine boiler outlet CO. For oxy-firing the outlet CO mole fraction was predicted to be about 2.1 times that of air-firing. The outlet CO was determined to be independent of the amount of CO in the recycle flue gas since CO (and likewise VOC) is destroyed in the high temperature high oxygen regions of the near burner region. Thus, the increase in CO can be explained by the greater equilibrium CO concentration for oxy-firing which at the equilibrium temperature is approximately 2.1 times that of air-firing. The CO at the model outlet is predicted to be 282 ppm (0.055 lb/Mbtu to the CPU) compared to the air-fired emission of 133 ppm (0.14 lb/Mbtu to the stack).

Mercury exits the furnace at 2.0 ppbv and is reduced to 1.1 ppbv (1.6 lb/Tbtu) by the SCR catalyst before exiting to the CPU compared to the air-fired boiler outlet Hg of 0.2 ppbv (1.1 lb/Tbtu). For the recycled primary flue gas all Hg is essentially removed by the wet-FGD and water quenching.

HCl exits the furnace and the model (to the CPU) at 560 ppmv (143 lb/Bbtu) compared to the air-fired boiler outlet HCl of 3.7 ppmv (5.0 lb/Bbtu). For the recycled primary flue gas nearly all HCl is removed by the wet-FGD and water quenching.

O₂-Fired High Sulfur Fuel: Reduced SO_x

To reduce the potential for boiler corrosion, a case was run with reduced SO_x in the boiler. Waterwall (lower furnace) corrosion is function of waterwall temperature and the deposit and gas composition. Since in O₂-firing compared to air-firing, the waterwall temperature is nearly the same and the composition of the ash deposit is not expected to change significantly, the predominant effect on waterwall corrosion is from the gas composition. The most important corrosive gas species is H₂S, which occurs in reducing conditions, especially due to combustion staging. In general, switching from air-firing to O₂-firing increases the level of H₂S by 50-100%. However, because in the current design, the OFA is reduced from 25% to 0% when switching from air to O₂, the maximum H₂S actually decreases by nearly 50%. Consequently the corrosion of the waterwalls is not expected to be an issue even when only the primary gas is treated by the FGD.

However, superheater/reheater corrosion is caused by the presence of liquid alkali iron trisulfates that form on tube surfaces beneath the ash deposit and is a function of the alkali content of the deposited ash and the SO_x concentration in the upper furnace (oxidizing conditions). As shown

in the previous section, without any FGD treatment of the secondary gas, the SO_x level is 5690 ppm which corresponds to the air-firing of a 7.7% S (dry) coal (compared to the air-fired SO_x of 2100 ppm). Consequently, it may be necessary to treat a portion of the secondary gas by the FGD depending on the maximum SO_x level for which the superheater/reheater materials were designed. After a survey of operating FW units, a reasonable upper limit for SO_x was judged to be about 4000 ppm (corresponding to air-firing of a 5.5% S (dry) coal). This was applied as the rough minimum FGD treatment objective. However, a more rigorous material corrosion investigation is required to determine what level of FGD treatment is actually required for specific boiler materials, fuels, and boiler conditions. Such an investigation is currently being performed by FW under DOE/NETL contract, DE- NT0005262, the results of which could be used, when available, to establish a more accurate minimum FGD treatment.

Iterations were performed between the Aspen-Plus system model and FW-FIRE CFD furnace model to determine the required amount of flue gas recirculation. The quantity of flue gas recirculation was selected to minimize changes to the boiler by approximately matching heat transfer and maximum metal temperatures. The flue gas recirculation flow to the boiler is 72% of the flue gas (34% as secondary gas at 310°F after ID fan and 38% as primary + secondary gas at 102°F after wet FGD and quenching tower) and is recuperated to 675°F. Required oxidant flow is 712 klb/hr and the average O₂ in feed gas is 26.4%v. The fluegas exit O₂ concentration is the same as the air-fired case at 3.0%v. Furnace exit gas temperature (FEGT) is 2119°F, which is about 80°F less than for the air-fired case.

The plant fires 334 klb/hr of bituminous coal to produce 2.99 Mlb/hr 2415 psia 1000°F main steam, 2.63 Mlb/hr 603 psia 1000°F reheat steam at the turbine and discharges to the condenser at 2.75" Hg. Boiler island gross power is 465 MWe, auxiliary power for the boiler is 40 MWe, and boiler efficiency is 92.4%. Heat extracted for the ASU and CPU is 47 Mbtu/hr. The boiler island net efficiency is 37.3% (air-fired boiler net efficiency = 36.7%) without counting the power consumption of the ASU and CPU, but including heat extraction to the ASU and CPU (without heat extraction to the ASU and CPU net power would increase by 4.4 MWe). Net efficiency including ASU/CPU duty (119.7 MWe from Praxair) and quench tower and ASU/CPU cooling water duty (2.4 MWe) is 26.6%. CO₂ capture efficiency of the CPU is over 99%.

Power plant feed flow rates are presented in Table 9 and steam cycle auxiliary power is presented in Table 10. Flue gas composition and emissions are presented in Table 11. The CO₂ purity of the gas to the CPU is 68.5%v (83.8%dv).

The lower economizer duty is increased by 35% compared to air-firing by increasing surface area by 45%. All other HRA banks are the same size and have nearly the same performance as air-firing. Flue gas exiting the HRA is cooled by 10°F to heat up the vent gas from the CPU through a circulated working fluid (970 psia hot water). A low pressure economizer (LPECO) is added downstream of air heater to cool the fluegas to 322°F (107 Mbtu/hr) by extracted condensate at 223°F. The LPECO fully replaces one feedwater heater just before the deaerator. A tubular gas-to-gas heater is added downstream of the ID fan to heat the secondary gas, which passed through the FGD, from 102°F to 253°F by cooling the flue gas to 310°F (26 Mbtu/hr). This heating is required to avoid acid gas dew point corrosion after mixing with the high SO_x secondary gas (without FGD treatment).

The water attenuation flow for control of the main steam temperature is 78 klb/hr (2.8%), which is slightly less than for the air-fired case (3.9%). Due to increased end steam flow of 2074 klb/hr (Air=2010 klb/hr), the condenser cooling duty in the steam cycle rises to 1981 Mbtu/hr (Air=1884 Mbtu/hr) resulting in a cooling water flow increase of 7 k-gpm (3,700 klb/hr) compared to the air-fired case. Additional cooling water is required by the quench tower (7,200 klb/hr) and the ASU/CPU (29,671 k/hr from Praxair) resulting in additional cooling power load of 2.4 MWe.

FW-FIRE was used to determine the furnace performance. Figure 11 presents a plot of gas temperature in a vertical plane through the second burner column. The heat flux at the furnace water wall is shown in Figure 12. The maximum heat flux is approximately 82,300 Btu/hr-ft² and the total heat absorbed is 1863 Mbtu/hr. The flame shapes and heat flux patterns are nearly identical to those under air-firing indicating the capability of the furnace to operate effectively under O₂-firing. Table 12 presents the maximum tube wall temperatures and shows that the maximum increase in wall temperature is less than 10°F for oxy-fuel firing.

Figure 11 – Gas Temperature: O₂-Fired High Sulfur Fuel, Reduced Sox

Figure 12 – Wall Heat Flux: O₂-Fired High Sulfur Fuel, Reduced SO_x

Table 12 – Maximum Tube Wall Temperatures: O₂-Fired High Sulfur Fuel

		Air-Fired	O ₂ -Fired	O ₂ -Fired
<u>Wall Temp.</u>				Reduced SO _x
Waterwall	F	802	811	810
Div Wall	F	858	861	862
Final SH	F	1042	1045	1044
Final RH	F	1065	1068	1069
Primary SH	F	777	771	772
Upper Econ	F	573	578	581
Primary RH	F	879	886	886
Lower Econ	F	543	551	552

Since the wet FGD is used to treat the recycled primary gas + 36% of the secondary gas (55% of the recycled gas is treated by FGD), the SO₂ at furnace exit is about 4080 ppmv (assuming 99% FGD efficiency) compared to air-firing at 205 ppmv. The SO₂ leaving the model is 3840 ppmv (1.6 lb/Mbtu) to the CPU compared to air-firing at 35 ppmv (0.08 lb/Mbtu) to the stack. For the FGD treatment of the recycled primary flue gas and 36% of the secondary flue gas, 99% SO₂ is removed by the wet-FGD and water quenching producing a SO₂ concentration of 43 ppmv in the recycled flue gas.

Fluent was used to determine the SO₃/SO₂ conversion. Small amounts of SO₃ are formed in the boiler by oxidation of SO₂. For air-firing, the ratio of SO₃/SO₂ at the furnace outlet was 0.65%. For oxy-firing with no OFA, the ratio of SO₃/SO₂ at the furnace outlet was 0.73%. The outlet SO₃ was found to be independent of the amount of SO₃ in the recycle flue gas since SO₃ is destroyed in the high temperature regions of the furnace and only formed in the upper sections of the furnace as temperature reduces. Furthermore, the ratio of SO₃/SO₂ was determined to be independent of the amount of SO₂ in the recycle flue gas. Based on these Fluent simulations, it is concluded that the ratio of SO₃/SO₂ in the flue gas will be only about 12% higher for oxy-firing as air-firing. This can be explained by the fact that the formation of SO₃ occurs in the upper section of the furnace where the oxy-fired furnace O₂ concentration and temperature are very similar to that of the air-fired furnace. The 12% increase for oxy-firing is presumably due to longer residence time of oxy-fired flue gas in the furnace (volumetric flow of oxy-fired flue gas is 22% less than air-fired flue gas). Based on published test results, the SO₃/SO₂ generally peaks at about 1% near the furnace outlet and then decreases to as low as 0.2% due to adsorption and reaction (i.e. to form H₂SO₄) with fly ash. For conservatism, the SO₃ predicted by Fluent at the furnace outlet is applied. Additional SO₃ is created by the SCR catalyst where approximately 0.5% of SO₂ is oxidized to SO₃. Thus, the boiler outlet SO₃ concentration is 50.1 ppm for oxy-firing versus 13.4 ppm for air-firing. Model outlet SO₃ concentration is 47.4 ppm (25.1 lb/Bbtu) for oxy-firing versus 2.4 ppm (7.3 lb/Bbtu) for air-firing. For the recycled primary flue gas 95% SO₃ is removed by the wet-FGD and water quenching producing a SO₃ concentration of 2.7 ppmv. This SO₃ concentration can be reduced to 1.6 ppmv by removing the SCR catalyst, which may be required to protect the mill from acid corrosion. Normally H₂SO₄ levels and not SO₃

levels are measured at the stack due to the operating temperatures at the test locations. The SO₃ levels noted here are the precursors to the H₂SO₄ levels tested at the stack in air-firing.

FW-FIRE was used to determine boiler outlet NO_x. For oxy-firing with no ammonia injection into the SCR (full amount of NO_x was recycled in the flue gas) the NO_x was predicted to be 397 ppm at the boiler outlet (0.127 lb/Mbtu to CPU) compared to the air-fired boiler outlet NO_x of 166 ppm (0.25 lb/Mbtu) and SCR outlet NO_x of 40 ppm (0.069 lb/Mbtu to stack).

FW-FIRE was used to determine boiler outlet CO. For oxy-firing the outlet CO mole fraction was predicted to be about 2.1 times that of air-firing. The outlet CO was determined to be independent of the amount of CO in the recycle flue gas since CO (and likewise VOC) is destroyed in the high temperature high oxygen regions near the burners. Thus, the increase in CO can be explained by the greater equilibrium CO concentration for oxy-firing, which at the equilibrium temperature is approximately 2.1 times that of air-firing. The CO at the model outlet is predicted to be 284 ppm (0.053 lb/Mbtu to the CPU) compared to the air-fired emission of 133 ppm (0.14 lb/Mbtu to the stack).

Mercury exits the furnace at 1.7 ppbv and is reduced to 1.0 ppbv (1.3 lb/Tbtu) by the SCR catalyst before exiting to the CPU compared to the air-fired boiler outlet Hg of 0.2 ppbv (1.1 lb/Tbtu). For the recycled primary flue gas all Hg is essentially removed by the wet-FGD and water quenching.

HCl exits the furnace and the model (to the CPU) at 402 ppmv (97 lb/Bbtu) compared to the air-fired boiler outlet HCl of 3.7 ppmv (5.0 lb/Bbtu). For the recycled primary flue gas nearly all HCl is removed by the wet-FGD and water quenching.

Summary of Power Plant Retrofit Modifications

O₂-Fired Low Sulfur Fuel

1. HRA lower economizer enlargement
Additional surface area = 22,250 ft² (total surface area = 72,350 ft²).
Total heat transfer = 170 Mbtu/hr
Inlet/Outlet Water temperature = 495°F /546°F (total lower economizer)
Water flow rate = 2845 klb/hr
LMTD = 285°F (total lower economizer)
2. Low pressure economizer
Surface area = 144,200 ft² (Some tubes may require special coating to avoid flue gas acid corrosion)
Heat transfer = 135 Mbtu/hr
Inlet/outlet water temperature = 223°F /280°F
Water flow rate = 2376 klb/hr
LMTD = 125°F
3. O₂ distribution (725 klb/hr at 120°F)

- a. Piping (Cross Sectional Area = 60 ft²)
 - b. Mixing device
 - c. Flow controller
4. Hot gas recirculation (1650 klb/hr at 325°F)
 - a. Extraction port
 - b. Ducting to SA Fan (Cross Sectional Area = 185 ft²)
 - c. Flow controller
 - d. Air/gas switch junction
5. Cold gas recirculation (900 klb/hr at 102°F)
 - a. Extraction port
 - b. Ducting to PA Fan (Cross Sectional Area = 65 ft²)
 - c. Flow controller
 - d. Air/gas switch junction
 - e. Small reheat (i.e. 10°F) by fluegas recuperator (using low-pressure water circulation) to avoid acid dew point
6. High pressure steam extraction (18 klb/hr at 317 psia, 816°F) for CPU/ASU]
 - a. Steam extraction port
 - b. Piping
 - c. Flow control
 - d. De-superheater (including water spray and pump) to 425°F at 22 klb/hr
 - e. Condensate return piping
7. Low pressure steam extraction (13 klb/hr at 107 psia, 552°F) for CPU/ASU
 - a. Steam extraction port
 - b. Piping
 - c. Flow control
 - d. De-superheater (including water spray and pump) to 342°F at 14 klb/hr
 - e. Condensate return piping
8. CPU vent gas heater

Surface area = 3,560 ft²
Heat transfer = 10 Mbtu/hr
Inlet/outlet water temperature = 250°F /525°F
Water flow rate = 34 klb/hr
LMTD = 290°F
9. CPU vent gas heater circulation pump
 - a. Water flow rate = 34 klb/hr (70 gpm)
 - b. Pressure increase = 100 psi
 - c. Power = 6 hp
10. Quenching tower
 - a. Booster fan

- b. Ducting
 - c. Water circulation pump
 - d. Acid water treatment
11. Flue gas duct to CPU (1108 klb/hr at 325°F)
- a. Extraction port
 - b. Ducting (Cross Sectional Area = 125 ft²)
 - c. Flow controller
12. FGD modification
- a. Use oxygen instead of air for bubbling
 - b. O₂ flow = 0.7 klb/hr
13. Auxiliary air replacement by auxiliary CO₂
- a. Replace auxiliary air by CO₂ for air purge systems and solid transport systems
 - b. CO₂ flow = 1500 scfm (11 klb/hr) at 80 psig
14. Improved Boiler Sealing
- a. Reduce air in-leakage by mechanical sealing (as is practical)
15. Enlarge cooling tower
- a. Without ASU/CPU requirements, cooling tower increases by 20-25%
16. DCS controller and instrumentation for oxyfuel operation, switch over from air, etc.
17. Flue gas cooler

O₂-Fired High Sulfur Fuel

1. HRA lower economizer enlargement
Additional surface area = 22,250 ft² (total surface area = 72,350 ft²).
Total heat transfer = 171 Mbtu/hr
Inlet/Outlet Water temperature = 495°F /545°F (total lower economizer)
Water flow rate = 2922 klb/hr
LMTD = 285°F (total lower economizer)
2. Low pressure economizer
Surface area = 108,150 ft² (Some tubes may require special coating to avoid flue gas acid corrosion)
Heat transfer = 107 Mbtu/hr
Inlet/outlet water temperature = 223°F /268°F
Water flow rate = 2349 klb/hr
LMTD = 130°F
3. O₂ distribution (712 klb/hr at 120°F)

- a. Piping (Cross Sectional Area = 60 ft²)
 - b. Mixing device
 - c. Flow controller
4. Hot gas recirculation (1350 klb/hr at 310°F)
 - a. Extraction port
 - b. Ducting to SA Fan (Cross Sectional Area = 140 ft²)
 - c. Flow controller
 - d. Air/gas switch junction
5. Cold gas recirculation (1480 klb/hr at 102°F)
 - a. Extraction port
 - b. Ducting to PA Fan (Cross Sectional Area = 110 ft²)
 - c. Flow controller
 - d. Air/gas switch junction
 - e. Small reheat (i.e. 10°F) by fluegas recuperator (using low-pressure water circulation) to avoid acid dew point
6. High pressure steam extraction (18 klb/hr at 317 psia, 816°F) for CPU/ASU
 - a. Steam extraction port
 - b. Piping
 - c. Flow control
 - d. De-superheater (including water spray and pump) to 425°F at 22 klb/hr
 - e. Condensate return piping
7. Low pressure steam extraction (13 klb/hr at 107 psia, 552°F) for CPU/ASU
 - a. Steam extraction port
 - b. Piping
 - c. Flow control
 - d. De-superheater (including water spray and pump) to 342°F at 14 klb/hr
 - e. Condensate return piping
8. CPU vent gas heater

Surface area = 3,560 ft²
Heat transfer = 10 Mbtu/hr
Inlet/outlet water temperature = 250°F /525°F
Water flow rate = 34 klb/hr
LMTD = 290°F
9. CPU vent gas heater circulation pump
 - a. Water flow rate = 34 klb/hr (70 gpm)
 - b. Pressure increase = 100 psi
 - c. Power = 6 hp
10. Quenching tower
 - a. Booster fan
 - b. Ducting

- c. Water circulation pump
- d. Acid water treatment
- 11. Flue gas duct to CPU (970 klb/hr at 310°F)
 - a. Extraction port
 - b. Ducting (Cross Sectional Area = 100 ft²)
 - c. Flow controller
- 12. FGD modification
 - a. Use oxygen instead of air for bubbling
- 13. Auxiliary air replacement by auxiliary CO₂
 - a. Replace auxiliary air by CO₂ for air purge systems and solid transport systems
 - b. CO₂ flow = 1500 scfm (11 klb/hr) at 80 psig
- 14. Improved Boiler Sealing
 - a. Reduce air in-leakage by mechanical sealing (as is practical)
- 15. Enlarge cooling tower
 - a. Without ASU/CPU requirements, cooling tower increases by 20-25%
- 16. DCS controller and instrumentation for oxyfuel operation, switch over from air, etc.
- 17. Primary gas heater (tubular, gas-to-gas)
 - Surface area = 45,240 ft²
 - Heat transfer = 26 Mbtu/hr
 - Inlet/outlet primary gas temperature = 102°F /253°F
 - Primary air flow rate = 775 klb/hr
 - LMTD = 135°F
- 18. Flue gas cooler

Constraints and Limitations

Note that for the design proposed herein, all limitations have been checked and are acceptable at full load conditions.

1. ID fan surge limit

The maximum quantity of flue gas recirculation is limited by the existing ID fan capacity (i.e. surge limit). Above this limit the furnace pressure is not controllable. For example, this limit may be reached when flue gas recycle is above 75-80%.

2. Superheater attemperators

The minimum quantity of flue gas recirculation is limited by the attemperator design. Below this limit the attemperator flow reduces to zero and the superheat temperature is not controllable. This is produced by too much heat absorption and steam generation in the furnace resulting in HRA flue gas energy, which is insufficient to reach the design outlet steam temperature. For example, this limit may be reached when flue gas recycle is below 60-65%.

The maximum quantity of flue gas recirculation is limited by the maximum water spray capacity of the attemperator design. Above this limit the superheat temperature is not controllable. For example, this limit may be reached when flue gas recycle is above 75-80%.

3. Tube metal temperatures

The quantity of flue gas recirculation is limited by the design tube metal temperature margins. As the recycled flue gas flow rate is reduced, furnace waterwall temperatures increase and HRA tube metal temperatures decrease. As recycled flue gas flow rate is increased, furnace waterwall temperatures decrease and HRA tube metal temperatures increase.

4. Burner design

Burner design operational limits should be checked to determine limitations on flue gas recirculation conditions at full and part loads.

5. Cold gas recirculation SO_x removal

SO_x should be removed as much as possible from cold gas (primary gas) recirculation to avoid acid dew point corrosion in mill and downstream piping.

6. Steam turbine

The shutting off of low pressure feedwater heater(s) will increase flow to the low pressure turbine sections. Turbine design should be checked to determine if this can be accommodated. Note that in the design proposed herein, low pressure steam flow increases by 4%.

Oxyfuel Incremental Equipment Cost Estimate

O₂-Fired Low Sulfur Fuel

A conceptual level cost estimate was performed for the incremental equipment and installation cost of the oxyfuel retrofit in the boiler island and steam system (covering Foster Wheeler's scope of the plant simulation) for the reference 460 MW (gross) plant firing PRB low sulfur fuel. The estimate is summarized in Table 13.

Table 13 – Low Sulfur Fuel O₂-Fired Retrofit Incremental Cost Estimate

	M \$
Heat Exchangers	35.95
Piping and ducting	37.15
Boiler Sealing	7.85
Total	80.95

Details of the estimate assumptions and methodology are presented below.

1. HRA lower economizer enlargement

The heating surface of the lower economizer located below the convection pass Heat Recovery Area (HRA) was increased 22,250 ft², for a total lower economizer surface area of 72,350 ft². The additional heating surface is comprised of 2" OD carbon steel tubes.

It is typical design practice to include space for additional loops of heating surface in new unit designs. This practice allows for the addition of heating surface at a later date to achieve performance guarantees or to provide fuel switching flexibility. The original design of the reference unit for this study included space for additional loops of economizer surface. Therefore the additional surface noted above was added without need for modifications to the economizer inlet/outlet headers, boiler casing or boiler support steel.

2. Low pressure economizer

Low pressure economizer heating surface area of 144,200 ft² was added in the flues downstream of the regenerative air heater.

This additional surface was not considered in the original design of the unit. Therefore, in addition to the 2" OD carbon steel tubes, inlet and outlet headers, transfer piping to / from the new heating surface to the feedwater heater bay, plate casing, sootblowers and support steel were added. Valves for control and isolation of the water flow to the LPEC are included.

Due to the low temperature of the flue gas at this location, this LPEC does not need to be bypassed when not in service. Therefore, no bypass ducts or dampers have been considered.

3. O₂ distribution (725 klb/hr at 120°F)

O₂ supply (with flow control) from the ASU to the boiler area is by others. O₂ mixing with recirculated flue gas and its distribution within the boiler area has been considered by the boiler supplier.

An O₂ injection grid with a mixing device was added upstream of the FD fans. The proposed equipment is similar to the injection grids and mixing devices employed for SCR systems. However, the components and casing in the mixing area are stainless steel material for safety considerations. Costs include additional ducting and support steel.

4. Hot gas recirculation (1650 klb/hr at 325°F)

Ducts of 185 ft² cross sectional area for recirculation of hot flue gas (325°F) from upstream of a wet FGD system to the FD fan inlets have been added.

The ducts are of carbon steel material. Costs include the dampers for air/gas proportioning, turning vanes, expansion joints and support steel.

5. Cold gas recirculation (900 klb/hr at 102°F)

Ducts of 80 ft² cross sectional area for recirculation of cold flue gas (102°F) from downstream of the FGD system to the PA fans inlets have been added.

The ducts are of carbon steel material. Costs include the dampers for air/gas proportioning, turning vanes, expansion joints and support steel. In addition, small amount of reheat (i.e. 10°F) by fluegas recuperator (using low-pressure water circulation) was included to avoid acid dew point concerns in the mill system.

6. High pressure steam extraction (18 klb/hr at 317 psia, 816°F at turbine) for CPU/ASU

High pressure steam supply piping from the steam turbine or power piping to ASU and CPU including flow control valves, desuperheating station, lagging/insulation and pipe supports have been added.

High pressure steam condensate return piping from the ASU and CPU to the condenser hotwell including flow control valves and pipe supports have been added.

Steam supply and return piping inside the ASU and CPU is by others.

7. Low pressure steam extraction (13 klb/hr at 107 psia, 552°F at turbine) for CPU/ASU

Low pressure steam supply piping from the steam turbine or power piping to ASU (located next to the boiler) and CPU (located to the rear of the boiler, by the stack) including flow control valves, desuperheating station and pipe supports have been added.

Low pressure steam condensate return piping from the ASU and CPU to the condenser hotwell including flow control valves, lagging/insulation and pipe supports have been added.

Steam supply and return piping inside the ASU and CPU is by others.

8. CPU vent gas heater

A gas over tube bundle type heat exchanger with 3,560 ft² of heat transfer surface has been added. Casing and steel support is included.

9. CPU vent gas heater circulation pump

- a. Water flow rate = 34 klb/hr (70 gpm)
- b. Pressure increase = 100 psi
- c. Power = 6 hp

BY OTHERS (Not included in estimate).

10. Quenching tower

- a. Booster fan
- b. Ducting
- c. Water circulation pump
- d. Acid water treatment

BY OTHERS (Not included in estimate).

11. Flue gas duct to CPU (1108 klb/hr at 325°F)

Gas ducts of 125 ft² cross sectional area from quenching tower to CPU have been added. These ducts are sized for 1,108,000 lbs/hour of CO₂ gas at 325°F.

The ducts are of carbon steel material. Costs include the dampers for air/gas proportioning, turning vanes, expansion joints and support steel.

12. FGD modification

- a. Use oxygen instead of air for bubbling
- b. O₂ flow = 0.7 klb/hr

BY OTHERS (Not included in estimate).

13. Auxiliary air replacement by auxiliary CO₂

Piping as required to replace auxiliary air with CO₂ for air purge systems and solid transport systems has been added.

14. Improved Boiler Sealing

Improved seals at all locations in the boiler and flues/ducts have been added. Additional sealing of manways and observation ports by use of aspirating air type seals (using CO₂ instead of air) and a penthouse pressurization fan system (again using CO₂ instead of air) have been added. CO₂ piping and ducting have been added as required.

15. Enlarge cooling tower

- a. Without ASU/CPU requirements, cooling tower increases by 20-25%

BY OTHERS (Not included in estimate).

16. DCS controller and instrumentation for oxyfuel operation, switch over from air, etc.

BY OTHERS (Not included in estimate).

17. Flue gas cooler

BY OTHERS (Not included in estimate).

NOTE: A plot plan and sectional side view drawings of the reference unit were used to estimate lengths of runs of pipe, flues and ducts. Tie-in points were estimated based on the process requirements. However, due to the complexity of retrofit work, the details of the tie-ins to the reference unit have not been fully assessed.

O₂-Fired High Sulfur Fuel

A conceptual level cost estimate was performed for the incremental equipment and installation cost of the oxyfuel retrofit in the boiler island and steam system (covering Foster Wheeler's scope of the plant simulation) for the reference 460 MW (gross) plant firing bituminous high sulfur fuel. The estimate is summarized in Table 14.

Table 14 – High Sulfur Fuel O₂-Fired Retrofit Incremental Cost Estimate

	M \$
Heat Exchangers	33.87
Piping and ducting	42.70
Boiler Sealing	7.89
Total	84.46

Details of the estimate assumptions and methodology are presented below.

1. HRA lower economizer enlargement

The heating surface of the lower economizer located below the convection pass Heat Recovery Area (HRA) was increased 22,250 ft², for a total lower economizer surface area of 72,350 ft². The additional heating surface is comprised of 2" OD carbon steel tubes.

It is typical design practice to include space for additional loops of heating surface in new unit designs. This practice allows for the addition of heating surface at a later date to achieve performance guarantees or to provide fuel switching flexibility. The original design of the reference unit for this study included space for additional loops of economizer surface. Therefore the additional surface noted above was added without need for modifications to the economizer inlet/outlet headers, boiler casing or boiler support steel.

2. Low pressure economizer

Low pressure economizer heating surface area of 108,150 ft² was added in the flues downstream of the regenerative air heater.

This additional surface was not considered in the original design of the unit. Therefore, in addition to the 2" OD carbon steel tubes, inlet and outlet headers, transfer piping to / from the new heating surface to the feedwater heater bay, plate casing, sootblowers and support steel were added. Valves for control and isolation of the water flow to the LPEC are included.

Due to the low temperature of the flue gas at this location, this LPEC does not need to be bypassed when not in service. Therefore, no bypass ducts or dampers have been considered.

3. O₂ distribution (712 klb/hr at 120°F)

O₂ supply (with flow control) from the ASU to the boiler area is by others. O₂ mixing with recirculated flue gas and its distribution within the boiler area has been considered by the boiler supplier.

An O₂ injection grid with a mixing device was added upstream of the FD fans. The proposed equipment is similar to the injection grids and mixing devices employed for SCR systems. However, the components and casing in the mixing area are stainless steel material for safety considerations. Costs include additional ducting and support steel.

4. Hot gas recirculation (1350 klb/hr at 325°F)

Ducts of 140 ft² cross sectional area for recirculation of hot flue gas (325°F) from upstream of a wet FGD system to the FD fan inlets have been added.

The ducts are of carbon steel material. Costs include the dampers for air/gas proportioning, turning vanes, expansion joints and support steel.

5. Cold gas recirculation (1480 klb/hr at 102°F)

Ducts of 110 ft² cross sectional area for recirculation of cold flue gas (102°F) from downstream of the FGD system to the PA fans inlets have been added.

The ducts are of carbon steel material. Costs include the dampers for air/gas proportioning, turning vanes, expansion joints and support steel. In addition, small amount of reheat (i.e. 10°F) by fluegas recuperator (using low-pressure water circulation) was included to avoid acid dew point concerns in the mill system.

6. High pressure steam extraction (18 klb/hr at 317 psia, 816°F at turbine) for CPU/ASU

High pressure steam supply piping from the steam turbine to ASU and CPU including flow control valves, desuperheating station, lagging/insulation and pipe supports have been added.

High pressure steam condensate return piping from the ASU and CPU to the condenser hotwell including flow control valves and pipe supports have been added.

Steam supply and return piping inside the ASU and CPU is by others.

7. Low pressure steam extraction (13 klb/hr at 107 psia, 552°F at turbine) for CPU/ASU

Low pressure steam supply piping from the steam turbine to ASU (located next to the boiler) and CPU (located to the rear of the boiler, by the stack) including flow control valves, desuperheating station and pipe supports have been added.

Low pressure steam condensate return piping from the ASU and CPU to the condenser hotwell including flow control valves, lagging/insulation and pipe supports have been added.

Steam supply and return piping inside the ASU and CPU is by others.

8. CPU vent gas heater

A gas over tube bundle type heat exchanger with 3,560 ft² of heat transfer surface has been added. Casing and steel support is included.

9. CPU vent gas heater circulation pump

- a. Water flow rate = 34 klb/hr (70 gpm)
- b. Pressure increase = 100 psi
- c. Power = 6 hp

BY OTHERS (Not included in estimate).

10. Quenching tower

- a. Booster fan
- b. Ducting
- c. Water circulation pump
- d. Acid water treatment

BY OTHERS (Not included in estimate).

11. Flue gas duct to CPU (970 klb/hr at 325°F)

Gas ducts of 100 ft² cross sectional area from quenching tower to CPU have been added. These ducts are sized for 970,000 lbs/hour of CO₂ gas at 325°F.

The ducts are of carbon steel material. Costs include the dampers for air/gas proportioning, turning vanes, expansion joints and support steel.

12. FGD modification

- a. Use oxygen instead of air for bubbling

BY OTHERS (Not included in estimate).

13. Auxiliary air replacement by auxiliary CO₂

Piping as required to replace auxiliary air with CO₂ for air purge systems and solid transport systems has been added.

14. Improved Boiler Sealing

Improved seals at all locations in the boiler and flues/ducts have been added. Additional sealing of manways and observation ports by use of aspirating air type seals (using CO₂ instead of air) and a penthouse pressurization fan system (again using CO₂ instead of air) have been added. CO₂ piping and ducting have been added as required.

15. Enlarge cooling tower

- a. Without ASU/CPU requirements, cooling tower increases by 20-25%

BY OTHERS (Not included in estimate).

16. DCS controller and instrumentation for oxyfuel operation, switch over from air, etc.

BY OTHERS (Not included in estimate).

17. Primary Gas Heater (tubular, gas-to-gas) [PAHA/PAHB]

A tubular gas-to-gas heat exchanger with 45,240 ft² of heat transfer surface has been added. Casing and steel support is included.

18. Flue gas cooler

BY OTHERS (Not included in estimate).

NOTE: A plot plan and sectional side view drawings of the reference unit were used to estimate lengths of runs of pipe, flues and ducts. Tie-in points were estimated based on the process requirements. However, due to the complexity of retrofit work, the details of the tie-ins to the reference unit have not been fully assessed.

Conclusions

A reference air-fired plant burning low sulfur PRB and high sulfur bituminous fuel was simulated. The reference air-fired plant has a boiler efficiency (PRB/Bituminous) of 86.7%/89.3% and a plant net efficiency of 35.8/36.7%.

The plant design was modified to accommodate oxyfuel firing. The flue gas recirculation flow to the boiler is 68% for PRB and 72% for bituminous. Maximum increase in tube wall temperature is less than 10°F. At the furnace outlet, compared to air-firing, NO_x mole fraction is about the same for PRB-firing and 2.5 times for bituminous-firing and CO mole fraction is approximately double.

Retrofitting of existing power plants for oxyfuel combustion is prudent for large (> 300 MW) high efficiency coal plants (subcritical or supercritical). Alternatively, retrofitting is not attractive for old, lower efficiency, smaller units. For these plants, rebuilding or repowering is more suitable.

By proper selection of the quantity of flue gas recycle flow, the boiler (furnace, HRA, air-heater, etc.) can be completely reused in oxy-fuel mode. Boiler retrofit challenges mostly center on the flue gas recycle and the oxygen distribution systems. These challenges include physical routing of ducts and pipes, selection of materials to avoid corrosion, and proper mixing of gases. Another challenge is the development of a control system and control logic for power plant start-up, shutdown, and switchover from air-firing to oxyfuel firing. An additional area for development is improved sealing of the boiler to increase the effectiveness of the CO₂ capture. Overall power plant retrofit constraints include proximity to geologic sequestration, space availability (especially for ASU and CPU), and access to additional water supply for increased cooling load.

For the reference plant presented herein there appears to be no major obstacles to oxyfuel retrofit; however, specific constraints and limitations must be reevaluated by a detailed design.

Areas requiring further study include: 1) Furnace radiation heat transfer coefficients in oxy-firing vs. air-firing, 2) Burner design modifications for O₂-firing and 3) Retrofit costs vs. clean slate costs.

List of Acronyms and Abbreviations

ASU	Air Separation Unit
CFD	Computational fluid dynamics
CPU	CO ₂ Processing Unit
Div	Division
DOE	Department of Energy
Econ	Economizer
FD	Forced draft
FEGT	Furnace exit gas temperature
FGD	Flue Gas Desulfurization
FW	Foster Wheeler, feedwater
FW-FIRE	Fossil fuel, Water-walled Furnace Integrated Reaction and Emission Simulation
HHV	Higher Heating Value
HRA	Heat recovery area
ID	Induced draft
LMTD	Log mean temperature difference
LPECO	Low pressure economizer
NETL	National Energy Technology Laboratory
NO _x	Nitrogen Oxides
OD	Outside diameter
OFA	Over-fired air
PA	Primary Air
PC	Pulverized Coal
PM	Particulate Matter
PRB	Powder River Basin
RH	Reheater
SA	Secondary Air
SCR	Selective Catalytic Reactor
SH	Superheater
SO _x	Sulfur Oxides
VOC	Volatile Organic Compound