Advanced, Low/Zero Emission Boiler Design and Operation

Final Technical Report

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ABSTRACT

In partnership with the U.S. Department of Energy's National Energy Technology Laboratory, B&W and Air Liquide are developing and optimizing the oxy-combustion process for retrofitting existing boilers as well as new plants. The main objectives of the project is to: (1) demonstrate the feasibility of the oxy-combustion technology with flue gas recycle in a 5-million Btu/hr coal-fired pilot boiler, (2) measure its performances in terms of emissions and boiler efficiency while selecting the right oxygen injection and flue gas recycle strategies, and (3) perform technical and economic feasibility studies for application of the technology in demonstration and commercial scale boilers.

This document summarizes the work performed during the period of performance of the project (Oct 2002 to June 2007). Detailed technical results are reported in corresponding topical reports that are attached as an appendix to this report. **Task 1** (Site Preparation) has been completed in 2003. The experimental pilot-scale O₂/CO₂ combustion tests of **Task 2** (experimental test performance) has been completed in Q2 2004. Process simulation and cost assessment of **Task 3** (Techno-Economic Study) has been completed in Q1 2005. The topical report on **Task 3** has been finalized and submitted to DOE in Q3 2005. The calculations of **Task 4** (Retrofit Recommendation and Preliminary Design of a New Generation Boiler) has been completed in 2004. In **Task 6** (engineering study on retrofit applications), the engineering study on 25MW_e unit has been completed in Q2, 2008 along with the corresponding cost assessment. In **Task 7** (evaluation of new oxy-fuel power plants concepts), based on the design basis document prepared in 2005, the design and cost estimate of the Air Separation Units, the boiler islands and the CO₂ compression and trains have been completed, for both super and ultra-supercritical case study. Final report of Task-7 is published by DOE in Oct 2007.

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DESCRIPTION OF THE TASKS

1 TASK 1: SITE PREPARATION

The demonstration part of this technology was performed in a 5 million Btu/hr Small-Boiler Simulator (SBS) located in Alliance, Ohio. The air fired SBS unit has been converted to oxygen firing by adding recycle flue gas, oxygen supply system, condensing heat exchanger, safety interlocks etc.

2 TASK 2: COMBUSTION AND EMISSIONS PERFORMANCE OPTIMIZATION

All the tests scheduled in this project have been performed in 2003 and Q2 2004. Two sets of tests were performed utilizing PRB coal. The first set of tests were performed as a preliminary step to help us gain experience in introducing oxygen into the primary air zone safely and to prepare the way for full oxygen firing. The second set of tests is the main objective of this experimental campaign, enabling measurements of combustion and emissions performances under O_2/CO_2 conditions. The goal here was to gain a better understanding of oxygen-enriched operation of the entire boiler.

3 TASK 3: TECHNO-ECONOMIC STUDY

The objective of this task was to perform a techno-economic assessment of the oxy-combustion process in pulverized coal boilers (PC-OC), and compare it with two other CO_2 capture technologies: the MEA process in pulverized coal boilers (PC-MEA) and the Selexol process in IGCC plants (IGCC-S). A conventional air-blown pulverized coal (PC) power plant without CO_2 capture was selected as a baseline for cost analysis of CO_2 capture.

4 TASK 4: CONCEPTUAL BOILER DESIGN

This task includes two sub-tasks “Subtask 4.1 Recommendations for retrofit applications” and “Subtask 4.2 Preliminary Design of a New Generation Boiler”. Based on the information collected during the experimental part of the project, the objective of the subtask 4.1 is to provide recommendations on retrofitting an existing air-fired unit into an oxygen-fired unit with flue gas recycle. The objective of the subtask 4.2 was to determine the optimum configuration of a new oxygen-fired boiler with flue gas recycle. The minimum recycled flue gas required to reduce excessive flame temperature and to be compatible with today’s boiler tube materials and targeted steam production was assessed. Also, the optimum operating conditions needed to reduce or avoid air infiltrations were evaluated. The boiler conceptual design and some information on the related costs are provided.

5 TASK 6: ENGINEERING, PLANT DESIGN AND COST ASSESSMENT FOR RETROFIT APPLICATION

The objective of task 6 is to conduct engineering studies and plant design activities necessary to assess the technical characteristics and the budget requirements for a field demonstration of the technology on an existing 25MW_e pulverized coal-fired power plant. The objectives were, 1) to develop a viable design for a large-scale pilot retrofit design that will demonstrate that oxygen combustion for concentrating CO₂ can be applied to an existing power plant as well as a new plant, 2) develop the primary control philosophies that will allow it to be operated in a similar manner to conventional air firing, 3) develop a basic test plan, 4) determine the overall cost for the retrofit including the cost of equipment, installation, and commissioning followed by three (3) months of testing.

6 TASK 7: OXY-FUEL COMBUSTION SYSTEM ANALYSIS FOR NEW POWER PLANT APPLICATION

Task 7 extends the current scope of work to further assess the economics of the oxy-coal technology at commercial scale (550 MWe) when applied to higher efficiency steam cycles, as available for today's or tomorrow's new coal-fired power plants (super-critical and ultra super-critical). This task of the project was lead by Parsons Group and the results were published by DOE in a separate report.

APPENDICES

- 1 TOPICAL REPORT ON TASKS 1, 2 AND 4
- 2 TOPICAL REPORT ON TASK 3
- 3 TOPICAL REPORT ON TASK 6
- 4 TOPICAL REPORT ON TASK 7

Advanced, Low/Zero Emission Boiler Design and Operation

TOPICAL REPORT

TASK 1: Site Preparation

TASK 2: Combustion and Emissions Performance Optimization

TASK 4: Recommendations for Retrofit Applications and
Preliminary Design of a New Generation Boiler

Work Performed Under Contract No.: **DE-FC26-02NT41586**

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ABSTRACT

This report describes a promising oxygen-based (oxy-combustion) process enabling CO₂ capture from coal-fired power plants. The technology involves the replacement of the combustion air by pure oxygen dilution in the recycled flue gas. It is also applicable to existing pulverized coal-fired boilers with minimal pressure-part modifications. The oxygen and recycled flue gas levels are adjusted to achieve flame and furnace temperatures compatible with today's materials.

In this program, the oxy-combustion of a Powder River Basin (PRB) coal was successfully demonstrated in a 5-million Btu/hr pilot combustor. The results were compared to air-blown combustion. Various burner parameters were varied to obtain maximum benefits from the oxygen/flue gas configuration. The overall combustion characteristics were comparable to air firing. The NO_x emissions from oxy-combustion tests were significantly lower by nearly 65% when compared to the air-blown combustion, and the CO₂ content in flue gas was increased from 15% to 80% in O₂-fired mode. With oxy-combustion, the flue gas volume exiting the boiler was reduced by nearly 70%, thereby improving the economics of efficient capture, reuse, and sequestration of carbon dioxide. This report describes the experimental results of oxy-combustion from a pilot-scale boiler. Oxy-combustion effects on flame shape, length, temperature, and emissivity are also discussed. In addition, the heat transfer variations in the boiler and the convection pass are reviewed. Finally, the highlights of an engineering feasibility study for the application of oxy-combustion technology to existing and new boilers are summarized.

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EXECUTIVE SUMMARY

The main achievements in tasks 1, 2 and 4 are as follow:

Task 1 (Site Preparation) - An air-fired 1.5 MW_{th} boiler was retrofitted to enable O₂/CO₂ combustion capabilities.

Task 2 (Pilot-Tests performance) - Four weeks of oxy-combustion tests were performed (2003 & Q2 2004) to characterize and optimize coal combustion in an O₂/CO₂ environment versus air-fired operation. Performance tests included emissions (CO₂, NO_x, CO, Mercury), heat transfer, and flame characterization as described below:

- **Burner stoichiometry, Flame stability and NO_x emission:** To avoid getting a cooler and less stable flame under O₂/CO₂ oxidizing atmosphere vs. air, with the same burner, a higher burner stoichiometry will be necessary in the oxy-combustion mode. This is due to the higher molecular weight and heat capacity of CO₂ vs N₂. Increasing the burner stoichiometry in oxy-combustion is feasible without affecting the NO_x level since the recycled flue gas keeps the NO_x emission very low (<0.1 lb/ 10⁶Btu) through the 'reburn' mechanism.
- **Heat Transfer Analysis:** Heat transfer characterization has been performed using measurements of flame emissivity, flame gas temperature (FGT), furnace exit gas temperature (FEGT), and convective pass exit temperature (CPEGT). Measurements have been performed for both air-firing and oxy-firing conditions. The measured emissivity indicates that the flame emissivity under the oxy-combustion is similar to the flame emissivity under normal air firing conditions. Pilot results seem to indicate that under optimized conditions (including keeping constant the overall mass flow rate), the gas temperature profiles are similar in oxy-firing and in air-firing conditions (previous tests showed 70°F lower average FEGT in oxy-firing).
- **Mercury Measurement:** Mercury sampling was performed at the convection pass outlet of the pilot-scale boiler under normal air-firing and oxy-firing conditions. The samples collected during the past reporting period have been analyzed at Western Kentucky University. The analytical chemistry and data reduction and analysis indicates a lower ratio of elemental to total mercury content of the flue gases from the oxy-fired process vs the air-fired baseline. If confirmed, such characteristic of the oxy-combustion process should enable easier capture of the mercury.

Task 4 (Retrofit Recommendation and Preliminary Oxy-Boiler Design) – This activity was completed in 2004, providing a conceptual boiler design for coal-firing with oxygen rather than air. The recommendations for retrofit application have also been discussed and reported.

1. INTRODUCTION

The key to a strong, stable, and secure economy is readily available, reasonably-priced energy. To maintain economic competitiveness and meet growing energy demands, the U.S. must improve the utilization of its domestic resources. Our vast resources of coal will play a strategic role in electric power production. Domestic coal is currently estimated to provide a 200-year supply of inexpensive energy at current production rates. Coal currently provides more than 50% of the United States electricity. Hence, coal will continue to be a prime source for electricity generation. Over the last three decades, utilities have implemented emission control equipment to control NO_x , SO_2 and particulate emissions to a large portion of coal-fired boilers.

Fossil fuel combustion is also a primary contributor of increased Greenhouse Gas (GHG) emissions. About one third of carbon emissions in the United States come from power plants (mostly coal-fired), one third from transportation and the remaining one third from industrial, commercial and residential sources. As fossil fuels continue to be the dominant fuel source for electricity generation, reducing carbon emissions by capturing and sequestering CO_2 from utility boilers is being considered to reduce these emissions and thereby control overall atmospheric GHG emissions.

Oxy-combustion, described below, is one combustion concept designed to greatly facilitate the capture and subsequent sequestration of CO_2 from fossil fuel power plants. The approach is applicable to both new and existing plants. One variation for the Oxygen-Fired Boiler concept is depicted in Figure 1. Oxygen is mixed with recycled flue gas to replace the normal combustion air at the burner. The oxygen is supplied by an on-site air separation unit. The flame temperature is maintained within 100°F above conventional PC boilers. High first stage temperatures are not desirable because this promotes thermal NO_x formation. The volume of flue gas leaving the boiler is considerably smaller than the conventional air-fired volume, and consists primarily of CO_2 . This concentrated stream of CO_2 is then available for capture and subsequent carbon sequestration without the need for expensive and energy-consuming separation systems.

Figure 1: Oxygen-Fired Boiler

An added benefit of the process for coal-fired systems is that it provides reduced NO_x emissions from the boiler. Thermal NO_x is reduced since there is less molecular nitrogen in the “combustion air.” In addition, some of the NO_x in the recycled flue gas will be reduced to molecular nitrogen via reburning by hydrocarbon radicals in the flame.

The oxy-combustion technology development initiative offers a variety of alternatives. To develop a retrofit technology applicable to existing boilers, a portion of the flue-gas is recycled thus replacing the nitrogen with CO₂ and keeping the boiler characteristics and dimensions the same as the air-fired base-case. For new plants, the volume of flue gas recycle (FGR) would be set to a minimum value enabling a more compact boiler design and resulting in reductions in boiler costs.

This report documents the work performed in Tasks 1, 2 and 4 of the DOE Contract No.: **DE-FC26-02NT41586**. The general objective of this project was to develop the oxy-combustion process to assess its performances as a clean and efficient process for CO₂ capture from pulverized coal-fired power plants. The report describes the experimental results of oxy-combustion from a 5-million Btu/hr pilot-scale boiler simulator. Effects of oxy-combustion on flame shape, length, temperature, and emissivity are discussed. Heat transfer variations in the boiler and the convection pass are also reviewed.

2. TASK 1 & 2: Site Preparation and Test Performance

Objectives

In partnership with the U.S. Department of Energy’s National Energy Technology Laboratory, B&W and Air Liquide are developing and optimizing the oxy-combustion process for retrofitting existing boilers as well as new plants. The main objectives of these tasks are to: (1) demonstrate the feasibility of the oxy-combustion technology with flue gas recycle in B&W’s 5-million Btu/hr coal-fired pilot boiler, (2) measure its performances in terms of emissions and boiler efficiency while selecting the right oxygen injection and flue gas recycle strategies, and (3) perform a feasibility study for application of the technology in commercial boilers. The objectives of task 1 and task 2 are to assess and optimize the performances of the Pulverized Coal Oxy-Combustion (PC-OC) process at “pilot-scale” test: 1.5MWth boiler simulator.

2.1. Task 1: Site Preparation

The demonstration part of this technology was performed in B&W’s 5 million Btu/hr Small-Boiler Simulator (SBS) located in Alliance, Ohio (Figure 2). Modular construction allows the examination of flame shape and stability, time/temperature history, multiple burners, staged combustion using overfire air (OFA) ports, flue gas recycle (FGR), fuel staging or reburning. Other variables in the combustion process can be examined in this facility as well. An indirect coal feed system was used. The indirectly fired system consists of a mass flow bin for dry pulverized fuel and a weigh belt feeder.

Figure 2: B&W Small Boiler Simulator (SBS)

This facility accurately replicates the combustion and convection heat transfer characteristics of a full-sized boiler. The convection pass section consists of four different sections simulating secondary superheater, reheater, primary superheater, and economizer. The convection pass is connected to an atmospheric pressure steam drum and the total heat transfer to the convection bank was inferred from the water flow rate during these tests.

A video camera is available for recording flame pictures and a FlameviewTM imaging system is available for temperature mapping of the flame. In addition, there are observation and sampling ports throughout the furnace so that in-furnace mapping and temperature can be measured. In this project, furnace exit gas temperatures were measured by high-velocity thermocouple probes for air-blown combustion and oxy-combustion tests.

The unit has a complete gas analysis system consisting of gas analyzers for O₂, NO_x, CO, CO₂, and SO₂ measurements. The flue gas is drawn from the convection pass exit by a vacuum pump, filtered, transported in a heated sample line to the refrigeration unit, and passed to the appropriate analyzers. The stack flyash can be isokinetically sampled and analyzed for carbon content to determine carbon utilization in the system. Major components of the oxy-combustion system are described next.

2.1.1. Process Design on the SBS Pilot Boiler for oxy-combustion operation

A process design schematic incorporating major equipment and instruments was developed. The corresponding modifications of the existing boiler were implemented. Appendix A shows the small boiler simulator (SBS) and the wet scrubber in an oxygen-fired mode. Appendix B illustrates the new equipment and ductwork (in red). Appendix C shows the last update of the oxygen supply system (the oxygen skid), as on June 2003.

2.1.2. Flue Gas Recycle (FGR) line to the Primary Air (PA) fan

This line includes an orifice, two pressure transducers, a thermocouple (TC), a control valve, and a remote controller to control and measure the flow rate and temperature of the FGR.

2.1.3. Air inlet modifications to the PA fan

It includes measurement and control equipment similar to FGR line.

2.1.4. Condensing Heat Exchanger (CHX) bypass line

A condensing heat exchanger (CHX) was used to cool down the recycled flue gas before it was mixed with oxygen and supplied to the primary and secondary oxidizer streams. As part of this project, an on/off manual valve was installed to enable switching from dry recycle process to wet recycle process and vice versa.

2.1.5. Oxygen supply to the boiler

Liquid oxygen was delivered to the test facility and stored on site in a 9000-gallon oxygen tank. The liquid oxygen is vaporized in an ambient vaporizer sized for winter operating conditions (-7°C/20°F air temperature). The oxygen is regulated to an appropriate pressure and delivered to the test area via a copper line. The oxygen delivery control system is integrated with the boiler simulator, allowing oxygen flows to be controlled and monitored from a central location. The oxygen is supplied to primary stream, secondary stream, and overfire zones. The oxygen was routed to the primary air line (after Primary stream fan) through an Oxynator™. During this project, the existing oxygen for the front overfire ports was modified to introduce oxygen into the burner. Also, the oxygen supply to the overfire ports and secondary stream Oxynator™ was modified. The electrical wiring was completed on site. The components of the oxygen valve skid & oxygen line were cleaned for oxygen service. The required additional piping was installed, as well as the new Oxynator™ in the PA line. The control system distributing the oxygen flows to various zones and monitoring the safety was developed and installed.

2.1.6. Two oxygen sensors

The oxygen levels in the primary and secondary streams, convection pass exit gas, and CHX exit flue gas, are measured with oxygen sensors. In this project, we added two oxygen sensors in the primary air/oxidant and recycled flue gas to account for air infiltration.

2.1.7. Safety interlocks

High oxygen concentration in the primary air line was considered a safety hazard. Safety interlocks provided the operator with an alarm at 25% oxygen concentration in the PA line for shutting down the oxygen and fuel, if the oxygen concentration reached 28%.

2.1.8. Modified coal feed system

Following a total system analysis performed, it was decided to modify the coal feed system to reduce the air inlet to the unit. Previously a compressed air aspirator was used for feeding pulverized coal to the burner. This system was modified by using the recycled flue gas in the primary line for PC pickup and transport. Two truckloads of PRB coal were received and stored. Twenty (20) tons of coal were consequently partially dried, pulverized and stored in the SBS facility.

2.1.9. Burner

A scaled-down version of the B&W DRB-4Z[®] low-NO_x burner was modified for oxygen firing and used in these tests (Figure 3). The burner was designed for staged combustion operation. This burner design has been installed in numerous commercial units with excellent results. The burner was modified to receive oxygen in the coal nozzle through a lance or through the transition zone of the burner. Three lances were fabricated for optimization.

Figure 3: B&W DRB-4Z[®] low-NO_x Burner

2.1.10. Data acquisition software

This system was modified for oxygen enriched FGR to the primary, secondary or tertiary air lines (versus air). The software enables us to collect the data and to provide on-line operational data.

2.2. Task 2: Combustion and Emissions Performance Optimization

All tests scheduled in this project were performed in 2003 and during the second quarter of 2004.

2.2.1. Coal characteristics

Testing was performed in the SBS using a PRB coal purchased from DTE Company. The coal analysis was performed by ISGS, and the results are shown below:

Table 1: PRB Coal Analysis

Moisture (As Received)	26.85 %
Ash (dry)	6.29 %
Volatile (dry)	47.20 %
Carbon (dry)	72.21 %
Hydrogen (dry)	5.00 %
Nitrogen(dry)	0.92 %
Sulfur (dry)	0.41 %
BTU (dry)	12,505 Btu/lb

2.2.2. Test matrix

Two sets of tests were performed as described below:

1. Optimization of oxygen injection in the primary combustion zone

This first series of tests were performed as a preliminary step to help us gain experience in introducing oxygen into the primary air zone safely and to prepare the way for full oxygen firing. The oxygen in the primary zone could be added through an OxynatorTM into the primary air line (premixing of oxygen in flue gas and pulverized coal), or at the burner through a lance or through the transition zone (annulus around the coal pipe).

2. Full-oxy combustion with flue gas recycle.

This second series of test is of course the main objective of this experimental campaign, enabling measurements of combustion and emissions performances under O₂/CO₂ conditions. Our goal here was to gain a better understanding of oxygen-enriched operation in the entire boiler.

2.2.3. Principle of flame radiance measurement

The total radiation from the flame and the spectral radiance from the flame were measured in these tests. The total radiation emitted from an area of the flame was measured with a thermopile detector. Thermopile detectors have a spectrally flat response to thermal radiation over the wavelength range of about 0.2 to 50 micrometers. The detector approximates a “grey body” in which the absorptance (emittance) is constant with wavelength. A protective window reduces the responsive wavelength range to about 0.6 to 18 micrometers. For temperatures in the range of 1500°F to 3000°F, this wavelength range contains about 95% or more of the total thermal radiation.

The spectral radiance was measured using a computer controlled scanning monochrometer. The monochrometer is equipped with multiple photodetectors and gratings that can span a measurement range of about 0.2 to 15 micrometers. Silver coated first surface mirrors are used to direct the flame radiation to the thermopile and to the scanning monochrometer. The light collection mirror is an off-axis parabolic mirror fastened to a rotating mount. This arrangement permits selection of thermal radiation from the blackbody calibration source or the flame by simply rotating a single mirror.

The equipment is shown in Figure 4. The thermopile detector is mounted at the same optical height as the monochrometer, and so can be simply inserted in the path of the collected light to make a measurement. The light entering the monochrometer is modulated with a chopper wheel, and the electronic signal from the photodetectors is amplified and processed with a lock-in

amplifier. The lock-in amplifier and monochromator are interfaced to a computer which operates the equipment.

Figure 4: Photograph of the Thermopile Detector, Scanning Monochromator and Light Collection Mirrors

For the combustion tests, the equipment was mounted on a hydraulic table and positioned adjacent to a furnace viewport. The viewport sight glass was replaced with a metal flange equipped with a 1½ inch pipe nipple in the center. This arrangement allowed a view of the flame that was unobstructed by a window, could be easily opened and closed, and could be purged with only small amounts air.

The measured signals from the thermopile detector and scanning monochromator were standardized to an 1835°F blackbody at the start and end of a test. The standardization permitted evaluation of the total and spectral thermal radiant heat flux from the flame. Flame temperature was measured using a two-color pyrometry. A hand-held pyrometer was used to measure flame temperature using the same viewport as the radiance measurements. Flame temperature was also measured using a Diamond Power Specialty Corp. Flameview. The Flameview uses video images filtered at two wavelengths to determine a spatially resolved flame temperature map.

With the flame radiance and flame temperature, the total emittance and spectral emittance of the flame was calculated.

2.2.4. Principle of mercury measurements

Mercury speciation sampling was conducted during the test program. The Ontario Hydro Method (OHM) was used to obtain mercury speciation profile at the convection pass outlet (hereafter, CPO) of the SBS:

- Elemental mercury (Hg^0),
- Oxidized mercury (Hg^{2+}),
- Particulate mercury (Hg^P),

The total mercury (Hg^T) is defined by $Hg^T = Hg^0 + Hg^{2+} + Hg^P$.

OHM sampling campaigns were performed on air-fired baseline and oxy-combustion conditions. Each sampling campaign was performed in duplicate. The OHM sampling also provides the particulate matter (PM) concentration and the moisture content of the flue gas at the point of measurement. PM concentration is determined using the OHM filter pre- and post-test weights. The OHM impinger weights (pre- and post-test) are used to determine the moisture content of the flue gas.

2.2.5. Results and discussion

Technical Feasibility of the Oxy-combustion Process and Switching procedure from air-fired to oxygen-fired operation

As mentioned earlier, an existing pilot-scale boiler was modified to enable oxygen combustion in flue gas recycle. A safe and smooth transition from air to oxy-combustion was achieved. Transitioning from primary air to oxygen-enriched flue gas was initially achieved with the addition of oxygen only through a burner lance. After full primary air substitution with flue gas and oxygen introduction through the lance, a portion of oxygen was removed from the lance and reintroduced in the primary air line via an OxynatorTM. The oxygen to the secondary stream and overfire port was introduced through another OxynatorTM.

CO₂ Capture Capability: flue gas volume and CO₂ content

One area of concern during the current series of tests was the amount of air infiltration into the unit. During these tests, the maximum CO₂ concentration that could be obtained was 80% versus 15% in air-firing (on a dry basis, corrected to 3% boiler exit oxygen concentration). The infiltrated air was approximately 5% of the total boiler oxidant requirement (air, oxygen, flue gas). The SBS boiler is a balanced draft that operates under slightly negative pressure. During the tests, the furnace pressure was increased to slightly positive to minimize air leaks into the boiler. The main sources of air leaks are suspected to be the I.D. fan, baghouse and scrubber that were operating under higher negative pressures. This level of air leakage is a good representation of potential air leaks in commercial boiler retrofits from the ESP, air heater, etc. However, for retrofit units, there are other alternative boiler upgrade and operating procedures that could be employed to reduce air leaks. Since the flue gas volume for oxy-combustion is reduced by 80% (vs air-fired operation), the CO₂ purification is highly facilitated.

Overall Combustion Characteristics in O₂/CO₂ environment – burner stoichiometry

Based on visual observations and 2-color pyrometry (Flameview) measurements, the oxy-combustion flame was judged to be cooler than the air-fired flame. During these tests, the coal feed system had been improved providing a very smooth combustion condition with very small convective pass exit O₂ fluctuations. As we switched from air firing to oxy-combustion while maintaining a flame stoichiometry of approximately 0.85, the combustion was affected negatively, judging from the high CO levels observed. We knew that the burner velocities were lower since nitrogen was substituted by CO₂. Since one burner was used for start-up on air, transitioning to oxygen, and full oxy-combustion, the throat velocity and burner stoichiometry could not have been simultaneously maintained. Therefore, we increased the burner stoichiometry from 0.85 for air-fired operation to 1.0-1.05 during oxy-combustion to increase the burner velocity and flame temperature. The combination effects of higher burner stoichiometry and velocity resulted in a more stable and brighter flame. Staged oxy-combustion for controlling

the NO_x emissions was less sensitive to burner stoichiometry variations (in the range of 0.8 to 1.05) than normal air firing.

NO_x Emission Characteristics

Measured NO_x emissions were approximately 65% lower in oxy-combustion conditions than the baseline (See Figure 5) data. During the baseline tests at the design stoichiometry of 0.9, the NO_x level was $0.28 \text{ lb}/10^6 \text{ Btu}$. The baseline NO_x emission ranged from 0.22 to $0.4 \text{ lb}/10^6 \text{ Btu}$ when burner stoichiometry varied from 0.57 to 1.21. By contrast, during the oxy-combustion tests NO_x levels ranged from 0.065 to $0.106 \text{ lb}/10^6 \text{ Btu}$ as the burner stoichiometry varied from about 0.8 to 1.05. Overall mass flow rates through the boiler for air firing and oxygen firing conditions were similar.

Under oxy-combustion conditions, NO_x emissions are impacted by the burner stoichiometry, oxygen flow rate into the primary air zone, and by the overall flue gas recycle rate (See Figure 6). This can be explained by higher flame temperature resulting from increased O_2 content in the primary air zone of the boiler or from lower flue gas flow rate. Higher temperatures in the reducing zone of the boiler promote the conversion of recirculated NO_x and devolatilized fuel nitrogen to molecular nitrogen. Since the flue gas recycle flow rates in the primary and secondary burner zone are controlled by various practical considerations (pulverizer operation and heat absorption for steam generation), NO_x formation can be reduced primarily by optimizing the primary zone oxygen distribution as well as the burner stoichiometry in such a way that a desirable flame shape is maintained.

Figure 5: NO_x Emissions

Figure 6: Impact of FGR Flow Rate on NO_x Emissions

Temperature Profiles and Heat Absorption

Oxy-combustion was originally thought to change the heat transfer characteristics to the point that addition of extra heat transfer surfaces would be necessary to generate steam. In order to assess the heat transfer effects, two tests were performed at the optimum combustion conditions. Two firing conditions were selected for air-blown combustion and oxy-combustion for exploring the effect of oxy-combustion on the boiler temperature profiles, and overall heat absorption in the furnace. In these two cases the boiler load, overall mass flow rate through the boiler, and excess oxygen were the same. Flame temperatures, furnace exit gas temperature, and convection pass exit temperature were measured in both tests. The total heat absorption in the convective pass was estimated from the intake water flow rates to the drum in the period of 8 hours. The results are illustrated in Table 2. The boiler temperature profiles were similar for both cases. Oxy-combustion flame temperature, measured with 2-color pyrometer, was 2480°F; about 100°F more than air firing combustion conditions (2370°F) presumably due to different burner stoichiometry and flue gas molecular weight. The convection pass exit was very similar indicating that slightly more heat was transferred from gas to water in the convection pass. This agreed well with the convection pass heat absorption, measured by intake water to steam drum. The higher convection pass heat absorption under the oxy-combustion can be explained by higher residence time available for heat transfer to the tube banks since volume of flue gas is smaller (higher molecular weight) in oxy-combustion tests. The smaller flue gas volume in the oxy-combustion conditions also reduces the flue gas velocity and reduces the heat transfer coefficient between flue gas and steam. But in this pilot, the effect of higher residence time is more than the lower heat transfer coefficient. In order to further explain these results, participants performed flame temperature and emissivity measurements.

Table 2: Temperature and Heat Transfer Measurements

	Air Firing	Oxy-combustion
Boiler Mass Flow Rate	4,167 lb/hr	4,030 lb/hr
Burner Stoichiometry	0.86	1.05
FEGT	1797 F	1880 F
Convection Pass Temperature	555 F	547 F
Convection Pass Heat Absorption	43.6%	46.8%

It should be mentioned that these results are impacted by the oxy-combustion burner stoichiometry. In a different test, the oxy-combustion temperature profiles under substoichiometric conditions showed that the average furnace exit gas temperature, under oxy-combustion conditions, was slightly lower than for the air firing case. The convection pass exit gas temperature was measured for both oxy-combustion and air cases. It was generally higher in oxy-conditions. Further pilot and/or full-scale studies are required to characterize the boiler heat transfer and steam generation properties. However, the results indicate that additional heat transfer surface may not be required.

Flame Radiance Results

Optical measurements of flame radiation were recorded at two test conditions representing conventional air-fired coal combustion, and oxygen-fired coal combustion. Several measurements were made at each test condition. The average spectral emittance for air-fired and oxygen-fired combustion over the range of 0.6 to 14 microns is shown in Figure 7.

Figure 7: Average Spectral Emittance for Air-fired and Oxygen-fired Combustion Over the Range of 0.6 to 14 Microns

The magnitudes of the spectral emittance for the two modes of firing are nearly identical. Differences between air and oxygen firing are most pronounced at wavelengths of about 2.6 μm , 4.3 μm , and for wavelengths from 12 to 14 μm . These wavelengths correspond to wavebands specific to CO_2 . This is also the likely reason that the measured emittance exceeds a value of 1 at wavelengths of about 4.6 and 13 μm . The similarities in spectral emittance for the two firing conditions indicate that thermal radiation is primarily dependent on the coal, flyash, and soot particles in the flame, and not on the concentration of CO_2 .

The total emittance that corresponds to the spectral data in Figure 7, and the total emittance measured using the thermopile detector are given in Table 3, along with the average flame temperatures.

Table 3: Total Emittance and Average Flame Temperature in Air and Oxygen Fired Conditions

Measurement Method	Combustion Condition	Average Temperature (F)	Total Emittance		
			Average	Maximum	Minimum
Thermopile	Air Fired	2370	0.69	0.70	0.69
	Oxygen Fired	2480	0.68	0.71	0.64
Integrated Spectra	Air Fired	2370	0.78	0.79	0.75
	Oxygen Fired	2480	0.78	0.80	0.77

As shown in Table 3, the total emittance using each measurement method is consistent, but the two methods differ by about 10%. The most likely cause for the difference is a fixed error due to slight misalignment of the optics. Because of the procedure used to perform the measurements, the thermopile measurements of total emittance are probably the most accurate on an absolute

basis. Table 3 also shows that the average temperature for the oxygen-fired combustion is about 100°F hotter than for air fired combustion.

Mercury Results

Two mercury speciation sampling campaigns using OHM were performed:

- 1) Air-fired baseline conditions
- 2) Oxy-combustion conditions.

The objective of these tests was to determine if oxy-combustion process changes the speciation of mercury. These measurements were performed at the convection pass exit prior to all environmental equipment including the baghouse and condensing heat exchanger which reduces the SO₂ and mercury emissions. The total mercury emissions from the unit were not measured.

Each sampling campaign was performed in duplicate.

Mercury, PM and moisture flow rate and concentration

Elemental mercury (Hg⁰), oxidized mercury (Hg²⁺), particulate mercury (Hg^P), and total mercury (Hg^T) concentration results are summarized in Table 4, along with particulate matter (PM) concentration and moisture content of the flue gas at the point of measurement.

Table 4: Mercury Speciation (in µg/dscm), PM Concentration (in mg/dscm), and Moisture Content (in %) at the CPO; Corrected to 3%O₂

Test	Hg ⁰	Hg ²⁺	Hg ^P	Hg ^T	% Hg ²⁺	PM	moisture
Baseline 1	2.5	1.6	0.4	4.4	36	2991	13.8
Baseline 2	1.6	1.9	0.3	3.9	50	2241	10.9
Baseline, av.	2.1±0.5	1.8±0.1	0.3±0.0	4.1±0.2	43±7	2616±375	12.4±1.5
Oxy 1	2.2	3.2	0.3	5.6	56	4227	19.9
Oxy 2	3.1	2.9	0.1	6.1	48	4828	21.9
Oxy, av.	2.7±0.5	3.1±0.1	0.2±0.1	5.9±0.2	52±4	4528±301	20.9±1.0

When analyzing the mercury results, one should realize that generally the concentrations of most emissions in the oxy-combustion process increase because of the absence of nitrogen. Some emissions decrease (like NO_x) as explained before. However, the total mass flow rates of these emissions do not increase. For example, while CO₂ concentration increases in the flue gas the total mass flow rate of CO₂ stay the same because it depends on the mass flow rate of coal that is being fired. The mercury results discussed below are not an indicator of total mercury emissions and only provide insight to mercury speciation. . Figure 8 and 9 illustrate the Hg emissions at convection pass outlet before all environmental control equipment (see Appendix B for the equipment layout).

Figure 8: Mercury Speciation Results for the Baseline and Oxy-combustion Processes

Figure 9: Mercury Emissions at the Convection Pass Exit

As expected, due to the high temperature of the flue gas at the CPO (521-555 °F; 272-291 °C), very little mercury was associated with PM at this location (less than 0.5 µg/dscm; reported

detection limit of OHM is about 0.1-0.5 $\mu\text{g/dscm}$). Duplicate experiments indicate a very good experimental precision for Hg^{T} and Hg^{2+} . Hg^0 measurements exhibited larger variations.

The oxy-combustion process resulted in higher mercury concentrations in the flue gas. The Hg^{T} level in the flue gas was concentrated by a factor of 44%. It appears that the Hg^{2+} level in the flue gas was concentrated by a higher factor than that of Hg^0 (72% vs. 29%). Due to the insolubility of Hg^0 , a higher degree of concentration in the Hg^0 content of the flue gas is expected during the oxy-combustion recycle process. Hg^{2+} is expected to be partially removed from the recycled flue gas in the condensing heat exchanger. This removal should result in a lower accumulation in the Hg^{2+} content of the flue gas. More studies are needed to determine the effect of oxy-combustion process on the concentration of Hg^0 and Hg^{2+} contents in the flue gas. It should be noted that the OHM measurements were performed upstream of the particulate collection device. At this location, the filter of the OHM may introduce a speciation bias in the flue gas mercury profile. In order to minimize these biases, the filter of the OHM was maintained at the CPO flue gas temperature. The high filter temperature should have prevented the adsorption and further reaction of Hg^0 and Hg^{2+} over the filter of the OHM. However, at locations upstream of particulate control devices, the speciation results as reported by the OHM should be interpreted with great caution. The total mercury results as reported by the OHM are valid and reliable.

As expected, the PM concentration and the moisture content of flue gas were concentrated during the oxy-combustion process by a factor of 73% and 69%, respectively.

Mercury Mass Balance Evaluation for the Baseline tests

Based on the coal mercury input, a mercury mass balance across the furnace of the SBS can be carried out. The ultimate and proximate analysis of the Black Thunder PRB coal (as fired) used in this study on as determined basis is illustrated in Table 5.

Table 5: Ultimate and Proximate Analysis of the Black Thunder PRB Coal

Parameter	Value
Moisture, %	19.95
Ash, %	5.10
Volatile Matter, %	35.04
Carbon, %	55.87
Hydrogen	6.05
Nitrogen, %	0.78
Sulfur, %	0.28
Heating Value, Btu/lb	9452
Mercury, ppm	0.07
Chlorine, ppm	118

The input of mercury into the system is through the coal and the output of mercury at the CPO is by the flue gas. There are no other streams of mercury into the system for the baseline operation. The oxy-combustion process, however, includes other input mercury streams; the mercury contents of those streams were not measured during this test program. Thus, mercury balance for the oxy-combustion process could not be carried out. The results of mercury mass balance for the two tests conducted during the baseline operation are summarized in Table 6.

Table 6: Hg Mass Balance for the Baseline Operation of the SBS

Parameter	Test 1	Test 2
Heat input, 10 ⁶ Btu/hr	3.90	3.95
Calculated coal flow, lb/hr	413	418
Measured coal flow, lb/hr	419	410
Hg input with coal, mg/hr	13.1	13.3
Flue gas flow at CPO, daf, lb/hr	3727	3713
Volumetric flue gas flow at CPO, dscm/hr	1324	1315
Hg output at CPO, mg/hr	5.6	5.0
Hg output at CPO, lb/10 ¹² Btu	3.2	2.8
% Hg recovered at CPO	43%	38%

As indicated in Table 6, only about 38-43% of inlet mercury was recovered at the outlet of the convection pass. There are cold stainless steel surfaces (water tubes) in the furnace of the SBS. The surface temperature should be about 100°C (boiling water temperature). Mercury has a high affinity for adsorption by the residual carbon (LOI) of the fly ash at this low temperature. It is believed that prior to reaching a steady state of combustion (with regard to mercury emission), significant amounts of mercury can be adsorbed by the fly ash deposited on these cold surfaces. Weeks of continuous operation of the SBS may be required to achieve a steady-state mercury emission condition. Also, as indicated in Table 6, significant amount of Hg²⁺ was observed at the CPO. Typically, the combustion of PRB coals is believed to result in very low Hg²⁺ (less than 20%) in the flue gas. This is especially true at the high temperature locations such as the convective pass outlet. The oxidation of Hg⁰ to Hg²⁺ is a heterogeneous adsorption- and/or reaction-controlled process. Iron is believed to be an effective catalyst for such oxidation. The presence of cold stainless steel surfaces in the SBS provides an ideal condition for the adsorption/oxidation of Hg⁰. The high Hg²⁺ observed at the CPO may be due to the biases created by the cold surfaces.

As explained earlier, a mercury mass balance for the oxy-combustion tests is not possible because the mercury was measured at the CPO and not after the baghouse and CHX.

An X-ray Fluorescence (XRF) analysis was also performed on the ash portion of a coal sample obtained during this test program (See Table 7). It should be noted that the OHM and coal/ash analyses were performed by Western Kentucky University.

Unburned Combustibles

Generally, oxy-combustion can improve the unburned combustibles in the fly ash and CO emissions, especially with high volatile bituminous coals. However, the PRB coal is very reactive and the baseline unburned combustibles ranged only from 0.62 to 0.8%. Under oxy-combustion tests, unburned combustibles were similar ranging from 0.52 to 0.86%. Under the best conditions, CO levels ranged from 12 to 15 ppm at the baseline conditions versus 70 to 91 ppm under the oxy-combustion conditions.

Table 7: XRF Analysis of the Ash Portion of the PRB Coal Sample

Component	Weight percent
Na ₂ O	0.99
MgO	4.45
Al ₂ O ₃	4.93
SiO ₂	33.57
CaO	19.71
K ₂ O	0.46
SO ₃	10.48
P ₂ O ₅	1.02
BaO	0.47
SrO	0.27
Fe ₂ O ₃	5.58
MnO	0.01
TiO ₂	1.41

2.3. Conclusion of experimental tests

From the experimental investigations, the following conclusions can be derived:

- The oxy-combustion technology with flue gas recycle has been successfully demonstrated and characterized in a 5-million Btu/hr coal-burning pilot facility.
- The overall oxy-combustion characteristics were comparable to the air firing case in spite of enormous change of oxidant composition from air to oxygen-enriched flue gas.
- The NO_x emissions from oxy-combustion were significantly lower (65%) than the air-blown case.
- The steam generation and heat transfer in the furnace and the convection pass changed only modestly. In commercial application, site-specific feasibility studies will determine the boiler performance but based on this pilot-scale study no additional heat transfer surfaces are expected.
- The air infiltration in the boiler under O₂-conditions has been reduced to a final level of approximately 5% of the overall stoichiometry, increasing the initial CO₂ content in flue gas from 15% in air-fired mode to eventually 80% under O₂-fired condition.
- Alternative boiler upgrade procedures are expected to reduce even more the air infiltration to achieve higher CO₂ concentration in flue gas for further sequestration or reuse.
- The flue gas volume exiting the boiler has been reduced by 70%, thus making easier any additional flue gas treatment which may be necessary before stack exhaust or CO₂ reuse or sequestration.

3. TASK 4: RECOMMENDATIONS FOR RETROFIT APPLICATIONS AND CONCEPTUAL BOILER DESIGN

In this task, the participants of the project provide engineering data to help apply the PC-OC technology to both retrofit and new power plants. Two subtasks are included: Subtask 4.1 “Recommendations for Retrofit Applications” and Subtask 4.2 “Preliminary Design of a New Generation Boiler”.

3.4. Subtask 4.1 Recommendations for retrofit applications

3.4.1. Objective and deliverable of subtask 4.1

Objective

Based on the information collected during the experimental part of the project, the objective of this subtask is to provide recommendations how to retrofit an existing air-fired unit into an oxygen-fired unit with flue gas recycle. Recommended flue gas recycle rate in a retrofitted boiler is provided and plant adaptations enabling reduction of air infiltration and maximization of flue gas CO₂ content are described. Those recommendations are specifically helpful to demonstrate that the proposed PC-OC is a viable retrofit technology for an existing unit.

Deliverable

“The recipient shall provide the Government with a list of technical recommendations describing the best practices for retrofitting an existing coal-fired boiler with oxygen/flue gas operation. Those recommendations have to be in a format that is readily comprehensible to a design engineer and that may be directly applicable to a potential proposed demonstration”.

3.4.2. Modifications to boiler and auxiliaries

Although the retrofit of an existing coal fired boiler from air firing to oxygen firing with gas recycle can vary from plant to plant depending on existing equipment, firing method, and coal type, there are many commonalities. It is expected that since the equipment was originally designed for firing with air, the basic philosophy would be to maintain load capacity with either air or oxygen firing. Whether the plant is a PC unit or cyclone fired burning lignite to bituminous coals, the changes to existing equipment would be minimal.

Process Flow and Data

Figure 9 illustrates the process flow for a typical coal fired plant. Some equipment may or may not exist at any given plant but it illustrates the concept of retrofitting a plant for firing with oxygen and gas recycle.

Figure 10: Typical Process Flow Diagram

The flue gas from the electrostatic precipitator (3) flows to a combination of gas cooler and gas reheater. A portion of the flue gas is cooled slightly as it passes through the gas reheater after which it is recombined with the main gas flow (4) and is further cooled to approximately 130°F in the water quench cooler. This process condenses out a large amount of the moisture and produces a saturated gas. A portion of the gas leaving the cooler (5) then passes through a reheater where it is heated and recombined with the main gas flow (6) to slightly superheat it to around 145°F in order to prevent water and wet particulate from flowing to the rest of the system while providing a reasonable temperature to the fans and air heaters to minimize the impact on fan size as much as practical and maintain precipitator efficiency. The majority of the gas is then recycled through new forced draft (FD) (12) and primary air (PA) fans (7) to the existing air heaters and on to the boiler (11) (13) to control steam temperature as well to obtain a reasonable flame temperature. Although the original fans may be capable of the duty, it may be more practical to provide new fans to simplify startup and conversion downtime.

Oxygen is injected into the primary air stream between the airheater outlet and mills (15) to produce a gas stream to the mills with 18% O₂ by volume, on a wet basis. Oxygen is

also injected into the secondary air stream (16) to produce a gas stream to the windbox with 25% O₂ by volume on a wet basis. The remaining oxygen needed for combustion is injected directly into the boiler in a separate stream (14). The coal is combusted in the furnace using this gas mixture and is carried through the convection pass and to the economizer outlet. The gas is cooled in the airheater and flows to the dry ESP or fabric filter.

From the particulate removal device the gas may flow into a scrubber after which about 70% of the gas flow is recycled back to the boiler as described above. In order to maintain equilibrium, an amount of gas equivalent to the oxygen injected plus the products of combustion less the moisture removed in the cooler, about 30% of the total gas flow leaving the dry ESP, must be extracted from the gas side after the gas cooler and reheater (17). The gas pressure must be boosted (18) in order to convey it to downstream equipment that may process the CO₂ laden gas stream.

The boiler output would not change between the two firing schemes. The most notable difference is that the boiler efficiency would be lower with oxygen firing because the flue gas temperature leaving the airheaters must be slightly higher. Except for boiler efficiency there are no adverse effects expected on boiler or auxiliary equipment performance. Air heater modifications may be necessary in order to increase the boiler efficiency.

Furnace

No physical modifications are anticipated to the furnace to accommodate conversion to oxygen firing with flue gas recycle. There will however be a drop in Furnace Exit Gas Temperature (FEGT) due to the difference in emissivity of the flue gas because of the concentration of CO₂ and moisture. Gas residence time is also expected to be longer due to the density difference between air and O₂ firing.

Superheater, Reheater, and Economizer

The convection pass typically contains the heating surface for the superheater, reheater and economizer. When calculating the performance for the boiler convection pass, enough gas recycle will be used to avoid changes to the heating surface. This is expected to be practical since the flue gas has a higher density so that gas velocities will actually be less than with air firing.

Burners

If the unit is a PC unit, oxygen-enriched recycled flue gas combustion will affect the burner operation. The recycled flue gas has a higher molecular weight than air and will reduce the burner throat velocity. As mentioned earlier, the mixing between flue gas and oxygen, especially in the pulverizer and primary air lines, limits the amount of oxygen that can be added to the flue gas. This is a safety issues that affects burner operation.

Calculations show that in order to introduce enough oxygen to the boiler, additional oxygen beyond that required for combustion gas (normally secondary air) and primary gas (normally primary air) has to be introduced into the boiler. In other words, if the total mass flow rate (air or oxygen firing) into the boiler remains constant, and the design limit for the oxygen concentration in the flue gas is 25%, there is insufficient oxygen for

complete combustion of the coal and a relatively small amount of additional oxygen must be introduced into the boiler to achieve the desired stoichiometry.

This can be accomplished by either modifying the existing burners. However, for performance reasons, it may be better to retrofit the unit with new burners specifically designed for oxygen and air firing to optimize and enhance combustion and lower NO_x under either condition

Pulverizers

In PC units when firing with air, coal is dried in the pulverizers and conveyed through the burner lines with air. When firing with oxygen, hot recycled gas from the primary airheater is mixed with enough oxygen to have about an 18% content (by volume on a wet basis) flowing to the pulverizers. There are three considerations for determining how much recycled gas to divert to the mills.

1. Maintain an equivalent amount of heat in the gas mixture as was used during air firing. This would insure an equal amount of drying and mill outlet temperature.
2. Maintain a minimum 27 ft³ of gas per lb of coal in the mills for fineness and classifier operation.
3. Maintain a minimum of 3000 fpm burner line velocity for coal conveying.

It is anticipated that the second and third considerations would be the governing criteria since as little recycled gas going through the primary airheater as possible is needed. This will be explained in a section 3.6 of the report. As with most of the other boiler auxiliary equipment there are no anticipated changes needed to the mills.

Airheaters

Some plants have combined primary/secondary airheaters (Tri-Sectors) others have separate primary and secondary regenerative airheaters. The performance would be the same in either case as they would be controlled to heat enough recycled gas on the primary side to achieve desired mill performance and the remaining gas would flow to the secondary airheater. Enough flue gas on the gas side of the primary airheater is used to heat the primary stream to its desired temperature while maintaining a suitable gas outlet temperature. Gas and air side draft loss is also an important consideration.

The remainder of the flue gas would be used on the secondary 'air' side. Depending on the coal being fired, the only minor problem may be that there is too much flue gas on the gas side of the secondary or too little recycled gas flow, depending on the perspective, causing the gas exit temperature on the secondary side to increase when firing with oxygen. Since the secondary side is the largest portion of the total gas flow, it causes the mixed exit gas temperature to increase. This then results in a decrease in boiler efficiency. Although the balance changes slightly, which effects boiler performance, no changes to the existing airheaters are expected to be necessary.

Electrostatic Precipitator (ESP) or Fabric Filter and Air Infiltration

The existing ESP's or fabric filters are expected to maintain their current removal efficiency since the gas velocities are expected to remain the same or slightly less. The effect of flue gas composition on ESP performance needs to be addressed prior to a

retrofit. From experience it is known that an ESP will usually have approximately 5% air infiltration. To minimize air infiltration for oxy-fuel firing CO₂ can be used to pressurize the penthouse which will reduce air infiltration from that portion of the ESP by at least half.

Quench Cooler and Reheater

In order to maximize the airheater performance and minimize fan size, the recycled gas must be at a temperature and volumetric flow that is reasonably close to the air flows with air firing. To accomplish this, the flue gas leaving the ID fan must be cooled. The cool, saturated gas must then be reheated slightly to insure the gas to the fans is dry and corrosion is prevented.

Due to their interconnections and close physical proximity, the quench spray water cooler and reheater will be discussed together. Figure 10 shows a schematic process flow diagram and the location of the components within the gas process and their arrangement. Although a direct contact quench cooler is discussed here, it is recognized that a condensing heat exchanger (CHX) could also be employed. Another alternative may be a wet scrubber system. The wet scrubber would cool the gas and neutralize the water condensing from the gas stream. This would eliminate the need for disposing of caustic water. A wet scrubber may also be desirable to reduce SO₂ concentration in the recycle gas stream to prevent coal ash corrosion in the boiler.

Quench Spray Water Cooler

The cooler is a “contact” heat exchanger, wherein counter flowing streams of cooling water (entering at the top) and hot gas (entering at the bottom) mix in a vertical cylindrical vessel. In the direction of gas flow, this vessel will contain a tray (to help distribute the gas uniformly and enhance cooling), two spray water elevations, and a mist eliminator. The heated sour wastewater is discharged to separate plant systems for treatment. Based on B&W’s experience with wet scrubbers, the cooler shell could be carbon steel with a Stebbins-tile lining. Due to the corrosive environment within the cooler and reheater, the materials would be Hastelloy or equivalent for the reheater.

Due to the disadvantages of quench cooling, some of which are site driven, other means of cooling the gas, such as a condensing heat exchanger, should be considered. If a wet scrubber is available, it could substitute for the quench cooler in a retrofit situation. This would be the most desirable situation the water discharge would not be as caustic as the discharge from the cooler. Quench cooling the flue gas may not be the least costly option. The quantity and low pH of the water leaving this cooler may present a significant problem with obtaining the quantity, treatment and discharge.

Gas Reheater

The gas reheater receives hot flue gas from the main flue gas recycle that is routed from the existing ID fan outlet to the quench cooler inlet, cools it slightly while “reheating” a portion of the gas stream leaving the cooler by about that same amount. The design is essentially the same as a conventional shell-and-tube tubular air heater (TAH) but with materials selected to resist the corrosive environment.

Unlike a conventional TAH where one side has flue gas and the other has air, both sides of the reheater will have gas flow. In the following discussion, the term “cool-side” is used to denote the colder of the two gas streams that is taken from the cooler outlet flue (see Figure 10). The reason for including a reheater is to insure a slight “superheat” for the gas stream going back to the boiler island to protect the FD and PA fans.

The hot-side gas is taken from the flue before entering the quench cooler (inlet to the gas cooler) or wet scrubber. It flows down through the tubes and is returned to the flue prior to entering the quench cooler. The gas being heated, or cool-side gas, will be extracted from the quench cooler outlet flue, flow through the shell side and then be recombined with the cooler outlet flue gas. The design has two passes on the cool-side (shell side) and a single pass on the hot-side (tube side). Entry of the one and exit of the other cool-side pass will be on the same side of the reheater so there will be a return bend on the opposite side of the shell for this cool-side stream.

Due to the corrosion potential, the gas reheater tubes will be constructed of Hastelloy C-276 tubing or equivalent to provide a reasonable service life. The reheater shell would be carbon steel with an epoxy lining. Since this lining is field-applied, it is more expensive than stainless steel, but significantly less than Hastelloy. Service life of the components in the cooler-reheater and adjacent flues on the outlet side is a concern, particularly regarding out-of-service corrosion and a means of maintaining temperature to avoid condensation when the unit is shut down may be wise.

Due to the small heat transfer duty, most of the flue gas (75-85%) will bypass the reheater. In the bypass flues to and from each side of the reheater, there is a damper that can be used for either isolation or throttling, if needed. A perforated plate upstream of the control dampers provides sufficient resistance to allow the dampers to throttle the main cooler inlet and outlet gas flows and force the desired quantity of gas through the reheater.

The hot-side flues before the quench cooler are unlined carbon steel. The cool-side flues downstream of the spray cooler will also be carbon steel but with epoxy lining. Dampers and perforated plates may be Hastelloy on the cool side, but carbon steel on the hot side (cooler inlet side).

Fans

It is possible that the existing fans could be used but they would have to be reviewed individually. To provide redundancy and a possible auxiliary power reduction, it is anticipated that new fans would usually be added to the process for oxygen firing. New fans would be used for the secondary recycle stream in parallel with the existing FD fans, which will be used when firing with air. New fans would also be added for the primary recycle stream in parallel with the existing PA fans, which will be used when firing with air.

The addition of new FD and PA fans will also minimize the down time required for the installation of the new firing system by allowing the entire system to be installed, except for final flue tie-ins, while the existing unit remains operational. The boiler will start up firing with air using the existing fans and then be transitioned over to the new fans to fire

with oxygen. Due to the nature of the gas, special considerations are given to sealing to prevent leakage into the boiler building and the potential for out-of-service corrosion.

The existing induced draft (ID) fans, if applicable, will remain in service during oxygen firing. The recycle and process gas will be taken from the outlet of the ID fans by the new fans. No changes are anticipated to the ID fans since their duty does not change in principle. However, the ID fan performance and the impact of the gas constituents (density) must be evaluated based on specific site conditions.

3.4.3. Conclusions of subtask 4.1

Retrofitting oxygen combustion with gas recycle on existing plants is a viable option. New equipment for removing moisture from the flue gas and reheating the gas after the cooler will be needed. Depending on the fuel being burned, a wet or dry scrubber may be needed to reduce the SO₂ concentration in the recycle gas. This would be necessary for protection against coal ash corrosion. It is also recommended that the unit be equipped with new burners for optimization of the oxygen combustion. All other existing equipment could possibly be used. This would need to be evaluated on a case by case basis since it would be very site specific.

3.5. Subtask 4.2 Preliminary Design of a New Generation Boiler

3.5.1. Objective and deliverable of subtask 4.2

Objective

The objective of this subtask was to determine the optimum configuration of a new oxygen-fired boiler with flue gas recycle. As opposed to the retrofit case, where the boiler has originally been designed for air-fired conditions, the new boiler does not have the same constraint in terms of steam cycle and boiler operating conditions. The new boiler will still feature some recycled flue gas to control the temperature and potentially improve the boiler efficiency, however likely in much lower volumes than in retrofit case to get maximum benefit from the smaller boiler size. The minimum recycled flue gas required to reduce excessive flame temperature and to be compatible with today's boiler tube materials and targeted steam production was assessed. Also, the optimum operating conditions needed to reduce or avoid air infiltrations were evaluated. The boiler conceptual design and some information on the related costs are provided.

Deliverable

The deliverable of this subtask is a conceptual design of new boilers taking maximum advantage of the oxygen-enriched flue gas firing mode using existing boiler tube materials. Along with the conceptual boiler design, heat and mass balance data and an insight into cost savings as compared to a conventional air-fired design are provided.

Optimizing Oxy-fuel Boiler by Flue Gas Recycle Reduction

Oxygen fired boilers are gaining interest as a means of addressing carbon emissions for firing coal to generate electricity. The advantages of using oxygen over air firing are the possibility of concentrating carbon dioxide for sequestration by means such as enhancement of oil and natural gas recovery and reduced emissions of NO_x. Earlier

studies have explored modifying a conventional air fired boiler to be capable of full load operation with air or oxygen by recirculating flue gas back to the boiler. The flue gas is recirculated to maintain the distribution of heat absorption throughout the furnace and convection pass the same as on air firing and maintain the gas mass flow necessary for heat transfer in the convection pass. In this manner, the boiler could be operated with up to full load capability on air or oxygen as required. Unlike previous work, this study investigates reduction of the quantity of flue gas recycle to as low as possible while still employing conventional technologies. The advantages envisioned are reduced boiler and equipment size and subsequent costs.

The initial concept of this boiler was to only fire with air for start up and oxygen fire thereafter, as opposed to the dual full load capability of oxygen and air in previous studies. Since the boiler will operate with oxygen, a cyclone combustion system was chosen as the starting point.

Cyclone furnaces have the advantage of good combustion and since most of the ash is tapped from the bottom of the furnace reducing slagging and fouling in the upper furnace, the furnace and convection pass enclosures can be smaller due to the decreased ash flow through the boiler. A smaller perimeter for the furnace also means more steam flow per foot of perimeter in the furnace walls. This will have the advantage of absorbing more heat from the combustion gases without overheating the wall tubes. Since reducing the gas recycle rate is expected to increase the adiabatic flame temperature, this will give more margin to furnace absorption before tube walls begin reach their temperature limits. Cyclone furnaces have been used since the 1940's but have fallen out of favor due to the high thermal NO_x emissions generated by the high combustion temperatures and nitrogen in air. Since this boiler will not be fired with air except for the brief period during startup, it seemed the logical choice for this study.

The steam turbine cycle used is for a 625 MWe plant and employs supercritical steam at advanced temperature conditions. This cycle was chosen due to its high efficiency. Table 8: Steam Conditions shows the critical steam side data. These steam conditions were held constant throughout the study.

Table 8: Steam Conditions

Parameter	Units	Value
Main Steam Flow	klbs/hr	3758.2
Reheat Steam Flow	klbs/hr	3037.3
Main Steam Temperature	F	1100
Reheat Steam Temperature	F	1100
Feedwater Temperature	F	560
Main Steam Pressure	Psig	3512

The boiler was treated as two separate sections, the furnace and the convection pass. The furnace was studied from a standpoint of required furnace height as a function of Furnace Exit Gas Temperature (FEGT). No consideration was given to the circulation system since it was beyond the scope of this study. The boiler convection pass performance was analyzed to meet steam conditions at the reduced gas flows to determine how much the

flue gas recycle could be reduced while still meeting performance using conventional materials and technologies.

3.5.2. Furnace

A special program for calculating furnace absorptions using different gas compositions was used for estimating FEGT at different recycle rates. Table 9 lists the different recycle rates along with the gas temperatures at the furnace exit and boiler fuel efficiencies. As a reference, the boiler fuel efficiency for the air-fired case was 84.7%. Fuel efficiency is calculated per the following equation:

Equation 1

$$eff = \left\{ 1 - \frac{(L - B)}{H_f} \right\} \times 100$$

Where;

eff = Boiler Fuel Efficiency

L = Heat Losses

B = Heat Credits

H_f = Fuel Higher Heating Value

The reason for the reduction in efficiency for the oxygen-fired cases is the inability to cool the gas exiting the air heater sufficiently to maintain the dry gas loss the same or less than air firing. It is noted that the term “air heater” refers to the boiler exit gas to recycled gas heat exchanger, and is used throughout since the function of the heat exchanger is the same as an air heater on an air fired unit. The high water content places a large heat load on any cooling equipment in the back end to lower the temperature below about 135°F. The original air/O₂ fired furnace size was held constant for these cases. Tables 10 & 11 provide the gas compositions for each case. These compositions were based on the oxidant being 95% oxygen with the remaining gas as nitrogen. The term full recycle refers to the amount of recycled gas needed to match air fired performance with the same heating surface arrangement.

Table 9: Effect of Recycled Flue Gas on Boiler Conditions

Case	Recycled Flow	Gas Flow Leaving Furnace	FEGT	Boiler Fuel Efficiency	Fuel Flow klbs/hr
Full Recycle	4540.5	6325.3	1860	82.13	818.5
20% Reduction	3655.0	5421.3	1899	83.02	809.8
25% Reduction	3405.4	5166.2	1905	83.53	807.3
44% Reduction	2554.0	4296.9	1898	84.32	798.9

Table 10: Composition of Flue Gas

Flue gas Component % Vol Wet	Full Recycle	20% FGR Reduction	25% FGR Reduction	44% FGR Reduction
% CO ₂	64.69	63.53	63.14	61.48
% H ₂ O	25.04	26.41	26.88	28.86
% SO ₂	0.071	0.08	0.085	0.10
% O ₂	3.47	3.39	3.36	3.24
% N ₂	6.73	6.59	6.54	6.32

Table 11: Composition of Flue Gas – On a Dry Basis

Flue gas Component % Vol Dry	Full Recycle	20% FGR Reduction	25% FGR Reduction	44% FGR Reduction
% CO ₂	86.29	86.33	86.35	86.43
% SO ₂	0.095	0.11	0.12	0.14
% O ₂	4.63	4.61	4.59	4.55
% N ₂	8.98	8.96	8.94	8.88

For comparison, it should be noted that for the same furnace design with air firing, the FEGT is approximately 2100°F. The lower FEGT with oxygen firing and gas recycle is due to the increased carbon dioxide and moisture content of the flue gas, which increases the flue gas emissivity and furnace absorption. The higher carbon dioxide content also increases the flue gas density, which results in a longer gas residence time in the furnace (lower velocity). Note that the FEGT begins to rise with reduced gas recycle until the 44% reduction case. The increased temperature is due to the increasing adiabatic flame temperatures in the burner zone. However, at the 44% FGR reduction case, the increased residence time in the furnace and the temperature difference between the furnace walls and gas begins to transfer more heat to the walls thus reducing furnace exit gas temperatures. This raises a potential problem that is beyond the scope of this study. The increased heat flux to the furnace walls has the potential to heat the furnace tubes beyond their allowable limits, which may require advances in metallurgy for wall tubes to permit these conditions.

For this study, a reduction in furnace height was needed to increase the FEGT to achieve acceptable heat transfer to the heating surface in the convection pass. A temperature of 2000°F was chosen for the target FEGT. This temperature lies between the air firing case and the gas recycle cases and allows for some error in FEGT prediction. The furnace height for the 25% FGR reduction case needed to be reduced by 15'-6" to meet this target

temperature. Note that the furnace could have been reduced further for the full recycle case since its predicted FEGT is 40°F lower. Further reductions in furnace height appear to be possible at reduced recycle rates as indicated by the trend for FEGT to decrease as the recycle rate is reduced. Further studies would be needed to determine if this trend would continue and to what extent. Further pilot data also needs to be generated while firing with this method to determine the residence time needed for full carbon burn out in the furnace. Standards today are largely empirical and based on firing coal with air. Due to the anticipated higher adiabatic flame temperatures with oxygen firing, the residence times required may be less.

One final note is that reduced gas recycle does have the advantage of increasing boiler efficiency. This is largely due to the reduced amount of heat being lost through the exit of the air heater, which is the final heat trap for the flue gas. The reduced heat loss is the result of the reduced gas flow. There may also be an increase in efficiency due the reduction in boiler surface area, which would reduce the radiation loss slightly. This effect has not been accounted for in this study

3.5.3. Boiler convection pass

After the furnace analysis, the convection pass was modeled to determine if the desired performance could be obtained with the reduced gas flows and to try to determine the minimum amount of gas recycle. Two criteria were established for this portion of the study.

- 1) The FEGT would be kept at 2000°F.
- 2) Conventional technology would be used to the furthest extent possible.

Figure 10 shows a typical example of a boiler with the convection pass used for conventional air firing. The first step was to model the boiler for air firing to achieve the previously given steam data. The boiler was 50 feet wide by 35 feet deep. This boiler model was then modified to simulate oxygen firing with gas recycle.

Table 12 illustrates some basic data from these calculations.

Table 12: Comparison of Boiler Heat Transfer for Baseline Oxy-combustion

	Air Fired	Oxygen Fired - Expected Recycle	Oxygen Fired – Reduced Recycle
Gas Flow leaving the Furnace (klb/hr)	5421.7	6325.3	5421.6
Total Heating Surface Area Circumferential (ft ²)	590,567	590,567	590,567
Radiant Surface Peripheral (ft ²)	13,040	13,040	13,040
% Gas Flow in RH Path	27	31	51
% Gas Flow in Pri Path	73	69	49
Maximum Gas Velocity (ft/sec)	45.2	39.9	49.6
Gas Velocity Entering the Pendant Convection Pass (ft/sec)	31	26	23
Gas Temperature Leaving Boiler (°F)	700	691	696
Gas Temperature Leaving the Air Heater (°F)	275	272	274
Boiler Width (ft)	50	50	50
Boiler Depth (ft)	35	35	35

The data in Table 12 illustrates boiler performance for air firing, expected gas recycle with oxygen firing, and with sufficient gas recycle to achieve the same gas mass flow leaving the furnace as with air firing. The boiler geometry was held constant in all three cases. These three cases were the basis of comparison for this study. They illustrate the boiler performance if the boiler had the dual capability of air and oxygen firing and the effects of gas recycle with oxygen firing. The following provides further explanation of the data in Table 12.

Gas Flow leaving the Furnace (klb/hr) – This is the flue gas leaving the furnace exit and entering the convection pass. It includes the recirculated flue gas, combustion gases, and oxidant. In this case, referring to Figure 1, the furnace exit is at the leading edge of the vertical bank of tubes directly above the arch.

Total Heating Surface Area Circumferential (ft²) – This is the bank heating surface from the furnace exit to the boiler exit (airheater inlet). It does not include any enclosure surface.

Radiant Surface Peripheral (ft²) - This is the radiant heating surface hanging directly above the furnace before the furnace exit. It can typically be platen surface as shown in Figure 10 or can be in the form of curtain walls or wing walls.

% Gas Flow in RH Path and % Gas Flow in Pri Path – In a parallel backend boiler such as this one, the heating surface in the down flow pass of the boiler (horizontal surface) is separated into two gas paths. The reheat temperature is controlled by control dampers at the exit of the gas paths that regulate the amount of gas that flows over the horizontal reheater. The reheat surface is on the left side in Figure 1 and primary superheater surface is in the right. Typically, the percentage of gas mass flow over the reheat pass should be around 20 to 30% of the total gas mass flow at full load. This allows sufficient control that gas weight may be increased at lower loads to maintain reheat temperature.

Maximum Gas Velocity (ft/sec) – This is the maximum gas velocity in the boiler and it can occur anywhere in the boiler depending on the conditions.

Gas Velocity Entering the Pendant Convection Pass (ft/sec) – The gas velocity entering the first bank of tubes after the furnace exit.

The remaining items are self-explanatory. Notice in the column for reduced recycle that the % gas flow in the reheat pass exceeded 50%. This indicates that reducing gas recycle any more than shown would require some surface modification since the boiler is already incapable of much additional load reduction.

At this point, a change in surface was needed before further reductions in gas recycle were possible. It was decided that additional reheat surface was needed. Adding more horizontal banks near the exit of the boiler was insufficient because the temperature of the flue gas was approaching the reheat inlet steam temperature. The solution was to replace the pendant secondary superheater outlet with an additional pendant bank of reheat surface. This, of course, lowered the amount of main steam superheater surface available. To compensate for this loss, additional superheater surface was added in the furnace in order to take advantage of the increased radiant heat from the hotter flame temperature. It was assumed that this additional surface would be in the form of curtain walls in the upper furnace. No attempt was made to arrange this surface physically but curtain walls have been used and the concept is not beyond current technology. In addition, the boiler roof was lowered 10 ft to increase gas velocity through the pendant surface to maintain a reasonable heat transfer coefficient. However, even though the velocity could have been increased more, reducing the height of the banks of surface any further would begin to have detrimental effects on the amount of heat transferred due to the reduction in heating surface. Table 13 below shows the results of this configuration.

Table 13: Effect of Flue Gas Recycle Flow Rate on Boiler Performance

	Base Recycle	25% Reduced Recycle	44% Reduced Recycle
Gas Flow leaving the Furnace klb/hr	6325.3	5166.2	4296.9
Total Heating Surface Area Circumferential (ft ²)	447,870	426,591	447,956
Radiant Heating Surface Peripheral (ft ²)	13,040	21,040	25,040
% Gas Flow in Path 1	27	13	38
% Gas Flow in Path 2	73	87	62
Maximum Gas Velocity ft/sec	40.5	53.4	43.4
Gas Velocity Entering the Pendant Convection Pass ft/sec	28	33	28
Gas Temperature Leaving Boiler °F	730	711	717
Gas Temperature Leaving the Airheater °F	364	354	356

The Base Gas Recycle case is essentially the same as the Expected Recycle case shown in Table 13 with the difference being a reduction in heating surface for the Base Recycle case. This was done to obtain a fair comparison of convective heat transfer. The convection pass performance is measured from the roof inlet tubes to the main steam outlet. The heating surface was arranged for the roof inlet enthalpy to be equal in all three cases. The higher surface area shown in Table 12 also reflects the dual firing modes of air and oxygen.

Notice the increasing amount of peripheral surface required in both Reduced Recycle cases in Table 13. This represents the additional radiant furnace surface needed for the main steam heat transfer performance. In the 44% reduction case, this radiant surface approached double that required for the base case. Also notice that the % gas flow in the horizontal reheat pass (Path 1) is so high that increasing gas weight at reduced loads and achieving boiler turn down will begin to have difficulties. It is this reduced gas flow, which marks the limit of this surface arrangement. To reduce this flow would require additional pendant or radiant reheat surface plus additional main steam radiant surface. The problems associated with this will be explained in the next section.

3.5.4. Further studies for advancing in technology

Further pilot scale test programs and engineering analyses will be required to further reduce the amount of recirculated gas below a 44% reduction. First, predictions of FEGT and local furnace absorption rates should be confirmed by pilot testing or demonstration units. This information will not only confirm or correct the FEGT calculations but will also serve to understand furnace heat fluxes and problems associated with increasing adiabatic flame temperature better. Second, analysis of furnace circulation and wall temperatures will be necessary to understand the metallurgy that will be required for furnace construction. This need not only exists or further gas recycle reductions but also to apply the results of this study.

The surface arrangement used in this study also presents some challenges due to the additional pendant reheat bank. Reheaters are limited in the amount of steam pressure drop through the boiler because of steam turbine cycles. Any additional reheat pressure drop other than that set by the turbine will result in a lower plant heat rate. Although it is believed that the arrangement used in this study will not adversely affect the reheater pressure drop, any additional surface would only increase the required pressure drop through the boiler. To confirm that the additional pressure drop is not a significant concern, a complete pressure drop analysis should be performed.

Since reheaters are pressure drop limited, they are designed with lower steam mass flows than the main steam heating surface. This lower mass flow increases the tube temperatures because of the reduced cooling accomplished by the steam. This is particularly troublesome in higher gas temperatures. Tube metal temperatures in the reheat outlet bank were calculated in this study and some tubes exceeded fireside surface metal temperatures of 1200°F. Although this is not beyond current metallurgy, it does make an expensive reheater. Any additional reheat surface in even hotter gas streams will increase these temperatures even further.

3.5.5. Costs

Budgetary pricings were developed for the boiler and the price reductions associated with the height reductions in the furnace and convection pass. Figure 11 is a chart showing the percentage of cost of individual components for the boiler island with the boiler itself being 64% of the total.

Figure 12: Percentage of Cost of Individual Components for The Boiler Island

The base price for the boiler for air firing is estimated at \$164.4 (\$M). By reducing the furnace height by 15.5' to increase FEGT, a savings of \$213,000 is expected. The decrease of 10' in the pendant convection pass is further expected to realize a savings of \$279,000. The savings due to a reduction of bank heating surface in the pendant convection caused by the height reduction are considered to be about the same as the addition of radiant surface in the furnace. This brings the boiler price down to \$163.9 (\$M) which in itself is not a significant savings. However, the savings in flues, fabric filter, scrubber and recirculating fan are expected to be significant because the gas flow treated is 68% of that required for the base case. This is a 32% reduction of the total gas flow compared to the base case and the reduction in costs to the flues and environmental equipment were assumed to be directly proportional. In addition, since the boiler produces very little NO_x, the SCR can be assumed to be eliminated. These reductions add up to approximately a 15% cost reduction in the total boiler island costs. In addition, the reduced boiler and backend equipment would have a favorable impact on building steel and foundation costs.

3.5.6. Conclusion of subtask 4.2

Lowering the amount of recirculated flue gas when firing with oxygen appears to be feasible. In addition, a 44% reduction seems practical using current technology. There are however, additional issues that must be resolved that require further pilot scale testing and engineering studies prior to demonstration scale testing to confirmation this initial assessment. Reduction in the amount of flue gas recycle for an oxygen-fired boiler designed to concentrate carbon dioxide for sequestration offers the considerable potential for reductions in material, construction and operating costs.

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6. LIST OF ACRONYMS AND ABBREVIATIONS

AAL	American Air Liquide
ACI	Activated Carbon Injection
BSR	Burner Stoichiometric Ratio
B&W	The Babcock and Wilcox Company
Cap	Capture Power Plant
CHX	Condensing Heat Exchanger
CPEGT	Convective Pass Exit Gas Temperature
COE	Cost of Electricity
CPO	Convective Pass Outlet
DOE	Department of Energy
EOR	Enhanced Oil Recovery
EPA	Environmental Energy Agency
EPRI	Electric Power Research Institute
ESP	Electro-Static Precipitator
FD Fan	Forced Draft Fan
FEGT	Furnace Exit Gas Temperature
FG	Flue Gas
FGD	Flue Gas Desulfurization
FGR / RFG	Flue Gas Recycle / Recycled flue gas
FGT	Flame Gas Temperature
FOM	Fixed Operating & Maintenance costs
GHG	Green House Gases
Hg	Mercury
HMI	Human Machine Interface
HRSG	Heat Recovery Steam Generator
ID Fan	Induced Draft Fan
IEA	International Energy Agency
IGCC	Integrated Gasification Combined Cycle
IGCC-S	IGCC equipped with a Selexol unit
ISGS	Illinois State Geological Survey
LOI	Lost On Ignition (same as UBC)
LP	Low Pressure
LSD	Lime Spray Dryer
MEA	Mono-Ethanol-Amine
10 ⁶ Btu/h	Million Btu/hour
Mills	1,000 mills = \$1
mt or metric t	metric ton
NETL	National Energy Technology Laboratory

OC	Oxy-Combustion
OHM	Ontario Hydro Method (mercury measurement)
O&M	Operating and Maintenance
PA	Primary Air
PACI	Pulverized Activated Carbon Injection
PC	Pulverized Coal
PC-MEA	PC Boiler equipped with MEA unit
PC-OC	Oxy-Combustion in PC boilers
PC _{ref}	Reference air-blown PC plant without CO ₂ capture
PM	Particulate Matter
PRB	Powder River Basin
PO	Primary Oxidant
Ref	Reference Power Plant
t or ton	short ton
SA	Secondary Air
SBS	Small Boiler Simulator
SCR	Selective Catalytic Reduction
SI	Système International d'Unité
SNCR	Selective Non Catalytic Reduction
Syn-gas	Synthesis gas (H ₂ /CO)
SO	Secondary Oxidant
TA	Tertiary Air
TBD	To be defined
TO	Tertiary Oxidant
TPD	(metric) Ton per Day
UBC	Unburned Carbon in Ash (same as LOI)
VOM	Variable Operating & Maintenance costs
XRF	X-ray Fluorescence

UNITS

dscm	Dry Standard Cubic Meter
µm	Micronmeter (micron) = 10 ⁻⁶ m

7. APPENDICES

Appendix A.

SBS (Small Boiler Simulator) in an Oxygen Firing Mode.

Appendix B.

New equipment and duct work.

Appendix C.

Modified Oxygen Skid.

Appendix A. SBS (SMALL BOILER SIMULATOR) IN AN OXYGEN FIRING MODE

Figure A-1: SBS (Small Boiler Simulator) Schematics

Appendix B. NEW EQUIPMENT AND DUCT WORK (IN RED) INSTALLED ON THE SBS

Figure B-1: New Equipment and Duct Work Installed on the SBS

Appendix C. MODIFIED OXYGEN SKID

Skid area: simplified legs.

Components are the same as on
hardcopy P&ID rev 06-04-2002

O2COM11PD0200

Blue: new/modified

Figure C-1: Updated Modified Oxygen Skid

Advanced, Low/Zero Emission Boiler Design and Operation

TOPICAL REPORT – TASK 3

Performance Simulation and Economics Assessment of three CO₂ capture coal-fired power plants:

- Sub-critical pulverized coal power plants equipped with Oxy-combustion (New and Retrofit Plants)
- Sub-critical pulverized coal power plants equipped with Post-combustion (New and Retrofit Plants)
- IGCC Power plant equipped with Selexol scrubbing (New Plant)

Work Performed Under Contract No.: **DE-FC26-02NT41586**

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ABSTRACT

This report describes the performance simulation and cost assessment of processes designed to capture and compress CO₂ from sub-critical pulverized coal fired power plants (PC) and Integrated Gasification Combined Cycle (IGCC) units. A plant capacity of 533 MW_e gross power output firing western PRB coal was considered. The CO₂ was compressed to 83 bara (1,204 psia). Oxy-Combustion (PC-OC) and amine scrubbing technologies (PC-MEA) are considered as technology options to capture CO₂ from PC plants, and Selexol process (IGCC-S) from IGCC plant. Both new plant and retrofit applications are investigated. Results of the mass and energy balance were obtained from steady state simulations. Based on recent studies from DOE, EPA and EPRI, cost models were developed to estimate the capital, operating, electricity and the CO₂ avoidance costs for each technology.

The performances and economics of the three CO₂ capture processes are compared to a reference sub-critical PC plant of same gross power output (533MW_e) without CO₂ capture, referred to as PC_{ref}. The net power output and net generation efficiency of the PC_{ref} plant are about 500MW_e and 37%, respectively.

Both the PC-OC process and the PC-MEA process are technically applicable for plant retrofit to capture CO₂ emissions. In retrofit applications of sub-critical PC plants, the net power output decreased to 358 MW_e for the retrofit PC-OC plant, and to 350MW_e for the PC-MEA plant. The energy penalty resulted in the net generation efficiency decrease to 27.3% for the PC-OC, and to 25.8% for the PC-MEA units. The required capital cost per unit of net power output of the retrofitted unit is \$595/kW_e and 665/kW_e for the PC-OC and for the PC-MEA, respectively. The increase in cost of electricity due to the retrofit is about 4.1 cents/kWh using PC-OC, and 4.9 cents/kWh using PC-MEA. The cost of CO₂ avoidance is \$48/metric ton (mt) for the retrofit PC-OC, compared to \$64/mt for the retrofit PC-MEA. The PC-OC process is economically more favorable than the PC-MEA process for CO₂ capture from the existing PC power plants.

For new plants applications, the net power output was about 363MW_e for the new sub-critical PC-OC plant and 350MW_e for the PC-MEA plant, with corresponding net generation efficiencies of 27.9% and 25.8%. An IGCC-S process of same gross capacity as the PC units would end-up with 435MW_e net power output and 36.2% net generation efficiency. The COE of the CO₂ capture plants increases from 5.0 cents/kWh in PC_{ref} to 8.1 cents/kWh for new sub-critical PC-OC plant, 9.0 cents/kWh for new sub-critical PC-MEA plant, and 7.2 cents/kWh for IGCC-S plants. The CO₂ avoidance cost is about \$36/mt for the PC-OC plant, \$52/mt for the PC-MEA plant, and \$27/mt for the IGCC-S plant.

Recommendations are also provided to further improve the economics of the PC-OC technology, including using super or ultra-supercritical steam cycles rather than sub-critical when applied to new power-plants.

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EXECUTIVE SUMMARY

A techno-economic study was performed to evaluate the performance and cost of three coal-fired power generation systems with CO₂ capture: pulverized coal (PC) boiler fired in oxygen-enriched flue gases (oxy-combustion in PC boilers, referred to as PC-OC), air-fired PC boiler equipped with amine scrubbing (PC-MEA) and IGCC unit equipped with selexol scrubbing (IGCC-S). A conventional PC power plant without CO₂ capture was also investigated as a baseline for comparison (PC_{ref}). The coal used in this study is the western sub-bituminous PRB coal. This report describes the results obtained on 533 MW_e gross output capacity for all the plant configurations. The CO₂ product is compressed to supercritical state at 83bara (1,204psia). In the PC-OC process, neither air infiltration nor CO₂ purification system other than flue gas drying is considered.

The PC unit (PC_{ref}, PC-OC and PC-MEA) features a sub-critical steam cycle. The primary purpose of the study was indeed the investigation of CO₂ capture technologies applicable to existing PC boilers (retrofit), most of them in the US being sub-critical. Many new PC units constructed today feature higher efficiency supercritical steam cycle. Economics assessment of CO₂ capture technologies applied to supercritical PC plants are not included in this study.

A process simulation software, Chemcad, was employed to perform steady state simulations of power generation systems and CO₂ capture processes. Based on several recent studies funded by DOE, EPA and EPRI, cost models were developed to estimate the capital, O&M, electricity and CO₂ avoidance costs for each technology.

The following criteria have been used to compare the three CO₂ capture technologies investigated in this effort (PC-OC, PC-MEA, IGCC-S): Loss of net power output versus the same gross power output baseline PC plant (PC_{ref}), net plant efficiency (%), cost of electricity (COE, mills/kWh), Cost of CO₂ avoided (\$/metric ton), Cost of CO₂ captured (\$/mt), Capital and O&M costs (\$/kW_e net and mills/kWh respectively).

Performance results

Based on the process simulation performed, the net power output of the reference sub-critical PC_{ref} without CO₂ capture plant was about 500MW_e for both existing and new PC plants. PC_{ref} net generation efficiency was about 37%.

In retrofit applications, the power output decreased to 358MW_e for the retrofit PC-OC plant, and to 350MW_e for the PC-MEA plant. The net generation efficiency decreased to about 27.3% in the retrofit PC-OC, and 25.8% in the PC-MEA units. In the PC-OC plant, out of 28% net power reduction, about 20% is due to the ASU power consumption and the remaining 8% due to the CO₂ compression. In the PC-MEA plant, out of 30% net power reduction, about 23% is due to the amine regeneration process, required fans and pumps, and 7% due to the CO₂ compression.

For new plants applications, the net power output was about 363MW_e for the new sub-critical PC-OC plant with 27.9% net efficiency, 350MW_e for the PC-MEA plant with 25.8% net efficiency. An IGCC-S process of the same gross capacity as the PC units would end-up with 435MW_e net power output with 36.2% net efficiency.

Performance summary of retrofit plants

	Existing-PC_{ref}	Retrofit PC-OC	Retrofit PC-MEA
Gross Power Output	533.2 MW _e	533.2 MW _e	533.2 MW _e
MEA regeneration (steam)	-	-	98.4 MW _e
ASU Power consumption	-	101.5 MW _e	-
Auxiliary Power consumption other than ASU and MEA	31.8 MW _e	73.9 MW _e	85.1 MW _e
Net power Output	501.4 MW_e	357.9 MW_e	349.8 MW_e
Plant net efficiency	37 %	27.3 %	25.8 %
CO ₂ capture efficiency	0 %	96 %	90 %

Performance summary of new plants

	New-PC_{ref}	New PC-OC	New PC-MEA	IGCC-S
Gross Power Output	533 MW _e	533 MW _e	533 MW _e	536 MW _e
MEA regeneration (steam)	-	-	98.4 MW _e	-
ASU Power consumption	-	100.6 MW _e	-	36.8 MW _e
Auxiliary Power consumption other than ASU or MEA	31.8 MW _e	70.1 MW _e	85.1 MW _e	63.5 MW _e
Net power Output	501.4 MW_e	362.6 MW_e	349.8 MW_e	435.5 MW_e
Plant net efficiency	37 %	27.9 %	25.8 %	36.2 %
CO ₂ capture efficiency	0 %	96 %	90 %	90 %

Economics

The economic analysis showed that:

- **For retrofit Power Plants:** Both the PC-OC process and the PC-MEA process are technically applicable for PC plant retrofit to capture CO₂ emissions. The retrofit capital cost per unit net generation for the PC-OC process is about 11% lower than that of the PC-MEA retrofit, while the increase of O&M cost due to retrofit is about 35% lower. The cost of CO₂ avoidance is about \$48/mt for the PC-OC retrofit compared to \$64/mt for the MEA retrofit. The PC-OC process is economically favorable for CO₂ capture from existing PC power plants.

Cost of CO₂ capture technologies applicable to Existing PC power plants

	CO₂ capture plants	
	Retrofit PC-OC	Retrofit PC-MEA
Net Power Output	358 MW _e	350 MW _e
Total Plant Cost (TPC), \$/kW_e	595	665
Levelized capital costs, mills/kWh	14.58	16.97
Levelized O&M, mills/kWh	6.41	9.89
Cost of electricity loss, mills/kWh	20.16	21.80
Increase in COE, mills/kWh	41.15	48.66
Cost of CO₂ avoidance, \$/mt CO₂	48.44	63.51
Cost of CO ₂ capture, \$/mt CO ₂	35.20	42.21

- **For new power Plants:** The COE of the sub-critical PC-OC plant increases by about 60% compared to PC_{ref} (no CO₂ capture). The COEs for the PC-MEA plant and the IGCC-S plant were increased by 79% and 43%, respectively. The CO₂ avoidance cost is \$36/mt for the PC-OC plant, \$52/mt for the PC-MEA plant, and \$26/mt for the IGCC-S plant.

Cost of CO₂ capture technologies applicable to New coal-fired Power Plants

	No CO ₂ capture	CO ₂ capture plants		
	New-PC _{ref}	New PC-OC	New PC-MEA	IGCC-S
Net Power Output	501 MW _e	363 MW _e	350 MW _e	435 MW _e
Total Plant Cost (TPC), \$/kW_e	1,115	1,861	2,000	1,841
Total Plant Investment (TPI) , \$/kW _e	1,216	2,028	2,181	2,097
Total capital Requirement (TCR) , \$/kW_e	1,276	2,123	2,289	2,187
Levelized capital costs, mills/kW _h	28.72	47.78	51.52	49.21
Levelized O&M, mills/kW _h	21.53	33.04	38.47	22.57
Levelized COE, mills/kW_h	50.24	80.82	89.99	71.78
Increase in COE, mills/kW _h	-	30.58	39.75	21.54
Cost of CO₂ avoidance, \$/mt CO₂	-	35.95	51.88	26.51
Cost of CO ₂ capture, \$/mt CO ₂	-	26.73	34.45	24.22

Sensitivity Analysis

Both the PC-OC and PC-MEA processes require a large amount of energy for capturing CO₂. The sensitivity analysis showed that the CO₂ avoidance cost is more sensitive to the absorbent regeneration heat in the PC-MEA process and the electricity use of the ASU in the PC-OC process than their respective capital cost requirements. The CO₂ avoidance cost for the PC-OC process decreases by about 33% when the electricity use of the ASU is reduced by 50%. The CO₂ avoidance cost of the MEA process decreases by about 29% when the absorbent regeneration heat is reduced by 50%.

Recommendations for future work

For new power plant applications, the economics and net generation efficiency of both the PC-OC and PC-MEA plants can be significantly improved by changing the steam cycle from sub-critical to super or ultra super-critical cycle. Increasing the power plant efficiency from 37% to the 40+ %, will result in smaller ASU requirement in the PC-OC plant, in lower heat requirement in the PC-MEA plant, and in reduced CO₂ volume in both processes. Integrating the ASU with the power plant can further reduce its power consumption and improve the economics of the process. Similarly, development of new absorbents with lower heat requirements for solvent regeneration can improve the economics of the PC-MEA process.

1. INTRODUCTION

1.1. Context

As large amounts of anthropogenic greenhouse gases are being emitted to the atmosphere, global warming has become one of the most important environmental issues today. Fossil fuel combustion is the major contributor of increased Greenhouse Gas (GHG) emissions. About one third of carbon emissions in the United States comes from power plants, one third from transportation and the remaining one third from industrial, commercial and residential sources. As fossil fuels continue to be the dominant fuel source and electricity generation is expected to grow, reducing carbon emissions by economically capturing and sequestering CO₂ from energy industries is imperative.

GHG emissions can be controlled by increasing energy efficiency, switching to less carbon-intensive energy, carbon sequestration etc. The first two options are not believed to be sufficient enough to achieve the GHG reduction targets. Inexpensive large-scale capture and sequestration methods are necessary to make such goals achievable. IEA research also suggested that carbon sequestration could play an important role in the deep reduction of CO₂ emissions in the first part of the 21st century ^[1].

The US generates about 25% of the worldwide anthropogenic CO₂ emissions ^[2]. The coal-fired power plants are the most significant contributors among various CO₂ emission sources. Currently, coal-fueled utilities consume more than 90% of the coal used in the US ^[3], and are responsible for about 73% of US CO₂ emissions ^[4]. There is clearly a compelling need to develop new and retrofit technologies to capture and sequester the CO₂ emissions from coal fired power plants.

1.2. CO₂ capture technologies for fossil fuel-fired combustion systems

CO₂ separation/capture from flue gases of various fossil fuel combustion systems can be described into three main categories: post-combustion, pre-combustion configurations and oxygen combustion (Figure 1) ^[5].

Figure 1. CO₂ capture via Oxy-combustion, Pre-combustion or Post-combustion¹

¹ Source: Carbon Capture Project (CCP)

1.2.1. Oxy-combustion (OC)

In OC process, oxygen is separated from air and then burned with hydrocarbon fuels to produce an exhaust with a high concentration of CO₂. To control the combustion temperature, oxygen may be diluted in recycled flue gases. The resulting benefits of the OC process include reduced flue gas volume for treatment, concentrated CO₂ in the flue gas convenient for further CO₂ treatment, reduced NO_x emissions, and applicability both for a new plant and retrofit ^[9].

1.2.2. Post-combustion

In the post-combustion configuration CO₂ is captured from the flue gas after combustion. This technology can thus be retrofitted to existing equipment. Different physical and chemical processes are being developed for this purpose, including absorption, adsorption, membranes and cryogenic separation ^[6]. Absorption is a promising approach for CO₂ capture on a power plant scale and at a CO₂ concentration of about 15-20% of the flue gas. The mono-ethanol-amine (MEA) absorption process was initially developed for separating CO₂ from flue gas streams mainly used for enhanced oil recovery in the 1970s, and is currently being widely studied for capturing the CO₂ from power plants ^[7].

1.2.3. Pre-combustion

In the pre-combustion separation, the carbon-containing fuel is transformed to a non carbon-containing fuel (hydrogen). Carbon in the fuel is converted to CO₂ prior to combustion of the fuel gas in the gas turbine combustor. The pre-combustion process, without CO₂ capture, is already used in several commercial-scale Integrated Gas Combined Cycle (IGCC) plants fueled by coal, residual oil or petroleum coke ^[8]. One advantage of an IGCC plant equipped with a CO₂ separation process is that increased CO₂ concentration and pressure in the gas stream significantly reduce the size of CO₂ capture equipment, so that a weaker physical absorbent, such as Selexol, could be used.

1.3. Objectives of the study

The objective of this study was to perform a techno-economic assessment of the oxy-combustion process in pulverized coal boilers (PC-OC), and compare it with two other CO₂ capture technologies: the MEA process in pulverized coal boilers (PC-MEA) and the Selexol process in IGCC plants (IGCC-S). A conventional air-blown pulverized coal (PC) power plant without CO₂ capture was selected as a baseline for cost analysis of CO₂ capture. Several case studies covering a practical range of plant sizes have been performed for each technology (Table 1).

Table 1. Case studies of techno-economic analysis

	CO ₂ capture installed?	Abbreviation	Plant Capacity, Gross
Air-blown PC	No (reference plant)	PC	533 MW _e
Air-blown PC with MEA process	Yes	PC-MEA	533 MW _e
PC Plant with oxy-Combustion process	Yes	PC-OC	533 MW _e
IGCC + Selexol	Yes	IGCC-S	533 MW _e

The Powder River Basin (PRB) coal was used in all cases. Since coal compositions varies from mine to mine, or even seam to seam, a typical PRB coal was selected for this study. The PRB coal used was the Antelope Mine coal. Its ultimate analysis data is listed in Table 2 ^[10].

Table 2. Ultimate analysis of the western PRB coal

Composition, Wt%	As-received basis	Dry basis
Moisture	26.70	-
Carbon	51.35	70.05
Hydrogen	3.59	4.90
Nitrogen	0.78	1.06
Chlorine	0.01	0.01
Sulfur	0.24	0.33
Oxygen	12.08	16.48
Ash	5.25	7.16
Total	99.99	99.99
HHV(Btu/lb)	8,800	12,005

2. PROCESS DESCRIPTION

“**No-capture**” plant refers to a power plant that does not feature the capability of capturing the CO₂, which is hence emitted into the atmosphere. The “no-capture” plant used as reference in this study to compare to the capture plants is a typical air-fired pulverized coal plant, referred to as “PC_{ref}”.

“**Capture plant**” refers to a power plant equipped with a technology enabling CO₂ capture. Three capture plants are considered in this analysis:

- Pulverized Coal Power Plant equipped with amine scrubbing for CO₂ capture, referred to as **PC-MEA**
- Pulverized Coal Power Plant equipped with the Oxy-combustion Process for CO₂ capture, referred to as **PC-OC**
- IGCC Power Plant equipped with Selexol Unit for CO₂ capture, referred to as **IGCC-S**

The technical processes of the reference “no-capture” plant (PC) as well as the three different “capture plants” (PC-MEA, PC-OC, IGCC-S) are described in this chapter. A schematic illustrates each process. For further reference in the Process Simulation Results section of this report, some numbers are indicated on the schematics of CO₂ capture plants. ❶ refers to the flue gases exiting the convective pass section of the PC boilers, or to the syngas exiting the gasifier of the IGCC unit. In all processes, ❷ refers to the CO₂ product as available for transport then sequestration. Depending on the process, one or two intermediate streams composition are given (❸, or ❹ and ❺).

2.1. *Reference PC Power Plant without CO₂ capture (PC_{ref})*

Two different no-capture/reference PC plants are considered in this study (for new and retrofit applications) to compare the performances and cost of capture plants. Both reference plants are air-blown pulverized coal power plants, referred to as PC_{ref}. Both plants use a single reheat sub-critical steam power cycle. A balanced draft radiant boiler arranged with a water-cooled dry-bottom furnace is employed. The plants differ by the flue gas treatment technologies they are equipped with.

2.1.1. Reference PC for new power plant applications (New-PC_{ref})

For a brand new plant to be constructed, the flue gas cleaning system of the PC_{ref} plant includes the removal of dust, SO₂ and NO_x. For the low sulfur coal used in this study, a Lime Spray Dryer (LSD) process was chosen as flue gas desulphurization (FGD) process since it is more economical than the wet scrubbing process, with only slightly lower SO₂ removal efficiency. The post-combustion process, Selective Catalytic Reduction (SCR), was chosen for the NO_x removal. Such equipments reflect common practices for new pulverized coal-fired power plants burning low-sulfur coal in the US.

The schematic of the air-blown New-PC_{ref} power plant considered as reference for the new power plant applications is shown in Figure 2.

Figure 2. *New-PC_{ref} w/o CO₂ capture – Schematic Diagram*

2.1.2. Reference PC power plant for retrofit application (Existing-PC_{ref})

For an existing PC plant to be retrofitted to become a capture plant, the flue gas cleaning system of the PC_{ref} plant only includes the removal of dust by Electrostatic Precipitator (ESP). No SO₂ and NO_x removal system is included. To retrofit the plant, SO₂ and/or NO_x removal system will be part of the new equipment added for the retrofit, and were included in the calculation of the cost of CO₂ avoided. This assumption that only ESP exists on existing units does not reflect all the US Power plants but a majority of them. Estimate of cost impact would SO₂ and/or LSD be already included on the plant will be provided in the cost assessment chapter.

The schematic of the air-blown Existing-PC_{ref} power plant considered as reference plant for retrofit application is shown in Figure 3.

Figure 3. *Existing-PC_{ref} w/o CO₂ capture - Schematic diagram*

2.2. PC Plant equipped with Oxy-combustion for CO₂ capture (PC-OC)

The same sub-critical steam-power cycle and boiler have been employed for the PC-OC process. Experimental results obtained from pilot-scale tests confirmed that with part of the flue gas recycled, the combustion conditions in the PC-OC process can be similar to the air-blown combustion process, enabling use of today's PC boiler technology for PC-OC application^[28].

The PC-OC process evaluated in this study is illustrated in Figure 4.

Figure 4. PC-OC for CO₂ capture - Schematic diagram

In order to maintain the low concentrations of SO₂ and dust in the recycled flue gas, the flue gas is recycled downstream of the LSD and ESP. This might be necessary for corrosion prevention and/or safety reasons in the duct and the boiler. For the PRB coal, an LSD unit was considered for SO₂ removal in the PC-OC process. However, an SCR unit is not considered in the PC-OC, as the NO_x emissions could be reduced fourfold compared to the air-blown PC plant^[28]. The cleaned flue gas was cooled in a condenser before passing through the multi-stage compressors with the inter-stage coolers. The CO₂ is compressed to 83 bara (1,204 psia), to insure a supercritical state of the CO₂ product (CO₂ supercritical pressure $P_c = 73.8$ bars).

2.3. PC Plant equipped with amine scrubbing for CO₂ capture (PC-MEA)

As shown in Figure 5, the PC-MEA process corresponds to the addition of:

- A Mono-Ethanol-Amine (MEA) unit to scrub the CO₂ out of the flue gases, and a CO₂ compression train compared to the reference air-blown PC_{ref} plant for new power plant.
- A FGD, SCR, MEA units and CO₂ compression train compared to the reference air-blown PC_{ref} plant for retrofitted power plant.

Figure 5. PC-MEA for CO₂ capture - Schematic diagram

The CO₂ scrubbing process used in this study is a Fluor-Daniel Econamine FG Process. In this process, a large amount of heat is required for MEA solvent regeneration. A significant part of steam drawn from the steam cycle is therefore used as the heat source.

The CO₂ stream from the MEA process is further compressed to 83 bars, to obtain a supercritical product. Most of the moisture in the stream was condensed and removed during compression.

2.4. IGCC Plant equipped with Selexol Unit for CO₂ capture (IGCC-S)

The IGCC combines a gas turbine cycle and a steam turbine cycle into an integrated process. The IGCC plant with Selexol process is illustrated in Figure 6.

Figure 6. Destec IGCC-S with CO₂ Capture - Schematic diagram

The Destec gasifier is a two-stage, entrained flow and slagging design. The first-stage partial combustion zone uses 95% purity oxygen at 2,500°F which gasifies 80% of the total coal slurry. The remaining 20% of coal slurry is added to the second-stage where coal undergoes de-volatilization, pyrolysis and partial gasification in order to cool the raw gas and enhance its heating value. The product from the gasifier is mostly a mixture of CO+H₂ (synthesis gas, or syn-gas). The particulates in the fuel gas are then removed in a candle filter. After mixing with steam, the wet syn-gas enters into a shift converter where most of the CO is converted to CO₂. The resulting gas is cleaned of H₂S and CO₂. The cleaned syn-gas is then saturated, reheated and expanded to recover electrical energy. The gas is then routed to a gas turbine burning with air. Gas turbine exhaust heat is recovered in an HRSG to produce steam for steam turbines.

A double-stage Selexol unit is used to preferentially remove H₂S and CO₂. Selexol is an organic absorbent, which can be used to physically absorb CO₂ at high pressure and low temperature. In the first stage H₂S contained in syngas (705 psi, 105°F) is preferentially removed in the first absorber by loading the CO₂-loaded lean solvent. The rich solvent is then regenerated in a stripper. The resulting H₂S rich acid gas stream is sent to the Claus unit for S recovery. In the second stage syngas from the first absorber (90°F, 700 psi) contacted with CO₂-unloaded lean solvent and 90% of the CO₂ is absorbed. A CO₂ balance is maintained by expanding the CO₂ saturated solution and then flashing CO₂ vapor off the liquid at reduced pressure. A 90% CO₂ removal efficiency can be achieved in the Selexol unit and is assumed in this study.

3. PROCESS SIMULATIONS

The process simulation software CHEMCAD, developed by Chemstation Inc., was employed in this study. The software package was used to perform steady state simulations for main chemical processes and equipment involved in the power generation system. Details of mass and energy balances were obtained from the simulation.

3.1. PC_{ref} and PC-MEA plants

3.1.1. Specifics of the simulation

Both PC_{ref} and PC-MEA processes have a similar boiler section (air-blown) and a similar flue gas composition upstream of the MEA unit. In both cases, the power plant was divided into several main process areas that include the boiler, steam turbine cycle, flue gas cleaning, MEA process and the CO_2 compression.

Boiler and Steam cycle

The detailed design and operating data for the steam cycle and the boiler can be found in a DOE-funded study ^[11]. A sub-critical steam cycle was assumed. The high-pressure turbine uses superheated steam at 2,415 psi and 1,000°F. The reheat steam enters the intermediate pressure turbine at 604 psi and 1,000°F. 15% excess air was assumed for combustion (ie, the overall stoichiometry of the combustion system was 1.15).

Flue Gas cleaning

In the flue gas cleaning units, contaminant removal efficiencies of 99%, 90% and 90% were assumed for the ESP, LSD and SCR units respectively. These values are typical for the current technologies ^[12].

MEA unit

The simulation of the MEA process was conducted using the amine-specific K-value model, enthalpy model, and chemical reaction constants developed in Chemcad. The main operating conditions used in the simulation were those typical for the Fluor-Daniel Econamine FG Process ^[13, 14]. For solvent regeneration the low pressure steam at 175 psi was drawn from the LP turbine and expanded to 60 psi prior to entering the stripper re-boiler. The steam condensed from the re-boiler was returned to the steam cycle.

CO_2 compression

The CO_2 stream exiting the MEA unit is compressed to 83 bars through multi-stage compressors with inter-stage coolers. The moisture in the CO_2 stream was condensed and removed during compression. The moisture in the final CO_2 liquid stream was kept below 0.16% by mole.

3.1.2. Process Simulation results

Process parameters, flue gas composition, CO₂ product characteristics

The main process parameters are summarized in Table 3. The compositions of the flue gas before and after cleaning and the composition of the supercritical CO₂ stream are listed in Table 4.

Table 3. New-PC_{ref} and PC-MEA - Main operating parameters

	New-PC _{ref}	New PC-MEA
Gross output, MW _e	533	533
Combustion		
Excess air, /	15%	15 %
Air flow rate, lb/hr	4,029,474	4,029,474
Coal feed rate, lb/hr	524,982	524,982
Steam generation		
Hot reheat steam, lb/hr	3,022,125	3,022,125
Superheat steam, lb/hr	3,422,824	3,422,824
Steam condensate, lb/hr	2,802,051	1,331,136
Main feed-water, lb/hr	3,321,228	3,321,228
Heat duty of cooling tower, 10 ⁶ Btu/hr	2,178	1,034
Flue gas		
Mass flow rate, lb/hr (TPD)	4,526,829 (49,272)	4,526,829 (49,272)
Temperature, °F (°C)	295 (146.1)	295 (146.1)
MEA unit		
Inlet gas volume, lb/hr	-	4,683,353
Temperature, °F	-	137
MEA make-up, lb/hr	-	1,827
LP steam consumption, lb/hr	-	1,470,914
CO ₂ captured, lb/hr (TPD)	-	894,115 (9,720)

Table 4. PC-MEA - Flue gas and CO₂ stream

	① Flue gas from boiler	② Cleaned flue gas	③ CO ₂ Product before compression	④ CO ₂ Product
O ₂	2.51 %	2.37 %	11 ppm	12 ppm
N ₂	71.65 %	67.80 %	172 ppm	180 ppm
Ar	0.10 %	0.096 %	1 ppm	1 ppm
CO ₂	14.55 %	13.77 %	95.69 %	99.86 %
H ₂ O	11.15 %	15.97 %	4,2861 ppm	1,209 ppm
SO ₂	257 ppm	24 ppm	8 ppm	9 ppm
NO _x *	117 ppm	11 ppm	31 ppb	8 ppb
NH ₃	n/a	3 ppm	155 ppb	19 ppb
Pressure, psia (bara)	14.2 (0.98)	13.8 (0.95)	25.0 (1.72)	1203.8 (83)
Temperature, °F (°C)	295.0 (146.1)	156.8 (69.3)	104.0 (40.0)	68.0 (20.0)

① ② ③ ④: flue gas location, see Figure 5

* PC_{ref} NO_x concentration listed in this table (117 ppm) is for the new plant, assuming an emission factor of 0.18 lb NO_x/10⁶Btu. NO_x concentration for the retrofit is about 163 ppm based on an emission factor of 0.25 lb/10⁶Btu.

Auxiliary Power consumption

Certain components in the power plant, such as pumps and conveyors, consume electricity. The auxiliary power use of the coal handling, pulverizers and ash handling were scaled linearly from the reference plant based on the coal feed rate or ash production. For all other main components energy usage was obtained from the process simulation.

The results of the auxiliary power use are summarized in Table 5.

Table 5. New-PC_{ref} and PC-MEA - Auxiliary power usage

	New-PC_{ref}	New PC-MEA
Gross output, MW _e	533	435
Electric-driven auxiliary load, kW_e		
Coal handling	339	339
Pulverizers	2,937	2,937
Primary air fans	1,212	1,212
Forced Draft fans	1,154	1,154
Induced draft fans	5,122	5,122
Seal air blowers	46	46
Steam turbine auxiliaries	884	884
Condensate pumps	949	891
Circulating water pumps	4,187	1,989
Cooling tower fans	2,367	1,125
Ash handling	1,424	1,424
Miscellaneous	2,411	2,411
Transformer loss	1,215	1,215
ESP	1,319	1,319
FGD	3,500	3,500
SCR	2,750	2,750
MEA: induced draft fans	0	15,837
pumps	0	2,980
CO ₂ compressor	0	37,951
Total Electric-driven load	31,816	85,086
Electric equivalent of steam driven Main feed pump	10,938	10,938

The total auxiliary use of electricity of the PC_{ref} plant (no CO₂ capture) is about 6% (32MW_e) of the total gross output (533MW_e). This total increases to 16% (85MW_e) of the original 533MW_e when a MEA unit is installed, mainly due to an additional need of 57MW_e (about 10% of the PC_{ref} gross output) to operate the MEA fans, MEA pumps and CO₂ compressors. The installation of the MEA process also reduces the gross output by

18% (98MW_e , 435MW_e vs 533MW_e) due to the heat duty required for amine regeneration. All together, the MEA unit along with the typical auxiliary equipment consume an electric equivalent of 183MW_e (34%) of the 533MW_e gross power of the reference plant PC_{ref} .

3.1.3. Performance summary

The overall performances of the PC_{ref} plant and the PC-MEA plants are shown in Table 6.

Table 6. New- PC_{ref} and PC-MEA - Overall process performances

	New-PC_{ref}	New PC-MEA
Gross output (terminal), MW_e	533	533
Coal feed as-received, lb/hr	524,982	524,982
Boiler efficiency	86.4 %	86.4 %
Steam turbine power, MW_e	533.2	434.9
Auxiliary power use, MW_e	31.8	85.1
Net power output, MW_e	501.4	349.8
Net generation efficiency, % (HHV)	37.0 %	25.8 %
Heat rate, Btu/kWh (HHV)	9,214	13,208

The net power output of the PC-MEA process (350MW_e) is about 30% lower than the net power available on the reference subcritical PC plant (501MW_e). This 152MW_e net power drop is mainly due to the steam used to regenerate the amine (98MW_e , 19% of the reference PC net output), the power needed to compress the CO_2 (38MW_e , 7%) and the MEA fans/pumps operation (19MW_e , 4%).

The power generation efficiency for the sub-critical PC plant without CO_2 capture is about 37%. When the MEA unit is installed, the net generation efficiency drops to about 26%.

3.2. PC-OC Plant

3.2.1. Specifics of the simulation

The PC-OC power plant was divided into several main process areas for simulation: the boiler, steam turbine cycle, flue gas cleaning and the CO_2 compression.

Boiler and Steam Cycle

The boiler and the steam cycle were simulated using the same design and performance data used for the air-blown PC plant. However, a 3% of excess oxygen (stoichiometry 1.03) was used for oxy-combustion compared to 15% excess air (stoichiometry 1.15) for air blown PC (PC_{ref} and PC-MEA). The purpose of that is to keep 3% excess O_2 in stack. The temperature of the flue gas leaving the boiler system was kept 295°F . A flue gas recycle ratio of 60% was assumed for the new plant, while 72% was assumed for the retrofit plant case. These ratios come from separate experimental and design data (tasks 2 and 4 of this project).

Flue Gas Cleaning

The contaminant removal efficiencies of 99% and 90% were assumed for the ESP and LSD.

CO₂ drying and compression

The cleaned flue gas was condensed at 104 °F prior to compression. A three-stage compressor was employed, and the final stream pressure of 83 bar was assumed for the same reason than in PC-MEA process. The CO₂ stream was cooled to 104 °F via inter-stage cooling. In this study, further purification of CO₂ stream is not considered.

3.2.2. Process simulation results

Process parameters, flue gas composition, CO₂ product characteristics

The main process parameters are summarized in Table 7 and the compositions of the flue gas from the boiler is listed in Table 8. The process performances of the new PC-OC and the retrofit PC-OC plants are slightly different due to the different flue gas recirculation ratios adopted.

Table 7. PC-OC (New and Retrofit) - Main operating parameters

	New PC-OC	Retrofit PC-OC
Gross output, MW _e	533 MW _e	533 MW _e
Combustion		
Excess O ₂	3%	3%
O ₂ flow rate (99% purity), lb/hr (TPD)	813,423 (8,854)	820,668 (8,932)
Coal feed rate, lb/hr (TPD)	504,064 (5,486)	508,554 (5,535)
Flue gas recycle ratio	0.60	0.72
Steam generation		
Hot reheat steam, lb/hr	3,022,125	3,022,125
Superheat steam, lb/hr	3,422,824	3,422,824
Steam condensate, lb/hr	2,802,051	2,802,051
Main feed water, lb/hr	3,321,228	3,321,228
Heat duty of cooling tower, 10 ⁶ Btu/hr	2,178	2,178
Flue gas stream from boiler		
Mass flow rate, lb/hr (TPD)	3,366,231 (36,639)	4,902,386 (53,359)
Temperature, °F (°C)	295 (146.1)	295 (146.1)
CO ₂ captured, lb/hr (TPD)	924,310 (10,062)	932,507 (10,151)

Table 8. PC-OC (New and Retrofit) - Flue gas and CO₂ stream

	New PC-OC (mol%)			Retrofit PC-OC (mol%)		
	① Flue gas from boiler	② Cleaned flue gas	④ CO ₂ product*	① Flue gas from boiler	② Cleaned flue gas	④ CO ₂ product*
O ₂	2.12 %	2.02 %	3.88 %	2.02 %	1.93 %	3.88 %
N ₂	0.32 %	3.04 %	0.58 %	0.30 %	0.29 %	0.58 %
Ar	0.60 %	0.58 %	1.11 %	0.57 %	0.55 %	1.11 %
CO ₂	51.26 %	49.18 %	94.27 %	48.92 %	46.94 %	94.26 %
H ₂ O	45.67 %	47.91 %	0.12 %	48.16 %	50.27 %	0.12 %
SO ₂	383 ppm	37 ppm	70 ppm	263 ppm	25 ppm	50 ppm
NO _x	73 ppm	70 ppm	337 ppm	62 ppm	60 ppm	422 ppm
Pressure, psia (bara)	14.2 (0.98)	13.9 (0.96)	1,203.8 (83.0)	14.2 (0.98)	13.9 (0.96)	1,203.8 (83.0)
Temperature, °F (°C)	295.0 (146.1)	201.9 (94.4)	68.0 (20.0)	295.0 (146.1)	202.2 (94.5)	68.0 (20.0)

* Further CO₂ purification hasn't been included in this economic analysis for the PC-OC.

Auxiliary Power Consumption

The auxiliary power usage of the PC-OC plants is listed in Table 9.

Table 9. PC-OC (New and Retrofit) - Auxiliary power usage

	New PC-OC	Retrofit PC-OC
Electric-driven Auxiliary load, kW_e		
Coal handling	326	329
Pulverizers	2,820	2,845
Primary air fans	959	1,530
Forced Draft fans	912	1,456
* Induced draft fans	1,461	1,544
Seal air blowers	36	57
Steam turbine auxiliaries	884	884
Condensate pumps	949	949
Circulating water pumps	4,187	4,187
Cooling tower fans	2,367	2,367
Ash handling	1,367	1,379
Miscellaneous	2,411	2,411
Transformer loss	1,215	1,215
ESP	1,006	1,465
LSD	3,201	3,597
CO ₂ Compressor	44,313	44,710
PC-OC flue gas recycle	1,651	2,933
ASU	100,572	101,468
Total electric-driven load	170,637	175,326
Electric equivalent of steam driven Main feed pump	10,938	10,938

* ID fans for driving the flue gas not re-circulated.

The ASU for O₂ production consumed about 100MW_e ^[16], or 19% of the gross output. CO₂ compression is the next largest power consumer, with about 8% of the gross output (44MW_e). Overall, the auxiliary power use in the PC-OC plant accounts for 32% of the gross output compared to 6% in the PC_{ref} plant.

3.2.3. Performance summary

Table 10 lists the overall performance of the PC-OC power plants.

Table 10. PC-OC (New and Retrofit) - Overall process performances

	New PC-OC	Retrofit PC-OC
Gross output, MW_e	533.23	533.23
Coal feed as-received, lb/hr (TPD)	504,064 (5,486)	08,554 (5,535)
Boiler efficiency	90.0 %	89.2 %
Steam turbine power, MW _e	533.2	533.2
ASU Power, MW _e	100.6	101.5
Other aux. power use, MW _e	70.1	73.9
Net power output, MW_e	362.6	357.9
Net efficiency, % (HHV)	27.9 %	27.3 %
Heat rate, Btu/kWh (HHV)	12,233	12,504

The PC-OC net power output (358 to 363MW_e) is 27% to 28% lower than the net power output of the PC_{ref} plant. Such reduction is mainly due to the power consumption by ASU (100MW_e, 20% of PC_{ref} net output) and the CO₂ compressors (45MW_e, 8% PC_{ref} net output).

The generation efficiency dropped to about 28% in the new PC-OC and to about 27% in the retrofit PC-OC, compared to about 37% in the PC_{ref}.

3.3. IGCC-S Plant

3.3.1. Specifics of the simulation

The main design and operating parameters used in the simulation were referred from several recent studies ^[17, 18, 19]. The PRB coal that has been investigated in this study has lower heat value and higher moisture content than those in the literature, such as the Illinois coal. Compared to the Illinois coal, the use of the PRB coal resulted in the larger mass flow, lower BTU raw syngas and different syngas composition. Therefore, based on the literature, the process parameters were adjusted for the PRB coal where ever necessary.

The process simulation included two process blocks. One included the gasifier, the shift reactors, the saturators, the Selexol unit and the gas turbine. The other included the steam cycle. The heat exchangers in the combined cycle were balanced between the two blocks. The Chemcad package was used for the simulation. However, only a simplified mass balance calculation was conducted for the Selexol unit due to unavailability of related chemical data of the solvent. Hence, the process of selexol unit is not simulated.

Therefore, the CO₂ product composition had to be assumed instead of calculated. As a simplified approach, the CO₂ extracted by the selexol unit is supposed to be pure (100% CO₂).

3.3.2. Process simulation results

Process parameters, flue gas composition, CO₂ product characteristics

Table 11 summarizes the main performance data, while Table 12 provides information on flue gas and CO₂ product.

Table 11. IGCC-S - Main operating parameters

Gross output, MW_e	536
Net output, MW_e	436
Coal feed rate, lb/hr (TPD)	466,262 (5,074)
O ₂ (95% purity) flow rate, lb/hr (TPD)	288,533 (3,140)
Water make-up, lb/hr	584,349
Raw fuel gas, lb/hr	934,997
Slag, lb/hr	24,471
Steam to gas shift reactor, lb/hr	293,937
Air to gas combustor, lb/hr	5,022,433
CO ₂ captured, lb/hr (TPD)	788,136 (8,579)
CO ₂ capture efficiency, %	89.8%

Table 12. IGCC-S - Flue gas and CO₂ stream

	① Syn-gas exiting the gasifier	② Syn-gas exiting the water shift reactor	*③ CO ₂ product before compression	*④ CO ₂ product
H ₂ O	25.42 %	0.16 %	-	-
O ₂	-	-	-	-
N ₂	0.37 %	0.37 %	-	-
Ar	0.88 %	0.86 %	-	-
H ₂	29.22 %	55.29 %	*100 %	*100 %
CO	29.63 %	2.73 %	-	-
CO ₂	13.84 %	39.95 %	-	-
CH ₄	0.57 %	0.56 %	-	-
H ₂ S	742 ppm	732 ppm	-	-
NH ₃	64 ppm	10.4 ppb	-	-
Pressure, psia (bara)	795 (54.8)**	704 (48.5)***	*N/A	*1203.8 (83.0)
Temperature, °F (°C)	550 (287.8)**	105 (40.6)***	*N/A	*68.0 (20.0)

*Not simulated. CO₂ product is assumed pure when exiting the Selexol unit. Only CO₂ mass balance is considered. Compression energy and costs were estimated from scaling of DOE literature.

**Temperature and Pressure right after the raw gas cooler in the gasifier unit

*** Temperature and Pressure right before the Selexol Plant

Auxiliary Power consumption

The result of the auxiliary power is summarized in Table 13.

Table 13. IGCC-S - Auxiliary power usage

Gross output, MW_e	536 MW_e
Gross Power Summary, kW_e	
Gas Turbine Power	358,541
Steam Turbine Power	174,601
Generator Loss	-7,996
Turbo-Set Power	525,146
Fuel Gas Expander Power	10,611
Gross Power, kW_e	535,757
Auxiliary Load Summary, kW_e	
Coal Handling and Conveying	526
Coal Milling	1,213
Coal Slurry Pumps	235
Slag Handling and Dewatering	123
Recycle Gas Blower	127
Air Separation Plant	36,759
Oxygen Boost Compressor	16,259
Selexol Plant	7,977
Claus/TGTU	15
Tail Gas Recycle	154
Humidification Tower Pump	65
Humidifier Makeup Pump	141
Low-Pressure CO ₂ Compressor	865
High-Pressure CO ₂ Compressor	25,782
Condensate Pumps	329
High-Pressure Boiler Feed Pump	3,294
Low-Pressure Boiler Feed Pump	61
Miscellaneous Balance of Plant	1,092
Gas Turbine Auxiliaries	623
Steam Turbine Auxiliaries	244
Circulating Water Pumps	1,636
Cooling Tower Fans	1,028
Flash Bottoms Pump	55
Transformer Loss	1,661
Total	100,263

The total auxiliary power use is about 19% of the gross power output. The Selexol unit consumed about 8% (8MW_e) and CO₂ compression about 27% (26.6MW_e) of the total auxiliary power.

3.3.3. Performance summary

The overall performance data is listed in Table 14 and Table 15.

Table 14. IGCC-S – Breakdown of Gross Power Output

Steam Turbine, MW _e	358.5
Gas Turbine, MW _e	174.6
Generator Loss, MW _e	-8.0
Fuel Gas Expander, MW _e	10.6
Resulting Total Gross power, MW_e	535.7

Table 15. IGCC-S - Overall process performances

Gross output, MW_e	536
Coal feed as received, lb/hr (TPD)	466,262 (5,074)
Oxygen Feed , lb/hr (TPD)	288,533 (3,140)
Water , lb/hr (TPD)	584,349
Aux. Power Use, MW _e	100.3
Net Power, MW_e	435.5
Net efficiency, HHV (%)	36.18 %
Heat rate, Btu/kWh (HHV)	9,422

Compared to the IGCC plant without CO₂ capture, installation of the Selexol process reduced the net generation efficiency. An IGCC plant without CO₂ capture has a generation efficiency of 43%^[18]. In this study the IGCC-S plant reduced the generation efficiency to 36%, compared to 37% reported in the literature for the similar process configuration^[18]. The generation efficiency of the IGCC-S plant is comparable to the efficiency of a sub-critical air-blown PC plant with no CO₂ capture (about 37%), but lower than the efficiency of state-of-the-art air-blown supercritical boilers (about 42-43%).

3.4. Comparison of process performances

The performances of PC-OC, PC-MEA and IGCC-S are provided in the following section in terms of auxiliary power consumption, net power output, net generation efficiency and CO₂ removal efficiency (CO₂ captured/ CO₂ produced by coal combustion, %).

3.4.1. Comparison of CO₂ product composition

Table 16 shows the composition of the CO₂ from the PC-OC process (retrofit or New), the PC-MEA process and the IGCC-S plant.

For a fair comparison, the CO₂ product is compressed to the same pressure (83 bars) in the three capture processes. 83 bars is 10 bars greater than the critical pressure (“P_c”) of CO₂ (P_c=73.8bars). The CO₂ product is thus in supercritical stage for pipeline transportation. Depending on pipeline requirements and the CO₂ application, the required

CO₂ pressure may be higher than 83 bars. Additional compression would then have to be added to the three CO₂ capture processes presented in this report.

Table 16. PC-OC, PC-MEA and IGCC-S - Comparison of CO₂ products

	PC-OC		New or retrofit PC-MEA	IGCC-S
	New PC-OC	Retrofit PC-OC		
O ₂	3.88 %	3.88 %	12 ppm	-
N ₂	0.58 %	0.58 %	180 ppm	-
Ar	1.11 %	1.11 %	1 ppm	-
CO ₂	94.27 %	94.26 %	99.86 %	*100 %
H ₂ O	0.12 %	0.12 %	1,209 ppm	-
SO ₂	70 ppm	50 ppm	9 ppm	--
NO _x	337 ppm	422 ppm	32 ppb	
NH ₃	-	-	154 ppb	-
Pressure, psia (bara)	1,203.8 (83.0)	1,203.8 (83.0)	1,203.8 (83.0)	1,203.8 (83.0)
Temperature, °F (°C)	68 (20)	68 (20)	68 (20)	68 (20)

*Simplified assumption. Not simulated.

The CO₂ purity from the selexol or MEA units approaches 100% as those processes are extremely selective. The CO₂ purity from the PC-OC process is about 94%, since no purification has been simulated. Note also that no air infiltration has been included in this study, which would lower the CO₂ purity from the PC-OC process. Depending of the CO₂ application, the CO₂ purity of the PC-OC process can be increased to the required level by adding some purification equipment in CO₂ compression train.

3.4.2. Comparison of plant performances

Retrofit Applications

Figure 7 and Table 17 show the comparison of performances (Net power output, plant efficiency) of the two CO₂ capture technologies applicable for retrofit of existing PC plants: PC-OC and PC-MEA.

Table 17. Performance comparison for retrofit plants

	Existing-PC _{ref}	Retrofit PC-OC	Retrofit PC-MEA
Gross Power Output	533.2 MW _e	533.2 MW _e	533.2 MW _e
MEA regeneration (steam)	-	-	98.4 MW _e
ASU Power consumption	-	101.5 MW _e	-
Total auxiliary Power consumption other than ASU and MEA	31.8 MW _e	73.9 MW _e	85.1 MW _e
Net power Output	501.4 MW_e	357.9 MW_e	349.8 MW_e
Net Power output loss	0	- 143.5 MW _e (-28.6 %)	- 151.6 MW _e (-30.2 %)
Plant net efficiency	37 %	27.3 %	25.8 %
CO ₂ capture Efficiency ²	0 %	96 %	90 %

Figure 7. Performance comparison for retrofit plants

As mentioned before (see 3.1.2 and 3.2.2), the power consumption of the ASU and MEA system and the power needed to capture and compress the CO₂ to the required pressure (83bars in this study) increase the auxiliary power need of the PC plants from 32 MW_e (6% of the gross power output) in the PC_{ref} plant, to 175MW_e (32.8%) in the PC-OC and 183MW_e (34.3%) in the PC-MEA retrofit plants. As a result, the net power output of the

² The CO₂ capture efficiency is defined as the % CO₂ captured from the full flue gas stream of the same power plant

CO₂ capture plants is reduced from 500MW_e without capture to 358MW_e when capture is performed by PC-OC process, and 350MW_e when the PC-MEA system is installed. Note that for both amine and OC systems, the power consumption of the MEA and ASU is directly related to the coal flow rate, not to the MW_e. A more efficient power plant, like a super or ultra-supercritical plant, with net generation efficiency between 40 and 50% (rather than 37% in this study) would produce less CO₂ per MW_e and thus require a smaller ASU or MEA unit for CO₂ capture (thus smaller auxiliary power consumption).

New plants Applications

Table 18 provides a summary and comparison of the performances of three CO₂ capture technologies applicable to new coal-fired power plants. Note that a new PC plant will usually be supercritical, not sub-critical as the PC plants considered in this study. This table is thus for information purpose only and will be updated in future studies with supercritical, state of the art PC plants for a fair comparison with IGCC units.

Table 18. Performance comparison for new plants

	New-PC_{ref}	New PC-OC	New PC-MEA	IGCC-S
Gross Power Output	533 MW _e	533 MW _e	533 MW _e	536 MW _e
MEA regeneration (steam)	-	-	98.4 MW _e	-
ASU Power consumption	-	100.6 MW _e	-	36.8 MW _e
Auxiliary Power consumption other than ASU or MEA	31.8 MW _e	70.1 MW _e	85.1 MW _e	63.5 MW _e
Net power Output	501.4 MW_e	362.6 MW_e	349.8 MW_e	435.5 MW_e
Net Power output loss	0	- 138.8 MW _e (-27.7 %)	- 151.6 MW _e (-30.2 %)	- 65.9 MW _e (-13.1 %)
Plant net efficiency	37 %	27.9 %	25.8 %	36.2 %
CO ₂ capture efficiency	0 %	96 %	90 %	90 %

4. COST ESTIMATIONS

4.1. Methodology

The capital cost, the O&M cost and the cost of electricity for the power plants were estimated according to the methodology described below^[20].

4.1.1. Capital Cost

There are three definitions for the capital cost, the Total Plant Cost (TPC), the Total Plant Investment (TPI), and the Total Capital Requirement (TCR). Their definitions are defined in Table 19.

The Total Plant Cost consists of Bare Erected Cost (BEC), engineering & home office overheads and fees, and contingencies.

Table 19. Items in capital cost estimation

(a)	Bare erected cost
(b)	Engineering & home office fee
(c)	Process contingency
(d)	Project contingency
Total Plant Cost (TPC) = a + b + c + d	
(e)	Total cash expended
(f)	AFDC (escalation and interest in construction period)
Total plant investment (TPI) = TPC + e + f	
(g)	Royal allowance
(h)	Pre-production costs
(i)	Inventory capital
(j)	Initial catalyst & chemicals
(k)	Land cost
Total Capital Requirement (TCR) = TPI + g + h + i + j + k	

The Total Plant Investment includes the escalation of construction costs and interest during construction plus the Total Plant Cost. The Total Plant Investment can be calculated according to:

$$TPI = TPC \cdot A \cdot \frac{\left[\left(\frac{1+i}{1+ea} \right)^n - 1 \right]}{\left(\frac{1+i}{1+ea} \right)} \quad (1)$$

where A = cost expended per year, 1/n;
n = construction years;
i = weighted cost of capital;
ea = inflation rate.

The Total Capital Requirement covers all other expenditures needed to complete an entire plant. These include the preproduction, spare parts, inventory capital, initial chemicals

and land costs. The assumptions used for calculating the TCR are listed in Table 20 ^[12].

Table 20. Assumptions for estimating the Total Capital Requirement

Prepaid Royalties	0 % for PC-based plant; 0.5% for IGCC-S
Pre-production	
O&M cost (except for fuel)	30 days
Fuel	7.5 days
Plant modifications	2% of total plant investment (TPI)
Inventory capital	
Coal	60 days
Consumables (excluding water)	60 days
Spare Parts	0.5 % of total plant cost (TPC)
Initial Catalyst and Chemical Inventory	30 days
Land	\$6,500/Acre

The TPC for air-blown PC plant, the PC-OC plant and the IGCC-S plant was estimated as follows. Based on the TPC, the TPI and TCR were determined according to Equation (1) and Table 19.

PC_{ref} and PC-MEA Plants

The PC_{ref} and PC-MEA plants include the basic PC plant, the flue gas cleanup units LSD, SCR and the MEA unit if used.

(1) Basic components of a PC plant

In a conventional sub-critical air-blown PC plant without CO₂ capture, 13 process areas are involved as listed in Table 21.

Table 21. PC_{ref}, PC-MEA and PC-OC - Basic components

	Process area	Engineering fee*	Project contingency*	Parameter for plant size scaling
1	Coal handling	8.0%	21.6%	Coal feed rate
2	Coal preparation & feed	8.0%	21.6%	Coal feed rate
3	Feed water & misc.	8.0%	25.6%	Water feed rate
4	PC boiler & accessories	8.0%	10.8%	Plant capacity
5	Electro-static precipitator	8.0%	10.8%	Flue gas flow rate
6	HRSG, ducting and stack	8.0%	16.7%	Flue gas flow rate
7	Steam turbine generator	8.0%	13.5%	Steam flow rate
8	Cooling water system	8.0%	19.3%	Heat duty of cooling water tower
9	Ash/spent sorbent handling system	8.0%	16.4%	Ash in coal feeding
10	Accessory electric plant	8.0%	17.7%	Plant capacity
11	Instrumentation & control	8.0%	16.8%	Plant capacity
12	Improvements to site	8.0%	32.4%	Plant capacity
13	Buildings and structures	8.0%	27.0%	Plant capacity

* Percentages of the bare erected cost.

The bare erected costs for each process area were scaled from DOE studies on a 401MW_e.

(net) reference plant ^[12,20]. An exponential scaling factor of 0.72 was adopted^[21]. Specific scaling parameters to individual process areas are listed in the last column of Table 21. The cost scaling was carried out as follows,

$$\begin{aligned} BEC_{i1} &= BEC_{i0} \times (P_{i0}/P_{i1})^{0.72} \\ TPC &= BEC \times (1 + C1) \times (1 + C2) \times (1 + C3) \end{aligned} \quad (2)$$

where BEC = bare erected cost;

P = scaling parameter;

i = process area;

C1 = engineering fee;

C2 = process contingency;

C3 = project contingent.

The subscripts 1 and 0 stand for the plant under this study and the DOE reference plant, respectively. C1 and C3 are listed in Table 21; C2 was assumed 0 for the conventional processes.

(2) LSD for SO₂ removal

The total plant cost for the LSD unit with a 90% SO₂ removal efficiency was based on the cost model developed by the US EPA^[22, 23].

(3) SCR for NO_x removal

The cost model developed by the US EPA was adopted to estimate the capital cost of the SCR unit with a NO_x removal efficiency of 90%^[24]. This model is applicable for the coal-fired boilers with capacity of 100 to 850 MW_e and NO_x inlet concentrations of 0.15-2.5 lb/10⁶Btu.

The model gave the result of the TCR. The TPC was assumed to be 10% less than the TCR in the SCR case.

(4) MEA for CO₂ capture

The capital cost of the MEA unit was estimated by the scaling approach. Given the fact that the gas stream conditions, such as temperature and CO₂ concentration in the gas, are similar for different plants and coals, the inlet flow rate of the gas stream was selected as the scaling parameter. The MEA unit for a 402 MW_e PC plant in a recent DOE study^[18] was selected as the scaling reference. An exponential scaling factor of 0.6 was assumed^[13]. The total plant cost was estimated as follows.

$$TPC_{MEA} = TPC_{ref} \times \left(\frac{GF}{GF_{ref}} \right)^{0.6} \quad (3)$$

where TPC_{ref} = total plant cost of a reference MEA unit

GF = gas flow rate of the investigated plant

GF_{ref} = gas flow rate of a reference plant

PC-OC Plant

For the PC-OC plants, the total plant cost of equipment common to those involved in the

air-blown PC plant was estimated using the same cost models as described above. The capital cost of equipment specific to the PC-OC process was estimated according to the following approaches.

(1) ASU for O₂ generation

The total plant cost per unit oxygen production is tabulated in Table 22 ^[16]. 99% purity O₂ was used in the PC-OC process and 95% purity O₂ in the IGCC-S case.

Table 22. Capital cost of oxygen production

Oxygen production		O ₂ purity	Capital Cost	
lb/hr	mt/day	%	k\$	k\$/(mt/day)
917,768	10,000	99 %	138,059	13.81
275,330	3,000	95%	54,594	18.20

(2) Flue gas handling

The capital cost of the flue gas handling system involves the gas recirculation loop (ID fan and piping) and the ID fan for driving gas to the condenser and compressors. The existing ID fan of the boiler system can be used for re-circulating the flue gas; its cost therefore is not counted again.

The duct of the recirculation loop was assumed to be the steel duct and 300 ft long with a gas velocity of 25m/s. Its total plant cost is a sum over the estimated material cost of commercial steel, auxiliary materials such as insulation and valves (50% of the steel cost was assumed), and installation fees (1.5 times of the total material cost).

The ID fan for transporting gas to the condenser and compressor was based on the cost model available in the Chemcad. Two parallel fans were used, each handling half of the total gas flow.

(3) Flue Gas Condenser

The capital cost of the flue gas condenser was scaled from that of an 865 MW OC plant ^[25]. The capital cost was estimated according to the following power rule.

$$TPC_{conds} = C_{ref} \times \left(\frac{FG}{FG_{ref}} \right)^{0.72} \times \left(\frac{HD}{HD_{ref}} \right)^{0.72} \quad (4)$$

where C_{ref} = capital cost of the reference condenser, \$;

FG = flue gas flow rate, cfm;

HD = Heat duty of condenser, btu/h.

(4) CO₂ compression

The purchase prices of the compressor and inter-stage cooler were obtained from Chemcad.

IGCC-S Plant

The IGCC-S plant was categorized into 16 process areas for the cost estimation (Table

23). The total plant cost of main equipment was scaled from several IGCC case studies^[17,26,27]. The cost scaling was based on scaling parameters specific to individual process areas (Table 23). A power law with an exponent of 0.72 was also assumed for the IGCC plant^[21].

Table 23. IGCC-S plant - Classification of process areas

	Process area	Scaling parameters
1	Coal handling	Coal feed rate
2	Coal preparation & feed	Coal feed rate
3	Feed water & miscellaneous Bop systems	Water feed rate
4	Gasifier & accessories	Coal slurry feed rate
5	ASU and accessories	Oxygen feed rate
6	Gas cleanup and piping	Synthesis gas flow rate
7	CO ₂ compression	CO ₂ captured
8	Gas turbine and accessories	Gross capacity
9	HRSB, ducting and stack	Flue gas flow rate
10	Steam turbine generator	Steam flow rate
11	Cooling water system	Heat duty of cooling water tower
12	Ash /spent sorbent handling system	Ash in coal feed
13	Accessory electric plant	Gross capacity
14	Instrumentation & control	Gross capacity
15	Improvements to site	Gross capacity
16	Buildings and structures	Gross capacity

4.1.2. O&M cost

Fixed O&M cost

The fixed O&M cost consists of the operating labor, the maintenance material & labor, and the administrative & support labor.

The cost of operating labor (OL) was estimated based on the number of operating jobs (OJ) and the labor rate.

$$OL = \text{Labor rate} \times OJ \times 40 \text{ hrs / week} \times 52 \text{ weeks / yr} \quad (6)$$

Annual maintenance labor and material cost (ML & M) was estimated as a percentage of the bare erected cost. Among the total maintenance costs it was assumed that 40% was shared by the maintenance labor. The following equation was used for the annual maintenance cost estimation.

$$ML \& M = \sum_i F_i \times BEC_i \quad (7)$$

where F = maintenance factor;
BEC = bare erected cost;
i = process area.

Administrative & support labor cost was assumed 30% of the total costs of operating labor and maintenance labor.

$$A \& S = 0.3 \times (0.4 \times ML \& M + OL) \quad (8)$$

The assumptions used in the fixed O&M cost estimation are listed in Table 24.

Table 24. Assumptions used for estimating the fixed O&M cost

Fixed O&M cost	Assumptions
Operating labor	Labor rate: \$40/hr/person Labor shift: 4 shifts/day Operating jobs: (1) PC _{ref} plant without CO ₂ capture: 25 persons/shift ^[12] (2) PC-MEA : 27 persons/shift (2 for MEA ^[13]) (3) PC-OC plant: 29 persons/shift (2 for ASU, 2 for CO ₂ comp and purification, if needed) (4) IGCC-: 17 persons/shift (15 for IGCC ^[26] , and 2 for Selexol)
Maintenance	Percentage of TPC: (1) PC _{ref} plant without CO ₂ capture: EPRI guidelines ^[12] (2) PC-MEA and others related to CO ₂ capture: 2.5% ^[13] (3) ASU and others related to CO ₂ capture: 2.5% (4) IGCC-S: 2.2% ^[26]
Administrative & support cost	30% of O&M labor

Variable O&M cost

The variable O&M cost is due to the expenditure of coal, consumables (such as chemicals and water) and waste disposal. The consumptions of fuel/consumables and the generation of the waste were mainly obtained from the process simulation. The unit purchase price and the unit cost of waste treatment are listed in Table 25. A 70% loading factor was assumed for all types of power plants investigated.

Table 25. Unit prices of coal, consumables and waste treatment

Category	Unit price
PRB Coal	20 \$/t
Water	0.06 \$/t
Waste water treatment chemicals	0.13 \$/lb
Ash disposal	10 \$/t
Lime	65 \$/t
Steam	7 \$/t
Ammonia	225 \$/t
SCR catalyst	N/A (total cost estimated according to [10])
MEA	1250 \$/metric t
MEA inhibitor	250 \$/metric t
Waste disposal from MEA unit	175 \$/metric t
Caustic sodium	2000 \$/metric t
Activated carbon	2000 \$ metric t
Selexol solvent	1.45 \$/lb
Claus Catalyst	470 \$/t
SCOT activated alumina	0.067 \$/lb
SCOT cobalt Catalyst	6000 \$/t
SCOT chemicals	275 \$/t
Credit of sulfur by-product	\$75/t

4.1.3. Cost of Electricity

The cost of electricity (COE) is the levelized (over the plant life – 20 years) coal pile-to-busbar cost of power expressed in mills/kWh. One cent of \$ equals ten mills.

The financial criteria used for estimating the cost of electricity (Table 26) were adopted from a recent DOE study ^[18]. Based on these assumptions a levelized carrying charge (applied to TCR) was calculated as 0.138. No inflation was considered for categories involved in the O& M costs.

A levelized busbar COE for a 70% design capacity factor can be calculated according to the following equation.

$$COE = \frac{13.8\% \times TCR + O \& M \text{ cost } t}{MW_{net} \times 0.7 \times 8760} \quad (9)$$

where COE: cost of electricity, mills/kWh;
 MW_{net} = new electricity output, MW_e;
 TCR: total capital requirement, \$;
 O&M cost: annual O&M cost, \$;

Table 26. Financial criteria for cost estimation

General characteristics		
Plant startup date (year)	2000	
Capital cost year dollars	2000	
Construction period	3 years for PC-based plant; 4 years for IGCC	
Capacity factor	70%	
Financial assumptions		
Project book life	20 years	
Book salvage value	0%	
Project tax life	20 years	
Tax depreciation method	ACRS class	
Property tax rate	1% per year	
Insurance tax rate	1% per year	
Federal income tax rate	34%	
State income tax rate	4.2%	
Investment tax credit% eligible	0%	
Capital structure	% of total	Cost, %
Common equity	42	12
Preferred stock	10	8.5
Debt	45	9.0
Weighted cost of capital (after tax)	8.76%	
Escalation rates (apparent)		
General escalation	0%	
Primary/secondary fuel price escalation	0%	

4.1.4. Costs of CO₂ Avoidance and CO₂ Capture

Two indicators are commonly used to calculate and compare the cost of CO₂ capture systems:

- The cost per ton of CO₂ captured, or CO₂ capture cost
- The cost per ton of CO₂ avoided, or CO₂ avoidance cost.

The definitions of both indicators are given below, along with corresponding mathematical expressions, for retrofit and new plant applications. The calculation of those indicators depends on the selection of a reference plant that does not feature any CO₂ capture capability. The reference plants used in this study for retrofit and new power plants are described.

In the CO₂ cost formula, the following abbreviations are made:

- cap = the capture plant (power plant with CO₂ capture equipment): PC-OC, PC-MEA or IGCC-S in this study
- ref = the reference plant (air-blown PC plant without CO₂ capture): new-PC_{ref} or existing-PC_{ref} in this study.
- \$/kWh = levelized COE

New Plant Application (new PC-OC, PC-MEA, IGCC-S)

For new power plant CO₂ costs assessment, the reference plant used in the formula is the new-PC_{ref} as defined in 2.1.1. As a reminder, the new-PC_{ref} features a SCR and a FGD and has same gross power output as the capture plants.

CO₂ capture cost (new plant)

The captured CO₂ is calculated as the difference between the CO₂ emitted into the atmosphere during a specific period of time (one day for example) by the reference plant (tonCO_{2emitted})_{ref} minus the CO₂ emitted by the capture plant (tonCO_{2emitted})_{cap}, divided by the kWh generated by the capture plant (kWh)_{cap} during the same period. The captured CO₂ is expressed in tons per kWh generated by the capture plant.

$$\text{CO}_2 \text{ captured (tons/kWh)} = \frac{(\text{tonCO}_{2\text{emitted}})_{\text{ref}} - (\text{tonCO}_{2\text{emitted}})_{\text{cap}}}{(\text{kWh})_{\text{cap}}} \quad (10)$$

The cost of captured CO₂ is defined as the increase in cost of electricity from the reference (\$/kWh)_{ref} to the capture plant (\$/kWh)_{cap}, per ton of captured CO₂. The CO₂ capture cost is given in \$ per ton of CO₂ captured.

$$\text{Cost of CO}_2 \text{ capture (\$/ton)} = \frac{(\$/\text{kWh})_{\text{cap}} - (\$/\text{kWh})_{\text{ref}}}{\text{CO}_2 \text{ captured}} \quad (11)$$

CO₂ avoidance cost (new plant)

The avoided CO₂ is calculated as the difference between:

1/ The CO₂ emitted into the atmosphere by the reference plant (tonCO_{2emitted})_{ref} per kWh generated by the reference plant (kWh)_{ref}, minus

2/ The CO₂ emitted into the atmosphere by the capture plant (tonCO_{2emitted})_{cap} per kWh generated by the capture plant (kWh)_{cap}

The avoided CO₂ is expressed in tons per kWh.

$$\text{CO}_2 \text{ avoided (tons/kWh)} = \frac{(\text{tonCO}_{2\text{emitted}})_{\text{ref}}}{(\text{kWh})_{\text{ref}}} - \frac{(\text{tonCO}_{2\text{emitted}})_{\text{cap}}}{(\text{kWh})_{\text{cap}}} \quad (12)$$

Note that the CO₂ avoided can also be expressed in a more meaningful way:

$$\text{CO}_2 \text{ avoided (tons/kWh)} = \frac{(\text{tonCO}_{2\text{emitted}})_{\text{ref}} - (\text{tonCO}_{2\text{emitted}})_{\text{cap}} - \left(\frac{(\text{tonCO}_{2\text{emitted}})_{\text{ref}}}{(\text{kWh})_{\text{ref}}} \Delta \text{kWh}_{\text{ref-cap}} \right)}{(\text{kWh})_{\text{cap}}} \quad (13)$$

where:

- ΔkWh_{ref-cap} = Decrease in kWh net output from reference plant to capture plant

Expression (11) highlights the difference between CO₂ captured and CO₂ avoided, by displaying the term $\left(\frac{\text{tonCO}_{2\text{emitted}}}{\text{kWh}} \right)_{\text{ref}} \Delta \text{kWh}_{\text{ref-cap}}$. This term represents the CO₂ that would be emitted by a no-capture power plant of same capacity than the lost net capacity from the reference plant to the capture plant $\Delta \text{kWh}_{\text{ref-cap}}$. This no-capture plant is assumed to discharge CO₂ into the atmosphere at the same rate than the reference plant, $\left(\frac{\text{tonCO}_{2\text{emitted}}}{\text{kWh}} \right)_{\text{ref}}$.

The cost of avoided CO₂ is defined as the increase in cost of electricity from the reference (\$/kWh)_{ref} to the capture plant (\$/kWh)_{cap}, per ton of avoided CO₂. The CO₂ capture cost is given in \$ per ton of CO₂ captured.

$$\text{Cost of CO}_2 \text{ avoidance (\$/ton)} = \frac{(\$/\text{kWh})_{\text{cap}} - (\$/\text{kWh})_{\text{ref}}}{\text{CO}_{2\text{ avoided}}} \quad (14)$$

Retrofit Plant Application (retrofit PC-OC, retrofit PC-MEA)

For retrofit power plant CO₂ costs assessment, the reference plant used in the formula is the retrofit-PC_{ref} as defined in 2.1.2. As a reminder, the retrofit-PC_{ref} has same gross power output as the capture plants but does not feature neither SCR nor FGD. The CO₂ capture and avoidance costs are calculated using formula that better highlight the meaning of those costs than formula (11) and (14) used for new plant applications.

CO₂ capture cost (retrofit plant)

$$\text{Cost of CO}_2 \text{ captured (\$/ton): } \frac{\frac{(\$_{\text{added_equipment}})_{\text{cap_vs_ref}}}{(\text{kWh})_{\text{cap}}} + \frac{(\$_{\text{electricity_loss}})_{\text{cap_vs_ref}}}{(\text{kWh})_{\text{cap}}}}{\text{CO}_{2\text{ captured}}} \quad (15)$$

Where:

- \$_{added_equipment} = levelized capital and O&M cost of the equipment added on an existing PC unit to capture the CO₂ (ASU, FGR, MEA, CO₂ compressor etc...)
- \$_{electricity_loss} = cost of the net power lost from existing PC unit to the retrofitted capture plant.

CO₂ avoidance cost (retrofit plant)

$$\text{Cost of CO}_2 \text{ avoided (\$/ton): } \frac{\frac{(\$_{\text{added_equipment}})_{\text{cap_vs_ref}}}{(\text{kWh})_{\text{cap}}} + \frac{(\$_{\text{electricity_loss}})_{\text{cap_vs_ref}}}{(\text{kWh})_{\text{cap}}}}{\text{CO}_{2\text{ avoided}}} \quad (16)$$

Comparison of CO₂ capture and avoidance costs.

For CO₂-capture systems that are very energy intensive, the costs per ton of CO₂ removed and avoided are very different. To take into account the reduced net power output resulting from CO₂ capture, the cost of CO₂ avoided is more relevant.

A theoretical example is provided in Table 27 to illustrate the difference between CO₂ avoidance and capture costs.

Table 27. CO₂ avoidance and capture costs - Theoretical example

	Reference Plant	CO ₂ -capture plant
Plant Capacity	100 MW _e	80MW _e
Loss of Power	-	20MW _e (Assumed loss of Power = 20% _e)
CO₂ emitted	2000 TPD	200TPD (Assumed CO ₂ capture rate = 90% $\Rightarrow$ 1800 TPD)
COE	\$0.05/kWh	\$0.07/kWh
CO₂ captured	-	$(2000-200)/(80*24) = \mathbf{0.94 \text{ tons/MWh}}$ 2000-200=1800 TPD of CO ₂ are captured
CO₂ avoided	-	$2000/(100*24)-200/(80*24) = \mathbf{0.73 \text{ tons/MWh}}$ 2000-200-20*2000/100=1400 TPD of CO ₂ are avoided
CO₂ captured cost	N/A	$(0.07-0.05)*1000/0.73 = \mathbf{27.4 / \text{ton CO}_2 \text{ captured}}$
CO₂ avoidance cost	N/A	$(0.07-0.05)*1000/0.94 = \mathbf{21.3 / \text{ton CO}_2 \text{ avoided}}$

As avoided CO₂ is always smaller than captured CO₂, the cost of CO₂ avoidance is always greater than the cost of CO₂ captured. Note that the reference plant and the state of the CO₂ product play a very significant role in those costs. A different reference plant (equipped with different flue gas treatment for example, or at same net power output instead of same gross power output than the capture plant) would result in different CO₂ cost of the capture plant. Similarly, CO₂ compression and purification impact the cost of the CO₂ capture and avoidance cost, as it involve additional capital and operating cost, along with more auxiliary power consumption. Comparison between CO₂ costs (capture and avoidance) from different studies should thus be performed using caution. All calculations should refer to the same plant capacity, should use the same reference plant, and should result to similar CO₂ final pressure and temperature.

4.2. Results of Cost Estimation

4.2.1. Cost of PC_{ref} and PC-MEA plants

A costs analysis was conducted on the 533MW_e (gross) plant. Table 28 summarizes the results for the PC_{ref} and PC-MEA processes.

Table 28. New PC_{ref} and PC-MEA - Cost analysis

	New PC_{ref}	New PC-MEA
Gross, MW_e	533	533
Net output, MW_e	501	350
Capital cost, \$/kW_e net		
Total Plant Cost	1,115	2,000
Total Plant investment	1,216	2,181
Total capital requirement	1,276	2,289
O & M costs, mills/kWh		
Fixed O&M	5.60	9.55
Variable O&M	3.02	5.14
Consumables	2.44	8.77
Fuel cost	10.47	15.01
Annual carrying charge, mills/kWh	28.72	51.52
Levelized cost		
Cost of electricity, mills/kWh	50.24	89.99
Cost of CO₂ avoidance, \$/mt	-	51.88
Cost of CO ₂ removal, \$/mt	-	34.45

The total capital cost requirement of the PC_{ref} plant was \$1,276/kWe and the cost of electricity was 50 mills/kWh.

The installation of an MEA unit increased the cost of electricity generation. Compared to the PC_{ref} plant, the capital cost per unit net output and the cost of electricity of the PC-MEA plant increased by about 79% (to 90 mills/kWh). The cost of CO₂ avoidance is about \$52/mt CO₂ avoided.

Table 29 lists the breakdown costs of the PC-MEA process.

Table 29. New PC-MEA – Breakdown of specific equipment costs

Gross output, MW _e	533
Net output, MW _e	350
Total Plant Cost, k\$	152,384
O&M cost, k\$	
1. Fixed O&M	
Operating Labor	701
Maintenance cost	3,810
Administration & support labor	667
Subtotal	5,178
2. Variable O&M	
Amine make-up	6,356
Inhibitor	1,271
NaOH	643
Activated carbon	371
Waste disposal	61
Water	2,622
Subtotal	11,324

Besides a penalty of 20% reduction of the gross power output, the PC-MEA process also contributed to about 22% of the capital cost and about of 19% the O&M cost.

4.2.2. Cost of the new PC-OC Plant

The cost estimation conducted for the PC-OC plant is summarized in Table 30.

Table 30. New PC-OC - Cost analysis

	New PC_{ref}	New PC-OC plant
Gross output, MW _e	533	533
Net output, MW _e	501	363
Capital cost, \$/kW_e		
Total Plant Cost	1,115	1,861
Total Plant investment	1,216	2,028
Total capital requirement	1,276	2,123
O& M costs, mills/kWh		
Fixed O & M	5.60	9.35
Variable O&M	3.02	5.04
Consumables	2.44	4.75
Fuel cost	10.47	13.90
Levelized cost		
Cost of electricity, mills/kWh	50.24	80.82
Cost of CO₂ avoidance, \$/mt	-	35.95
Cost of CO₂ removal, \$/mt	-	26.73

Compared to the New PC_{ref} plant, the PC-OC power plant increased the unit capital cost by 67% and cost of electricity by about 60% (to 81 mills/kWh). The cost of CO₂ avoidance was estimated as \$36 /mt.

The cost breakdown of the equipment specific to the oxy-combustion process is listed in Table 31.

Table 31. New PC-OC – Breakdown of specific equipment cost

	New PC-OC
TPC, k\$	
ASU	120,839
Flue gas recycle	2,729
Flue Gas Condenser	3,957
CO ₂ Compressor	32,716
Total	160,241
O&M cost, k\$	
1. FOM	
Operating Labor	1,402
Maintenance labor& materials	4,006
Administration & support labor	901
subtotal	6,309
2. VOM	
Cooling water	3,721

4.2.3. Cost of IGCC-S Plant

The cost estimation was conducted for 536MW_e (gross) IGCC plant installed with Selexol process (IGCC-S). The results are listed in Table 32.

Table 32. IGCC-S - Cost analysis

Gross output, MW_e	536 MW_e
Net output, MW _e	435.5
Capital cost, \$/kW _e	
Total Plant Cost	1,841
Total Plant investment	2,097
Total capital requirement	2,187
O&M costs, mills/kWh	
Fixed O&M	10.89
Consumables	1.06
Fuel cost	10.71
Sulfur credit	0.09
Annual carrying charge, mills/kWh	49.21
Levelized cost	
Cost of electricity, mills/kWh	71.78
Cost of CO₂ avoidance, \$/mt CO₂	26.51
Cost of CO ₂ capture, \$/mt CO ₂	24.22

The results showed that the levelized COE is 72 mills/kWh. The IGCC-S plant has a high capital cost per unit net output, about 65% more than the New PC_{ref} plant. The O&M cost per unit electricity generation is relatively low, a level same as the PC plant. The cost of CO₂ avoidance was estimated to be \$27/mt CO₂.

4.3. Comparison of the Costs

4.3.1. New Plant Application

The cost analyses above have shown the capital cost, O&M cost, cost of electricity and CO₂ avoidance cost for the new PC_{ref} plant, the new PC-MEA plant, the new PC-OC plant and the new IGCC-S plant. Those costs are summarized in Table 33.

Table 33. New PC-OC, PC-MEA and IGCC-S - Recap of costs

	No CO ₂ capture	CO ₂ capture plants		
	New-PC _{ref}	New PC-OC	New PC-MEA	IGCC-S
Net Power Output	501 MW _e	363 MW _e	350 MW _e	435 MW _e
Total Plant Cost (TPC), \$/kW_e	1,115	1,861	2,000	1,841
Total Plant Investment (TPI) , \$/kW _e	1,216	2,028	2,181	2,097
Total capital Requirement (TCR) , \$/kW_e	1,276	2,123	2,289	2,187
Levelized capital costs, mills/kWh	28.72	47.78	51.52	49.21
Levelized O&M, mills/kWh	21.53	33.04	38.47	22.57
Levelized COE, mills/kWh	50.24	80.82	89.99	71.78
Increase in COE, mills/kWh	-	30.58	39.75	21.54
Cost of CO₂ avoidance, \$/mt CO₂	-	35.95	51.88	26.51
Cost of CO ₂ capture, \$/mt CO ₂	-	26.73	34.45	24.22

Compared to the PC_{ref} plant, the PC-MEA, PC-OC and the IGCC-S plants increased the total plant cost by 79%, 67% and 65%, respectively (Figure 8).

Figure 8. New PC-OC, PC-MEA and IGCC-S - Total plant costs (\$/kWe)

Figure 9 displays the increase in cost of electricity for the same processes.

Figure 9. New PC-OC, PC-MEA and IGCC-S – COE and cost of CO₂ avoided

The results showed that the PC-MEA plant is the most expensive, about 79% (40 mills/kWh) higher than the PC_{ref} plant. The PC-OC plant resulted in 60% (31 mills/kWh) increase in the COE. The IGCC-S plant is the cheapest; its COE is about 43% (22 mills/kWh) higher than the conventional PC_{ref} plant.

The costs of CO₂ avoidance are also compared in Figure 9. The CO₂ emissions avoided for different CO₂ capture technologies were based on the same baseline, i.e. the New PC_{ref}. The PC-MEA process resulted in a CO₂ cost of \$52/mt CO₂ avoided, while those for the PC-OC plant are \$36/mt CO₂ avoided. The CO₂ avoidance cost for IGCC-S was \$27/mt CO₂ avoided.

4.3.2. Retrofit Application

As both the PC-MEA process and PC-OC process are also applicable for the existing plant, their retrofit costs were also evaluated and compared. The IGCC-S process was not discussed for retrofit as it is a new power generation technology and is not an option to retrofit the existing fleet.

In the retrofit case, the cost of the basic power plant are assumed to be paid off. The expenditure only involved those process components to be constructed or retrofitted. These cost items included the LSD, SCR, MEA and PC-OC specific equipment. The results of retrofit costs for PC-MEA plant and the PC-OC plant are summarized in Table 34.

Table 34. Retrofit PC-OC and PC-MEA - Capital and O&M costs

	Retrofit PC-OC	Retrofit PC-MEA
Net output, MW _e	358	350
Total Plant Cost, \$/kW_e		
ASU & CO ₂ compression	452	-
MEA & CO ₂ compression	-	436
SCR	-	86
LSD	142	143
Subtotal	595	665
O&M cost, mills/kWh		
1. Fixed O&M		
ASU & CO ₂ compression	2.90	-
MEA & CO ₂ compression	-	2.41
SCR	-	0.09
LSD	1.15	1.14
Subtotal	4.04	3.65
2. Variable O&M		
ASU&CO ₂ compression	1.73	-
MEA&CO ₂ compression	-	5.28
SCR	-	0.34
LSD	0.63	0.62
Subtotal	2.36	6.25
Total O&M cost	6.41	9.89

In the PC-MEA retrofit case, the total plant costs involving installations of MEA, LSD and SCR were estimated to be 665 \$/kW. The O&M cost due to plant retrofit was 9.89 mills/kWh, of which 78% was due to the MEA unit and CO₂ compression.

Compared to the PC-MEA retrofit, the PC-OC retrofit required both a lower capital cost and a lower O&M cost. The total capital cost of the PC-OC retrofit was estimated as \$595/kW_e. The total O&M cost was estimated as 6.41 mills/kWh, of which about 72% was attributed to the ASU and CO₂ compression.

The CO₂ avoidance cost was also estimated for these two retrofit cases. Two assumptions were made in this retrofit study. First, the capital cost of the existing power plant was assumed to be paid off. Second, a 20-year remaining life time was assumed for the retrofitted plant, and a factor of 0.138 was adopted for amortizing the new capitals. Therefore, the annual total cost is composed of the levelized new capital cost, the O&M costs of the LSD, SCR, and PC-MEA or PC-OC specific facilities, and the cost due to the electricity loss for CO₂ capture. The CO₂ avoidance cost is calculated assuming the Existing-PC_{ref} as reference plant.

Table 35 and Figure 10 displays the economics results for retrofit applications.

Table 35. Retrofit PC-OC and PC-MEA - Recap of costs

	CO ₂ capture plants	
	Retrofit PC-OC	Retrofit PC-MEA
Net Power Output	358 MW _e	350 MW _e
Total Plant Cost (TPC), \$/kW_e	595	665
Levelized capital costs, mills/kWh	14.58	16.97
Levelized O&M, mills/kWh	6.41	9.89
Cost of electricity loss, mills/kWh	20.16	21.80
Increase in COE, mills/kWh	41.15	48.66
Cost of CO₂ avoidance, \$/mt CO₂	48.44	63.51
Cost of CO ₂ capture, \$/mt CO ₂	35.20	42.21

Figure 10. Retrofit PC-OC and PC-MEA – COE and cost of CO₂ avoided

In the PC-MEA retrofit, a total of 151MW_e electricity output was lost, while in the PC-OC retrofit, 143MW_e was lost. The cost of the electricity loss was estimated based on the COE of the reference plant.

The CO₂ avoidance cost in the PC-MEA retrofit was about \$64/mt, while that in the PC-OC retrofit was about \$48/mt. These costs are larger than their counterparts in the new plant cases. One reason is the different reference plant used while calculating the CO₂ avoidance cost for New or Retrofit applications. For retrofit calculation, since the reference plant does not include neither LSD nor SCR, those units have to be added for PC-MEA applications (and thus add some extra cost accounted in the CO₂ avoidance cost). For the same reason, the retrofit PC-OC process does not account for the credit for not having an SCR, thus increasing the CO₂ avoidance cost. A second reason is that in the PC-OC retrofit case, a higher flue gas recycle ratio was used compared to the new plant case, resulting in higher mass flow, enlarged equipment and related cost increases.

5. SENSITIVITY ANALYSIS

The cost of CO₂ avoidance is affected by many performance parameters of CO₂ capture processes and economic assumptions used in the cost modeling. Due to the large auxiliary energy and capital cost required both in the PC-MEA process and the PC-OC process, the cost sensitivity analysis was only conducted for these related parameters.

For the PC-MEA process, two important parameters were studied: the requirement of the solvent regeneration heat and the capital cost of the absorption process. For the PC-OC process, the ASU electricity use and its capital cost are selected for this purpose.

Figure 11 displays the sensitivities of the CO₂ avoidance cost to the requirement of absorbent regeneration heat in the PC-MEA process and to the ASU electricity use in the PC-OC process. The following parameters and assumptions were used in the analysis:

- (1) 10%, 20%, 30%, 40%, 50% reductions for the absorbent regeneration heat;
- (2) 10%, 20%, 30%, 40%, 50% reductions for the ASU electricity use;
- (3) Other cost items were same as the baseline;
- (4) The plant output was 533 MW_e (gross).

Figure 11. Sensitivity of CO₂ avoidance cost to ASU or MEA energy need

The results revealed that the CO₂ avoidance cost significantly decreased with energy requirement improvement in both processes. The CO₂ avoidance cost for the PC-OC process decreased by about 33% when the energy requirement of the ASU was reduced by 50%. In the MEA process, a 50% reduction of the absorbent regeneration heat resulted in about 29% decrease of the CO₂ avoidance cost.

The sensitivity analysis of the CO₂ avoidance cost to the capital cost of the PC-MEA process and the capital cost of the ASU in the PC-OC process is shown in Figure 12. The following parameters and assumptions were used in the analysis:

- (1) -50%, -25%, 0%, 25%, 50% changes of the capital cost of the PC-MEA process;
- (2) -50%, -25%, 0%, 25%, 50% changes of the ASU capital cost in the PC-OC process;
- (3) The auxiliary power uses and other cost items were same as the baseline;
- (4) The plant output was 533 MW_e (gross).

Figure 12. Sensitivity of CO₂ avoidance cost to ASU or MEA capital costs

Figure 12 also shows the decrease in CO₂ avoidance cost with decrease of the capital cost. With a 50% reduction of the capital cost, the CO₂ avoidance cost decreased by 16% both for the PC-MEA process and for the PC-OC process. However, by comparing Figure 12 to Figure 11, the CO₂ avoidance cost is less sensitive to the capital costs of the PC-MEA process and the ASU in the PC-OC process than to the absorbent regeneration heat and ASU electricity use.

6. CONCLUSIONS

A techno-economic assessment of three 533MW_e gross coal-fired Power plants equipped with CO₂ capture capability was performed: an oxy-combustion pulverized-coal plant (PC-OC), an air-blown PC power plant with the MEA process (PC-MEA), and an IGCC plant with Selexol process (IGCC-S).

In all cases, the reference power plant is a 533MW_e gross (533MW_e net) air-fired PC power plant (PC_{ref}) without CO₂ capture system. PC_{ref} net generation efficiency is about 37% (sub-critical steam cycle). The flue gas treatment equipment of the PC_{ref} is slightly different for new plant or retrofit plant applications. In CO₂ capture plants, the CO₂ is compressed to 83 bara. The purification of CO₂ stream in the PC-OC process was not included in this study. To meet the gas specification potentially required for some specific CO₂ applications (such as EOR), additional purification processes might be necessary.

The following conclusions are obtained.

- (1) Both PC-OC and PC-MEA are suitable for retrofit applications in PC boilers. Compared to the conventional PC-MEA plant, an SCR may not be required in the PC-OC process (very low NO_x levels due to flue gas recycle process).
- (2) The volume of flue gas entering the LSD and ESP is comparable between the retrofit PC-OC, the PC-MEA and the PC_{ref} plant. The gas mass flow rate at a 60% recycle ratio in the new PC-OC plant is about 25% lower than the PC_{ref} plant. As a result, the costs related to the SO₂ and dust removal in the new PC-OC process are lower than in retrofit PC-OC.
- (3) A non-capture sub-critical air-fired PC_{ref} plants consumes about 6% of its 533MW_e gross power output as auxiliary power. A CO₂-capture PC-OC plant consumes about 32-33% of the same gross output, of which the ASU uses 19% of the total gross output, and the CO₂ compression about 8%. A CO₂-capture PC-MEA plant consumes about 34% of its gross power output, of which the MEA regeneration uses about 18%, and the CO₂ compression about 7%.
- (4) In retrofit applications to PC boilers, the net generation efficiency decreases from 37% without CO₂ capture, to 27.3% and 25.8% using retrofit PC-OC and PC-MEA respectively.
- (5) For new plants applications, the IGCC-S process leads to 36.2% net generation efficiency, compared to 27.9% and 25.8% using sub-critical PC-OC and PC-MEA, respectively. Most of new PC plants are built with higher efficiency supercritical steam cycles, that have not been considered in this study.
- (6) In retrofit applications, the increase in cost of electricity (COE) is about 41 mills/kWh using PC-OC, and 49 mills/kWh using PC-MEA. The cost of CO₂ avoided is about \$48/mt with PC-OC, \$64/mt with PC-MEA. The oxy-combustion process is economically favorable for CO₂ capture retrofit of existing PC power plants.

- (7) In new plant applications, and compared to non capture PC_{ref} at 50mills/kWh, the COE increases by 31mills/kWh using PC-OC system, by 40mills/kWh using PC-MEA and by 22 mills/kWh with IGCC-S. The CO_2 avoidance cost for a new sub-critical PC-OC plant is about \$36/mt, compared to \$52/mt for a new sub-critical PC-MEA plant, and to \$27/mt for IGCC-S process. Both PC-OC and PC-MEA processes would benefit from application to higher efficiency steam cycle (super and ultra supercritical steam cycles), with net generation efficiencies before CO_2 capture comparable or higher than IGCC efficiencies.
- (8) The sensitivity analysis showed that the CO_2 avoidance cost was more sensitive to the absorbent regeneration heat in the PC-MEA process and the electricity use of the ASU in the PC-OC process than their capital cost requirements. The CO_2 avoidance cost for the PC-OC process decreased by about 33% when the electricity use of the ASU reduced by 50%, while that for the PC-MEA process decreased by about 29% when the absorbent regeneration heat reduced by 50%.

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8. LIST OF ACRONYMS AND ABBREVIATIONS

BEC	Bare Erected Cost
Cap	Capture Power Plant
COE	Cost of Electricity
EOR	Enhanced Oil Recovery
ESP	Electro-Static Precipitator
FOM	Fixed Operating & Maintenance costs
GHG	Green House Gases
IEA	International Energy Agency
IGCC	Integrated Gasification Combined Cycle
IGCC-S	IGCC equipped with a Selexol unit
LP	Low Pressure
LSD	Lime Spray Dryer
10 ⁶ Btu/h	Million Btu/hour
Mills	1,000 mills = \$1
mt or metric t	metric ton
MEA	Mono-Ethanol-Amine
OC	Oxy-Combustion
O&M	Operating and Maintenance
PC	Pulverized Coal
PC-MEA	PC Boiler equipped with MEA unit
PC-OC	Oxy-Combustion in PC boilers
PCref	Reference air-blown PC plant without CO ₂ capture
PRB	Powder River Basin
Ref	Reference Power Plant
t or ton	short ton
TCR	Total Capital Requirement
TPC	Total Plant Cost
TPD	(metric) Ton per Day
TPI	Total Plant Investment
SCR	Selective Catalytic Reduction
SI	Système International d'Unité
Syn-gas	Synthesis gas (H ₂ /CO)
VOM	Variable Operating & Maintenance costs

9. APPENDICES

(Appendices available in a separate document)

Appendix 1.

Detailed tables resulting from performance simulation and cost assessment of the PC_{ref}, PC-OC and PC-MEA processes.

Appendix 2.

Detailed tables resulting from performance simulation and cost assessment of the IGCC-S process.

Advanced, Low/Zero Emission Boiler Design and Operation

TOPICAL REPORT – TASK 3

Performance Simulation and Economics Assessment of three CO₂ capture coal-fired power plants:

- Sub-critical pulverized coal power plants equipped with Oxy-combustion (New and Retrofit Plants)
- Sub-critical pulverized coal power plants equipped with Post-combustion (New and Retrofit Plants)
- IGCC Power plant equipped with Selexol scrubbing (New Plant)

APPENDICES

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Appendix 1. Detailed tables resulting from performance simulation and cost assessment of the PC_{ref}, PC-OC and PC-MEA processes. A3

Appendix 2. Detailed tables resulting from performance simulation and cost assessment of the IGCC-S process. A10

Appendix 1. Detailed tables resulting from performance simulation and cost assessment of the PC_{ref}, PC-OC and PC-MEA processes.

Table 1 : PC_{ref}, PC-MEA and PC-OC - Main Process Parameters

	New PC _{ref} *	New PC-MEA	New PC-OC	Retrofit PC-OC
Combustion				
Excess Air, /	15%	15%	3%	3%
Air flow rate, lb/h	4,029,474	4,029,474	n/a	
O ₂ /Ar flow rate, lb/h	n/a	n/a	813,423	843,885
Coal feed rate, lb/h	524,982	524,982	504,064	522,941
Flue gas recycle ratio, lb/h	n/a	n/a	0.72	0.60
Steam generation				
Hot reheat steam, lb/h	3,022,125	3,022,125	3,022,125	3,022,125
Superheat steam, lb/h	3,422,824	3,422,824	3,422,824	3,422,824
IP steam to MEA process	n/a	1,470,914	n/a	n/a
Steam condensate, lb/h	2,802,051	1,331,136	2,802,051	2,802,051
Main feed water, lb/h	3,321,228	3,321,228	3,321,228	3,321,228
Heat duty of cooling tower, 10 ⁶ Btu/h	2,178	1,034	2,178	2,178
Flue gas from boiler				
Flue gas mass flow rate, lb/h	4,526,829	4,526,829	3,366,231	4,902,386
Flue gas temperature, °F	295	295	295	295
Composition:				
N ₂ , vol%	2.51%	2.51%	2.11%	2.02%
O ₂ , vol%	71.65%	71.65%	0.32%	0.30%
Ar, vol%	0.10%	0.11%	0.60%	0.58%
CO ₂ , vol%	14.55%	14.55%	51.26%	48.92%
H ₂ O, vol%	11.15%	11.15%	45.67%	48.16%
SO ₂ , ppm	257 ppm	257 ppm	383 ppm	263 ppm
NO _x , ppm	117 ppm	117 ppm	73 ppm	62 ppm
Fly ash flow rate, lb/h	22,050	22,050	21,171	21,359

* The process performances of a new and a retrofit PC plant are similar except for the NO_x concentration. The 117 ppm NO_x concentration is for the new PC, and is about 163 ppm for the retrofit.

Table 2: PC_{ref}, PC-MEA and PC-OC - Costs of the Basic Power Plant Components

	New PC _{ref}	New PC-MEA	New PC-OC	Retrofit PC-OC
TPC, k\$				
1. Coal & sorbent handling	28,007	28,007	27,199	27,373
2. Coal & sorbent prep & feed	22,250	22,250	21,608	21,747
3. Feed water & misc. 80 P systems	35,854	35,854	35,854	35,854
4.1 PC boiler & accessories	95,121	95,121	95,121	95,121
4.4 boiler Bop	6,198	6,198	6,198	6,198
5. ESP	22,500	22,500	18,513	24,265
6. HRSG, ducting and Stack	25,286	25,286	20,429	26,779
7. Steam turbine generator	79,168	79,168	79,168	79,168
8. Cooling water system	28,778	16,841	28,778	28,778
9 Ash/Spent sorbent handling system	20,937	20,937	20,333	20,463
10 Accessory electric plant	30,019	30,019	30,019	30,019
11 Instrumentation & control	16,826	16,826	16,826	16,826
12 Improvements to site	11,555	11,555	11,555	11,555
13 Buildings and structures	56,854	56,854	56,854	56,854
subtotal	479,351	467,414	468,454	480,999
O&M cost, k\$/yr				
1. FOM				
Operation labor	8,760	8,760	8,760	8,760
Maintenance labor	4,451	4,389	4,335	4,467
Maintenance material	6,677	6,584	6,503	6,700
Administration support labor	3,963	3,945	3,929	3,968
subtotal	23,851	23,677	23,527	23,894
2. VOM				
Water making up	3,040	3,040	3,040	3,040
Chemicals (water treatment)	1,624	1,624	1,624	1,624
Ash disposal	845	845	811	819
Coal cost	32,192	32,192	30,909	31,185
subtotal	37,702	37,702	36,385	36,668

Table 3: PC-OC - Costs of Air Separation Unit (ASU) and CO₂ Compression

	New PC-OC	Retrofit PC-OC
TPC, k\$		
(1) ASU	120,839	121,915
(2) Flue gas recycle line	2,729	2,857
(3) PC-OC Condenser	3,957	4,270
(5) CO ₂ Compressor	32,716	32,904
subtotal	160,241	161,946
O&M cost, k\$/yr		
1. FOM		
(1) Operating Labor	1,402	1,402
(2) Maintenance labor & materials	4,006	4,049
(3) Administration & support labor	901	906
subtotal	6,309	6,357
2. VOM		
(1) Water used in PC-OC condenser	3,721	3,807

Table 4: PC_{ref}, PC-MEA and PC-OC - Costs of the Lime Spray Dryer (LSD)

	New PC _{ref}	New PC-MEA	New PC-OC	Retrofit PC-OC
TPC, k\$				
(1) Reagent feed equipment	10,534	10,534	10,572	10,620
(2) SO ₂ removal equipment	19,856	19,856	16,623	18,031
(3) Flue gas handling equipment	7,061	7,061	6,894	8,759
(4) Waste handling equipment	1,471	1,471	1,483	1,469
(5) Support equipment	11,182	11,182	10,377	11,984
subtotal	50,104	50,104	45,948	50,863
O&M cost, k\$/yr				
Fixed O&M				
(1) Operating Labor	1,277	1,277	1,106	1,322
(2) Maintenance labor& materials	705	705	647	716
(3) Administration& support labor	468	468	409	482
subtotal	2,450	2,450	2,162	2,520
Variable O&M				
(1) Lime reagent	861	861	875	893
(2) Waste disposal	471	471	478	488
(3) Fresh water	1	1	1	1
subtotal	1,333	1,333	1,354	1,382

Table 5: PC_{ref} and PC-MEA - Costs of the Selective Catalytic Reduction (SCR)

	New PC_{ref}	New PC-MEA
TPC, k\$	29,776	29,776
O&M cost, k\$/yr		
1. Fixed O&M	197	197
2. Variable O&M		
(1) NH_3 use	202	202
(2) Catalyst replace	449	449
subtotal	651	651

Table 6: PC-MEA - Costs of the Mono-Ethanol-Amine (MEA)

	New PC-MEA
TPC, k\$	152,384
O&M cost, k\$/yr	
1. FOM	
(1) Operating Labor	701
(2) Maintenance labor& materials	3,810
(3) Administration& support labor	667
subtotal	5,178
2. VOM	
(1) MEA make-up	6,356
(2) Inhibitor	1,271
(3) Reagent Na caustics	643
(4) Reagent AC	371
(5) Waste disposal	61
(6) Water	2,622
subtotal	11,324

Table 7: PC_{ref} , PC-MEA and PC-OC - Power Generation and Auxiliary Use

	New PC_{ref}	New PC-MEA	New PC-OC	Retrofit PC-OC
Gross output, kW _e	533,230	434,850	533,230	533,230
Auxiliary load summary, kW_e				
Coal handling	339	339	326	329
Pulverizers	2,937	2,937	2,820	2,845
Primary air fans	1,212	1,212	959	1,530
Forced Draft fans	1,154	1,154	912	1,456
Induced draft fans	5,122	5,122	1,461	1,544
Seal air blowers	46	46	36	57
Steam turbine auxiliaries	884	884	884	884
Condensate pumps	949	891	949	949
* Main feed pump	10,938	10,938	10,938	10,938
Circulating water pumps	4,187	1,989	4,187	4,187
Cooling tower fans	2,367	1,125	2,367	2,367
Ash handling	1,424	1,424	1,367	1,379
Miscellaneous	2,411	2,411	2,411	2,411
Transformer loss	1,215	1,215	1,215	1,215
ESP	1,319	1,319	1,006	1,465
LSD	3,500	3,500	3,201	3,597
SCR	2,750	2,750		
MEA: gas induced fan		15,837		
MEA pump		2,980		
CO ₂ Compressor		37,951	44,313	44,710
PC-OC flue gas recycle			1,651	2,933
ASU			100,572	101,468
Total	31,816	85,086	170,637	175,326

* Feed-water pumps are turbine-driven, and not included in the subtotal auxiliary use.

Table 8: Air-Blown PC_{ref} , PC-MEA and PC-OC - Overall Process Performances

	New PC_{ref}	New PC-MEA	New PC-OC	Retrofit PC-OC
Coal Flow Rate (lb/hr)	524,982	524,982	504,064	508,554
Steam Turbine Power (MW _e)	533.2	434.9	533.2	533.2
Boiler efficiency (%)	86.4%	86.4%	90.0%	89.2%
ASU Power (MW _e)			100.57	101.47
Other Aux. Power (MW _e)	31.8	85.1	70.1	73.9
Net Power (MW _e)	501.4	349.8	362.6	357.9
Net efficiency, HHV (%)	37.0%	25.8%	27.9%	27.3%
Net heat rate, HHV (Btu/kWh)	9,214	13,208	12,233	12,504
CO ₂ Removal, k tonne/year	0	2,472	2,517	2,539

*Table 9: PC_{ref} , PC-MEA and PC-OC - Costs of Electricity and of CO₂ avoidance/capture
(New Power Plants)*

	New PC_{ref}		New PC-MEA		New PC-OC	
Capital costs	k\$	\$/kW	k\$	\$/kW	k\$	\$/kW
Total Plant Cost (TPC)	559,231	1,115	699,678	2,000	674,643	1,861
Capital investment						
Total cash expended	559,231		699,678		674,643	
AFDC	50,419		63,082		60,824	
Total Plant investment (TPI)	609,650	1,216	762,760	2,181	735,468	2,028
Royal allowance						
Preproduction costs	15,971	32	20,393	58	19,162	53
Inventory capital	12,116	24	15,477	44	13,110	36
Initial catalyst & chemicals						
Land cost	2,080	4	2,080	6	2,101	6
Total capital requirement (TCR)	639,817	1,276	800,710	2,289	769,841	2,123
O & M costs	k\$/yr	mills/kWh	k\$/yr	mills/kWh	k\$/yr	mills/kWh
Fixed O & M	17,223	5.60	20476	9.55	20,799	9.35
Variable O & M	9,274	3.02	11026	5.14	11,199	5.04
Consumables	7,494	2.44	18818	8.77	10,551	4.75
Fuel cost	32,192	10.47	32192	15.01	30,909	13.90
	\$/kW-yr	mills/kWh	\$/kW-yr	mills/kWh	\$/kW-yr	mills/kWh
Levelized capital costs	176	28.72	316	51.52	293	47.78
Levelized cost of electricity		50.24		89.99		80.82
Cost of CO₂ avoidance, \$/mt				51.88		35.95
Cost of CO₂ capture, \$/mt				34.45		26.73

Table 10: PC-MEA and PC-OC - Costs of Electricity and of CO₂ avoidance/capture (retrofit Power Plants)

	New PC-MEA		New PC-OC	
Net output, MW _e	350		358	
Power generation, GWh	2,145		2,195	
Total Plant Cost	k\$	\$/kW_e	k\$	\$/kW_e
ASU & OC & compressor	-	-	161,946	452
MEA (inclgd compressor)	152,384	436	-	-
SCR	29,947	86	-	-
LSD	50,104	143	50,863	142
Total	232,436	665	212,809	595
O&M cost	k\$/y	mills/kWh	k\$/y	mills/kWh
1. FOM				
ASU & OC & compressor			6,357	2.90
MEA (inclgd compressor)	5,178	2.41		
SCR	198	0.09		
LSD	2,450	1.14	2,520	1.15
subtotal	7,825	3.65	8,876	4.04
2. VOM				
ASU & OC & compressor	-	-	3,807	1.73
MEA (inclgd compressor)	11,325	5.28	-	-
SCR	739	0.34	-	-
LSD	1,333	0.62	1,382	0.63
subtotal	13,397	6.25	5,189	2.36
Total O&M cost	21,222	9.89	14,065	6.41
	k\$/y	mills/kWh	k\$/y	mills/kWh
Cost due to electricity loss	46,757	21.80	44,247	20.16
Increase of COE, mills/kWh		48.66		41.15
Cost of CO₂ avoidance, \$/mt		63.51		48.44
Cost of CO₂ capture, \$/mt		42.21		35.20

Appendix 2. Detailed tables resulting from performance simulation and cost assessment of the IGCC-S process.

Table 11: IGCC-S Power Plant - Process Performance

GROSS POWER SUMMARY, kW_e	
Gas Turbine Power	358,541
Steam Turbine Power	174,601
Generator Loss	-7,996
Turbo-Set Power	525,146
Fuel Gas Expander Power	10,611
Gross Plant Power	535,757
AUXILIARY LOAD SUMMARY, kW_e	
Coal Handling and Conveying	526
Coal Milling	1,213
Coal Slurry Pumps	235
Slag Handling and Dewatering	123
Recycle Gas Blower	127
Air Separation unit (ASU)	36,759
Oxygen Boost Compressor	16,259
Selexol Plant	7,977
Claus/TGTU	15
Tail Gas Recycle	154
Humidification Tower Pump	65
Humidifier Makeup Pump	141
Low Pressure CO ₂ Compressor	865
High Pressure CO ₂ Compressor	25,782
Condensate Pumps	329
High Pressure Boiler Feed Pump	3,294
Low Pressure Boiler Feed Pump	61
Miscellaneous Balance of Plant	1,092
Gas Turbine Auxiliaries	623
Steam Turbine Auxiliaries	244
Circulating Water Pumps	1,636
Cooling Tower Fans	1,028
Flash Bottoms Pump	55
Transformer Loss	1,661
Total Auxiliary Power Requirement	100,263
NET PLANT POWER, kW_e	435,493
PLANT EFFICIENCY	
Net Efficiency, % (HHV)	36.18 %
Net Heat Rate, Btu/kWh (HHV)	9,422
CONSUMABLES	
As-Received Coal Feed, lb/h	466,262
Oxygen (95% pure), lb/h	288,533
Water, lb/h	584,349

Table 12: IGCC-S Power Plant - Capital Cost

Gross output, MW_e	536 MW _e
PLANT SECTION DESCRIPTION, k\$	
Coal handle	19,101
Coal Slurry Preparation	31,997
Feed water	14,345
Oxygen Plant	57,218
Destec Gasifier	163,378
Gas Cleanup & Piping	77,567
CO2 Compression	55,649
Gas Turbine System	66,249
HRSG	21,767
/Steam Turbine	25,144
Cooling water tower	13,901
Ash	9,092
Electrical	42,905
Control/Instrumentation	13,944
Site improvement	8,999
Civil/Structural/Architectural	7,860
Subtotal, Process Plant Cost, k\$	629,117
Engineering Fees, k\$	62,912
Process Contingency, k\$	15,360
Project Contingency, k\$	94,367
Total Plant Cost (TPC) , k\$	801,755
TPC \$/kW_e	1,841
Total cash expended, k\$	801,755
Allowance for funds used during construction, k\$	111,638
Total Plant Investment (TPI) , k\$	913,393
Prepaid Royalties, k\$	3,146
Initial Catalyst and Chemical Inventory, k\$	194
Startup Costs, k\$	21,731
Spare Parts, k\$	4,009
Working Capital, k\$	8,493
Land, k\$	1,300
Total Capital Requirement (TCR) , k\$	952,265
TCR, \$/kW_e	2,187

Table 13: IGCC-S Power Plant - Annual O&M Costs

Gross output, MW_e	536 MW_e
Operating Cost, k\$/yr	
Coal (PRB)	28,591
Consumable Materials	
Water	340
Selexol Solvent	1,458
Claus Catalyst	0
SCOT Activated Alumina	1
SCOT Cobalt Catalyst	1
SCOT Chemicals	3
Ash/Sorbent Disposal Costs	1,032
Plant labor	
Operating Labor (including benefit)	5,957
Supervision & Clerical (30% of O&M cost)	3,904
Maintenance Costs	17,639
Royalties (1% of fuel cost)	286
Other Operating Costs	1,301
Subtotal	60,511
By-Product Credits, k\$/yr	
Sulfur product	244
Net Operating Costs, k\$/yr	60,267

Table 14: IGCC-S Power Plant - Costs of Electricity Generation and of CO₂ avoidance/capture

Gross output, MW_e	536 MW_e	
	\$/kW_e	mill/kWh
Total Capital Requirement (TCR)	2,187	
Annual carry-on charges	302	49.21
	k\$/yr	mill/kWh
O&M cost	60,267	22.57
Levelized cost		
Cost of electricity, mills/kWh		71.78
Cost of CO ₂ avoidance, \$/mt		26.51
Cost of CO ₂ capture, \$/mt		24.22

Advanced, Low/Zero Emission Boiler Design and Operation

TOPICAL REPORT

TASK 6: Engineering, Plant Design and Cost assessment for retrofit application

Work Performed Under Contract No.: **DE-FC26-02NT41586**

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ABSTRACT

The key to a strong, stable, and secure economy is readily available, reasonably priced energy. To maintain economic competitiveness and meet growing energy demands, the U.S. must improve the utilization of its domestic resources of which coal is among the largest. Coal currently provides more than 50% of the electricity generated in the United States -- coal will continue to be a primary source for electric power generation into the future.

Fossil fuel combustion is a primary contributor of increased Greenhouse Gas (GHG) emissions. As fossil fuels continue to be the dominant fuel source for electricity generation, reducing carbon emissions by capturing and sequestering CO₂ from utility boilers could be required to reduce the buildup of GHG in the atmosphere. Predictions of future global energy use suggest a continued increase in carbon emissions and rising concentrations of CO₂ in the atmosphere unless major changes are made in the way we produce and use energy - in particular, how we manage carbon. But, before carbon can be managed, it must be concentrated and captured.

To respond to this challenge and to keep coal a primary source for electric power generation, The Babcock & Wilcox Company (B&W) in partnership with Air Liquide (AL), have developed a multi-emissions control, oxy-combustion technology with flue gas recycle to produce a concentrated CO₂ stream that could be prepared for sequestration. The technology will be applicable to both new and existing coal-fired plants. B&W and AL proposed to demonstrate this technology in a two-phase program. During Phase 1, the process configuration was established, heat and mass balances for various load conditions were made, all equipment was designed, and the total cost for Phase 2 was determined. Following successful completion of Phase 1 goals and securing of funding, the Phase 2 oxy-combustion demonstration was to take place at the 25 MWe pulverized coal-fired power plant unit located at the City of Hamilton Municipal Electric Power Plant in Hamilton, Ohio. Phase 2 would consist of final engineering, fabrication, installation of the equipment, startup, operation, testing, and site restoration.

The field demonstration project goals are to: 1) transition the plant to full oxygen with flue gas recycle operation, 2) operate at full load, 3) obtain data and experience at full load, low loads, and load changing, 4) obtain heat transfer data and slagging and fouling information, 5) obtain wet scrubber performance data and determine impact on the scrubber processes, and 6) obtain significant reduction in NO_x, SO_x, Hg, and particulate in the exit stream which will continue beyond the demonstration project and permit the City of Hamilton to burn high-sulfur Ohio coal. Once the demonstration is completed, this technology will provide the utility with the option of continuing to burn Ohio coal.

Phase 1 began in October of 2005 and was completed in June 2006 co-funded by the Department of Energy, Air Liquide, and B&W. The objective of Phase 1 was to design the conversion of the City of Hamilton's 25MWe Unit 8 steam generating system into an oxycombustion process capable of full load and to develop the cost and schedule to fabricate construct and test the facility in Phase 2. This report summarizes that effort.

ACRONYMS & ABBREVIATIONS

A	AAL	American Air Liquide
	A/C	Air Conditioning
	acfm	Actual cubic feet per minute
	AH	Air heater
	AL	Air Liquide
	AQCS	Air Quality Control System
	Ar	Argon
	ASTM	American Society for Testing and Materials
	ASU	Air Separation Unit
	Avg	Average
B	Btu	British Thermal Units
	B&W	Babcock & Wilcox
	Blr	Boiler
	BOP	Balance of Plant
	BWRC	Babcock & Wilcox Research Center
C	C	Carbon
	CANMET	Canada Centre for Mineral and Energy Technology
	Ca/S	Molecular Ratio of Calcium to Sulfur
	CaCO ₃	Calcium Carbonate
	CaSO ₃	Calcium Sulfite
	CaSO ₄	Calcium Sulfate
	CCPC	Canadian Clean Power Coalition
	CEDF	Clean Environment Development Facility
	CEMs	Continuous Emissions Monitoring System
	CO	Carbon Monoxide
	COE	Cost of Electricity
	CoH	City of Hamilton, Hamilton, Ohio
	CO ₂	Carbon Dioxide
	CP	Convection Pass
	CPS	City Public Service
	CVAAS	Cold Vapor Atomic Absorption Spectroscopy
D	DCS	Digital Control System
	DOE	Department of Energy
E	ECO	Economizer
	EIO	Energy Industries of Ohio
	Ent.	Entering
	EPA	Environmental Protection Agency

	EPRI	Electric Power Research Institute
	ESP	Electrostatic Precipitator
F	F	Fahrenheit
	FD	Forced Draft (fan)
	FEGT	Furnace Exit Gas Temperature
	FGD	Flue Gas Desulfurization
	FRP	Fiber Reinforced Plastic
	ft.	Feet
	ft/sec	Feet per Second (velocity)
G	Gal	Gallons
	GCCI	Global Climate Change Initiative
	GDP	Gross Domestic Product
	GHG	Greenhouse gases
	Gpm	Gallons per Minute
	GW	Giga Watts
H	H	Hours
	H	Hydrogen
	HESP	Hot Electrostatic Precipitator
	Hg	Mercury
	HHV	Higher Heating Value of Fuel
	H & M	Heat and Mass
	Hp	Horse Power
	HP	High Pressure (steam)
	HVT	High Velocity Thermocouple
	H ₂ O	Water
	H ₂ SO ₄	Sulfuric Acid
I	ID	Induced Draft (fan)
	IEEE	Institute of Electrical and Electronic Engineers
	IGCC	Integrated Gasification Combined Cycle
	in. H ₂ O	Inches of water pressure
	in. wg	Inches of water gage pressure
	in.	Inches
K	klb/day	Kilo (1000) pounds per Day
	klb/h	Kilo (1000) pounds per Hour
	kW	Kilowatt
	kWh	Kilowatt Hour
L	Lb	Pounds

	lb/h	Pounds per hour
	L/G	Liquid to Gas Ratio
	LOX	Liquid Oxygen
	LP	Low Pressure
	LPA	Lewis Perry Associates
	Lvg	Leaving
M	M	Mega – million (10 ⁶)
	MBtu	Million (Mega) British Thermal Units
	MCA	Multi Contact Absorber
	MEA	Monoethanolamine
	MgCO ₃	Magnesium Carbonate
	MHVT	Multiple High Velocity Thermocouples
	MCC	Motor Control Centre
	MW	Mega Watts
	MWe	Million (Mega) electrical watts
	MWh	Megawatt hour
N	N/A	Not Applicable
	NEIL	National Energy Technology Laboratory
	NFPA	National Fire Protection Association
	NO _x	Oxides of Nitrogen (primarily NO ₂)
	N ₂	Nitrogen
	NPSH	Net Positive Suction Head
O	OAQA	Ohio Air Quality Authority
	OCDO	Ohio Coal Development Office
	OCTF	Oxygen Combustion Test Facility
	ODOD	Ohio Department of Development
	ODS	Oxygen Distribution System
	OFA	Overfire Air
	O&M	Operate and Maintain
	OSHA	Occupational Health and Safety Act
	OSS	Oxygen Supply System
	O ₂	Oxygen
P	PA	Primary Air
	PC	Pulverized Coal
	pH	Measure of Acidity or Alkalinity
	PM	Particulate Matter
	Ppm	Parts per million
	ppmdv	Parts per Million Dry Volume Basis

	psi	Pounds per Square Inch
	Psig	Pounds per Square Inch Gage Pressure
	PTC	Power Test Code
	P&ID	Process and Instrument Diagram
	PO	Primary Oxidant
	PRB	Powder River Basin
Q	Qty	Quantity
R	R&D	Research & Development
	Req'd	Required
	RFQ	Request for Quote
	Rpm	Revolutions per minute
S	S	Sulfur
	SBS	Small Boiler Simulator
	SCAH	Steam Coil Airheater
	SCPC	Supercritical Pulverized Coal
	SCR	Selective Catalytic Reduction
	Sec. SH	Secondary Superheater
	SO	Secondary Oxidant
	SO ₂	Sulfur Dioxide
	SO ₃	Sulfur Trioxide
	SO _x	Oxides of sulfur
T	TA	Tertiary Air
	TAC	Technical Advisory Committee
	TAH	Tubular Airheater
	TI	Titanium
	TO	Tertiary Oxidant
U	U.S.	United States
	USC	Ultra Supercritical
V	VAC	Volts Alternating Current
W	WB	Windbox
	Wt	Weight
	WFGD	Wet Flue Gas Desulfurization
	WWT	Waste Water Treatment
	WWTP	Waste Water Treatment Plant
Y	Yr	Year

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1.0 EXECUTIVE SUMMARY

On February 14, 2002, President Bush announced the Global Climate Change Initiative (GCCCI) with the goal of significantly reducing the greenhouse gas intensity of the United States Economy over the next 10 years, while sustaining the economic growth needed to finance investment in new, clean energy technologies. The GCCCI calls for increased investment in research and development to provide an improved basis for sound future decisions, for increased emphasis on capture and sequestration of CO₂, and for reductions in non-CO₂ greenhouse gas emissions. The programmatic timeline is to demonstrate a portfolio of safe, cost effective greenhouse gas capture, storage and mitigation technologies at commercial scale by 2012.

It is well known that about 50% of the electricity generated in the U.S. is from coal fired power plants while in Ohio it is nearly 90%. The abundance of domestic coal compared to other available fuels is also well documented. In the interest of energy security and lowest cost for power production, coal is clearly indispensable for the foreseeable future. Essentially all of the existing 322 GW of coal fueled generating capacity employs combustion and it is unrealistic to expect to implement any carbon management technology that requires rapid and wholesale replacement of the existing fleet, therefore a technology that can be retrofit to the existing fleet is of utmost importance, at least in the near term, while also looking at future plant designs will insure its acceptance for next generation systems. One such technology being developed is oxygen combustion with flue gas recycle.

This technology provides a concentrated CO₂ stream which can be prepared for sequestration or sold as a byproduct for enhanced oil recovery. In addition, it offers reductions in SO₂, NO_x, particulate, and mercury emissions from a coal-fired power plant.

In October of 2005, B&W in partnership with AL and with the cooperation of the City of Hamilton (CoH), Ohio, embarked on Phase 1 of a two phase project to design, retrofit and test the first power plant in the United States with oxygen combustion and flue gas recycle for concentrating CO₂ that could be prepared for permanent storage (sequestration). The total project includes the design, fabrication, installation, commissioning, and testing of the retrofit conversion of the 25 MWe Unit 8 at the City's municipal power plant for oxygen combustion with flue gas recycle. Due to the unit size, the cost of installing CO₂ purification and compression equipment, lack of a market for the CO₂, and the lack of a local storage reservoir, it was impractical to apply the compression and sequestration step in this project. In addition, due to cost, a liquid oxygen system was selected rather than a conventional cryogenic air separation unit as would be used at a commercial installation. However, all combustion and operational aspects of transitioning from air firing to oxygen with recycled flue gas will be tested, heat transfer will be verified, flame stability and load changing will be proven and the effectiveness of the process to concentrate CO₂ will be evaluated taking into account the air in-leakage of the plant due to its age. To minimize this impact the gas tightness of the existing equipment will be repaired to greatest extent practical and measured.

Phase 1 was funded by the Department of Energy, B&W and AL. The objectives of Phase 1 were, 1) to develop a viable design for a large-scale pilot retrofit design that will demonstrate

that oxygen combustion for concentrating CO₂ can be applied to an existing power plant as well as a new plant, 2) develop the primary control philosophies that will allow it to be operated in a similar manner to conventional air firing, 3) develop a basic test plan, 4) determine the overall cost for a large-scale pilot retrofit including the cost of equipment, installation, and commissioning followed by three (3) months of testing.

This phase of the project has been completed and is described in detail in this final report. The results of the study show that retrofitting Unit 8 is quite feasible but more costly than initially expected. The very early estimate based on a 'rule of thumb' cost for the wet scrubber systems and with out an understanding of the balance of plant impact was about \$30M, \$15M for oxygen supply and \$15M for the remaining equipment and testing. The final costs totaled \$55.88 M but were later reduced to \$51.58M after cost reductions (without eliminating any equipment). The need to add a full wet scrubber system to prevent undesirable concentration of SO₂ and SO₃ in the recycle loop, cool and dehumidify the recycle gas represents \$28.4 M, oxygen supply is \$13.3 M, testing and reporting is \$3.4M and the remaining \$6.48 M is flues, dampers, burner modifications, boiler modifications, control and instrumentation.

Although the large portion for the wet scrubber systems appears to be less important to the overall objectives, flue gas recycle burning a high sulfur coal with over 3% sulfur would not be possible without SO₂ removal not only to protect the equipment from corrosion but to meet stack compliance. Even if operated using their current compliance coal with less than 1% sulfur, the recycle system would experience concentrations equivalent to a 3.5% sulfur coal, though stack compliance could be met. In either case, moisture removal and flue gas cooling to permit use of the existing fans would be necessary. This function is integrated into the absorber in a unique manner.

2.0 INTRODUCTION

Unit 8 at City of Hamilton Ohio is a 25 MWe unit that is being converted for an oxygen-fired demonstration. A wet flue gas desulfurization (WFGD) system is being installed as a part of this project. One of the results of using oxygen rather than air is that the mass input to the combustion process is greatly reduced. Because of this, flue gas is recycled back to the boiler in sufficient quantity to increase the mass flow for heat transfer in the convection pass and to lower the adiabatic flame temperature in the boiler. The lack of nitrogen in the combustion stream increases the concentration of SO_2 , SO_3 , NO_x , H_2O , and Hg in the recycle stream and boiler to about 3.5 times their concentrations with conventional air-firing. Considerations for multi-pollutant removal are unique as is the question of whether that equipment should be located in the higher volume main flue gas recycle stream, in the product stream, or in the CO_2 production system. This paper describes those considerations and how they have been applied for the City of Hamilton Oxygen Combustion Demonstration Project.

Why Oxygen Combustion?

The case favoring the use of coal as the primary fuel for power generation in the U.S. is overwhelming. Coal is the most abundant domestic fuel and it remains the lowest cost fuel for power generation. A direct relationship has also been shown between electricity cost and the Gross Domestic Product (GDP) per capita (Figure 2.1). Current margins between electricity production and demand are at their lowest levels and significant economic growth is projected resulting in a steady increase in the use of coal as Figure 2.2 shows. Though the severity of the impact of carbon dioxide (CO_2) on global warming is still under debate, coal is unfortunately the most carbon intensive fuel, making an economical means of carbon management essential to ensure coal's continued use.

Figure 2.1 - Electricity Consumption vs. GDP

Coal is an abundant domestic resource and over half of the electrical power generation in the U.S. is coal-fired. In order to reduce reliance on imported oil and natural gas, and improve U.S. national energy security and economy, coal should also be used to produce synthetic natural gas

and liquid fuels for transportation. These are areas in which gasification can make a major impact. Although use of Integrated Gasification Combined Cycle (IGCC) to produce electricity has gained interest and popularity recently, pulverized coal combustion has been, and is expected to continue to be the workhorse technology for power generation in the U.S. and economical solutions to the emissions issues are rapidly becoming more available.

Figure 2.2 - U.S. Electricity Demand 1970- 2020

As Figure 2.3 shows, the emissions from pulverized coal-fired power plants have been reduced tremendously over the past three decades, driven primarily by technology and regulations, but also as a result of public concern and industry recognition of the need. Great strides have been made in SO_2 and NO_x reduction through scrubbing and selective catalytic reduction technologies. Fabric filters and improvements in electrostatic precipitators have reduced particulate emissions and more recently technologies such as wet electrostatic precipitators are capable of further reductions including $\text{PM}_{2.5}$.

Figure 2.3 - Coal-Fired Power Plant Emissions Improvement

With technologies available to address the primary pollutants and major programs to retrofit the existing fleet in progress, public and industry attention then turned to mercury. Though regulated only in a few states, economical solutions for mercury are now on the verge of

commercial availability for both eastern and western coals. With concerns about global warming intensifying and public interest being amplified by recent unusual natural events, development of ways to address carbon dioxide emissions has taken the forefront.

There are several promising technologies to address the capture of CO₂ from use of fossil fuels and all are depend upon development of a safe means of disposal. Assuming disposal technologies can be commercialized, the problem for coal combustion processes becomes one of extracting the CO₂ from the exhaust gas. A modern coal-fired plant with a heat rate just under 9000 Btu/kWh will produce about 1,800 lb of CO₂ per MWh. Unfortunately, it is diluted in the exhaust gas at about 15% by volume wet so it is difficult to extract.

Three approaches are presently seen as the front runners: oxygen combustion, amine scrubbing for new or existing plants that burn coal and IGCC for new plants that choose gasification. Oxygen combustion and IGCC concentrate the CO₂ for disposal where amine scrubbing extracts the CO₂ from the flue gas using a regenerable solvent-catalyst such as monoethanolamine or (MEA). Some current studies show oxygen combustion as the least costly while others lean toward IGCC, indicating that the jury is still out. None of the technologies have been demonstrated at significant size in an integrated full-scale system.

To concentrate CO₂, IGCC depends on a water shift reaction and a large gas turbine that burns hydrogen. These components were not demonstrated in IGCC plants built in the late 1980s, nor do they appear to be included in the plant designs currently being developed. Unfortunately, the principles have only been applied at small scale, and large scale hydrogen production and its combustion in a large gas turbine are yet to be demonstrated in an integrated plant. In fact, past attempts at full integration in Europe experienced problems yet to be resolved.

Amine scrubbing has not been demonstrated at utility plant scale and the various constituents in boiler flue gasses present challenging problems for the sorbents.

Only oxygen combustion is based on equipment and systems that are already commercially available at the required scale. Though there are some operational issues to be proven at large scale, oxygen combustion and the major operational processes have been demonstrated at pilot scale and a demonstration of the technology in a full-scale new unit has been suggested.

At this time, its greatest challenge is to reduce the capital and power cost associated with the oxygen supply and the CO₂ compression. Recent work to better integrate the cryogenic process with the power island needs and conditions has led to some cost reductions and additional opportunities exist. In addition, other methods of extracting oxygen are being developed that could greatly reduce the cost.

In spite of these significant cost penalties, recent studies show oxygen combustion as being quite competitive with the other technologies and, since it is composed of conventional equipment, likely to have a considerably lower operational risk. Figures 2.4 and 2.5 show results of recent studies by the U.S. Department of Energy, one with B&W and Air Liquide¹ and the other with Seltzer, A.H. et al. of Foster Wheeler², both indicating oxygen combustion will be the lowest cost solution for coal. SC represents current supercritical steam cycles with

steam conditions of around 3600 psi, 1100F, 1100F and ultra supercritical (USC) steam conditions are on the order of 4000 psi, 1300F, 1300F.

Figure 2.4 - Impact of CO₂ Capture on Plant Cost Relative to Air-Fired Plant (Ref. 1)

Figure 2.5 - Comparison of Levelized COE among Alternative Technologies (Ref. 2)

A more recent study by EPRI and CPS Energy showed that for either new installations with or without CO₂ capture and retrofit of CO₂ capture to an existing plant, supercritical pulverized coal (SCPC) with oxygen combustion results in a lower overall busbar cost than IGCC³.

Table 2.1 – CO ₂ Capture Busbar Costs (Ref. 3)				
	Base Case (100% PRB)	CO ₂ Capture (100% PRB)	% Increase	\$/Mt CO ₂ Avoided
IGCC 20-year levelized busbar cost (2006 \$ real)	\$45.03	\$65.41	45%	\$26.28
SCPC 20-year levelized busbar cost (2006 \$ real)	\$39.28	\$62.00	58%	\$29.64

What Is Oxygen Combustion?

Oxygen combustion is essentially a means of replacing the nitrogen in air with flue gas, which is predominately CO₂, in order to significantly increase the CO₂ concentration and facilitate its disposal. In its first generation form (Figure 2.6), the coal plant is started up using air in the same manner as is currently used. Once at a minimum stable load, the air inlet is gradually closed while the exit stream is also controlled to force the flue gases to be recycled to replace the air. Nearly pure oxygen is introduced into the recycled flue gas to maintain safe and optimal combustion conditions in the boiler. Additional oxygen is introduced at each burner to further enhance combustion and reduce NO_x.

Once the air intake is fully closed and the boiler is operating on recycled flue gas and oxygen, load can be raised. As the process proceeds, the lack of nitrogen will naturally concentrate the CO₂ and other constituents. The coal analysis, oxygen purity, air-in leakage, and combustion efficiency will determine the degree of CO₂ concentration that can be achieved in the flue gas, typically about 80% on a mass basis.

Because 65% to 80% of the flue gasses are recycled back to the boiler, the flow extracted from the process for final cleaning and compression prior to disposal is relatively small, 35% to 20% of the flow to the stack for an air-fired unit of the same capacity. The removal of most of the nitrogen from the process also concentrates the flue gas constituents in the recycle loop by a factor of about 3.5 compared to air firing. As a result, even low sulfur coals will produce flue gas with SO₂ concentrations more typical of medium sulfur coals. For example, a coal with sulfur content of about 1% will produce sulfur concentrations within the boiler and recycle loop equivalent to a coal with about 3.5% sulfur. To prevent excessive corrosion, a scrubber may be needed.

The air quality control system (AQCS) for an oxygen-fired unit can be designed to treat all of the flue gas leaving the airheater or just the gas that is to be sequestered. The concentration of NO_x, particulate, Hg, SO₃ and SO₂, the emission limits, and concerns for erosion dictate the design flow to the AQCS. It is expected that the SCR (if present) and particulate collector will be full capacity. Dry scrubbing systems will be close to full capacity. Wet scrubbers may be used to treat all or only part of the flue gas.

Water is also concentrated to higher levels than in conventional flue gas and much higher than in air. In addition to the effect of recycle on moisture concentration, water added by a WFGD system significantly increases this concentration. Coal moisture content and combustion considerations dictate the level of moisture removal and reheat necessary to protect the equipment from corrosion and ensure pulverizer and combustion performance. Depending on the circumstances, moisture removal can be located in the full stream or in the stream to the pulverizers. When using a wet scrubber, the recycle gas will be saturated at whatever final temperature it is cooled to, so it must be slightly reheated before it enters the fans. The degree of final cleaning and compression will vary depending on the means of transportation of the CO₂ stream and the tolerance of the final disposal reservoir. At a minimum, the moisture must be removed to protect the CO₂ compressor.

Although the properties of the flue gas differ from those with air firing due to the lack of nitrogen, it has been found that with the proper recycle ratio, an existing boiler can be converted

to oxygen combustion without changing heat transfer surfaces and only experiencing a small impact on fuel efficiency in the boiler island. For new units, optimized arrangements are being studied that offer some reduction in equipment size and improved performance.

B&W Experience

When global warming concerns put the spotlight on the CO₂ emissions from power generation in the late 1990s, B&W revitalized its efforts and in 2000 became a member of CANMET's CO₂ Consortium. Canada was approaching carbon management more aggressively than the U.S.

In 2001, B&W began a focused R&D effort and teamed with Air Liquide to provide the largest element in the process on cost and performance impact, the oxygen. Shortly afterward, combustion tests were run at 5MBtu in B&W's Small Boiler Simulator (SBS) facility on both Illinois #6 and Powder River Basin (PRB) coals.

In parallel, B&W participated with Fluor in a study sponsored by the Canadian Clean Power Coalition (CCPC) and very recently completed a similar but much more current study with Air Liquide and Parsons for the U.S. Department of Energy.

The results of the pilot testing showed that with stable flame and optimized oxygen and recycle flue gas, almost 85% CO₂ concentration and a 65% NO_x reduction could be achieved under pilot conditions. This NO_x reduction is not only due to the lack of nitrogen in the system but also a re-burn effect and relatively low leakage. Unburned combustibles were lower than with air firing while furnace exit gas temperature, convection pass heat absorption, and boiler exit gas temperature were similar to air firing. There were no negative impacts on boiler operation. In fact, the system could be easily started up, load changed, and shut down, and operational behavior and steam side performance were nearly the same as for a conventional air-fired unit.

This success has led to the pursuit of a reasonably sized demonstration in a power plant environment.

City of Hamilton Project

In 2005, B&W with Air Liquide responded to a solicitation by the Ohio Coal Development Office (OCDO) by proposing to retrofit and demonstrate oxygen combustion at a 25 MWe scale on Unit 8 at the City of Hamilton plant in Hamilton, Ohio.

The Unit 8 boiler is a 25 MW 1963 vintage B&W Sterling Power Boiler with six burners and three B&W EL pulverizers currently burning a compliance coal. For the demonstration, the unit will burn an Ohio coal with 3.5% sulfur from the Waterloo mine.

The entire oxygen combustion process will be demonstrated with the exception of the final cleaning and compression step. This is due to a lack of a reservoir or use for the separated CO₂, the cost of the equipment required for final cleaning and compression, and ongoing air-fired operation of the plant subsequent to the testing.

Due to the high sulfur in the coal, the retrofit arrangement requires some flue and fan modifications, addition of a wet scrubber, moisture removal, SO₃ control, and a steam coil to reheat the gas prior to the fans. Except for refurbishment of the hot electrostatic precipitator (ESP), tubular airheater, and boiler casing to minimize air-in leakage, the remaining equipment will be unchanged and full load capability is anticipated. Figure 2.7 shows a schematic of the retrofitted plant.

The project is being accomplished in two Phases. Phase 1 was estimated to cost about \$1.45 M which is co-funded by B&W and Air Liquide as well as the Department of Energy and the Ohio Coal Development Office. In Phase 1, which are completed, process and performance

engineering, equipment layout, loads and basic balance of plant design leading to development of plant general arrangement drawings, a test plan, and restoration requirements were completed and a detailed cost estimate developed.

Phase 1 will be followed by a period of several months to assemble the team of participants who will provide the funding for Phase 2. Proposals have been recently submitted to the U.S. Department of Energy (DOE), Ohio's Third Frontier program, the Ohio Coal Development Office. If funding is obtained expediently, this could be the first large scale application in the world. Phase 2 is expected to take about a year and a half and will consist of the completion of details, equipment fabrication and installation, and testing over a six month period, about the first three of which will be commissioning.

Although specific to the City of Hamilton boiler, this large-scale pilot will demonstrate the design concepts that will also be suitable for other existing or new coal-fired plants. During the commissioning, tuning and test period of approximately six months, the project will provide performance data and operating experience that will support scale up of the process to commercial size. It will also allow for optimization of control concepts under various plant operating scenarios including transition from air to recycle flue gas with oxygen, load increase and decrease, and equipment trips such as main fuel trip, Oxygen Supply System (OSS) trip, and/or pulverizer trip. The testing will also assist in identifying potential areas for optimizing the integration of the OSS and the steam generation process on a reference plant installation. The overall goal is to retrofit a plant of sufficient scale to prove the process, controls, safety and performance of oxygen combustion with flue gas recycle that produces a stream of concentrated CO₂ that could be prepared for permanent storage (sequestration).

Figure 2.7 – City Of Hamilton Oxygen Combustion Conversion Schematic

The specific objectives of Phase 2 are:

- 1) Finalize all engineering developed during Phase 1 to support procurement and fabrication,
- 2) Procure, fabricate and install the equipment,
- 3) Commission and tune the process,
- 4) Execute a thorough series of tests to define performance and operability under various operating conditions and equipment faults,
- 5) Show that oxygen combustion with flue gas recycle is a technically viable option for either retrofit or new unit application, and
- 6) Determine the economics for a reference plant with CO₂ capture and compression to prove the viability of the technology for carbon control. This phase of the project will be summarized in a report addressing each of these objectives.

The City of Hamilton project will utilize existing equipment to the greatest extent practical. Due to the boiler size and steam conditions, steam cycle efficiency is not high. More importantly, the age and configuration of the equipment will introduce much more air infiltration than would occur in a modern plant design. In addition, firing high sulfur coal will require the addition of sulfur removal capability as well as dehumidification and temperature control of the recycle flue gas entering the FD fan. Both of these functions have been integrated into a wet scrubber. Figure 2.7 shows both the existing and new equipment for this oxygen combustion conversion.

At full load under oxygen firing conditions, 26,100 lb/h of Waterloo coal will be burned requiring 58,600 lb/h of oxidant at 95% purity which will produce 331,500 lb/h of flue gas leaving the boiler. A small amount of additional oxygen will also be introduced into the wet scrubber to produce saleable gypsum.

Based on field testing completed during Phase 1, the air infiltration was so great that the O₂ concentration leaving the boiler was 5.2% of the flue gas (dry volumetric). For design purposes the air infiltration into the boiler was assumed to be 2% (of stoichiometric air) and an additional 6% was assumed for the existing hot electrostatic precipitator and tubular airheater casings. This results in a flue gas composition leaving the process of 66.85% CO₂, 9.76% H₂O, 0.02% SO₂, 5.31% O₂, 15.98% N₂ and 2.08% Ar. The air infiltration significantly increases the nitrogen content. It is also noteworthy that the oxygen level in the flue gas is higher than would occur in a new unit due to this leakage. During equipment installation, an effort will be made to significantly reduce boiler and hot ESP air-in leakage. Typically, a total of 2% leakage is assumed for a new plant. If the 6% could be completely eliminated (unlikely), the gas composition would be improved to 80.31% CO₂, 9.24% H₂O, 0.02% SO₂, 2.30% O₂, 5.63% N₂ and 2.50% Ar. The actual gas composition during testing is expected to be somewhere in between.

Assuming the higher leakages, the gas flow entering the wet scrubber is 353,800 lb/h at 351F and 351,300 lb/h leaves at 138F; most of the mass reduction is moisture removed. In order to prevent moist gas from entering the forced draft fans, it is reheated, typically 10F to 15F. However, since the existing airheater is being used and the surfaces will not be modified, the gas will be reheated to about 250F by a steam coil gas heater (SCAH) to maintain the

temperature to the pulverizers and windbox and approximate the coal conveying and combustion conditions currently experienced with air firing.

FGD Design

Following the demonstration project the City will continue to use the WFGD system to allow flexibility in coal purchasing. This required that the WFGD system be designed to treat the full gas flow both on oxygen firing and air firing. When burning coal with 3.5% sulfur content the air firing SO₂ concentration will be about 3520 ppm_{dv}. Operating with oxygen firing and only treating 26% of the gas flow (the gas to be sequestered) would have increased this concentration to about 9100 ppm_{dv}. Concerns about corrosion, achieving 95% SO₂ removal efficiency and over designing the wet scrubber for air firing dictated that all of the flue gas be treated.

The WFGD system is designed for an expected SO₂ removal efficiency of up to 98%. The B&W Multi Contact Absorber (MCA) design is being used. The absorber contains multiple stages of absorption trays and spray headers to maximize the SO₂ removal and optimize the capital and operating costs of the WFGD system. Table 2.2 lists the key design parameters.

Table 2.2 – Key Design Parameters	
Parameter	
Target SO ₂ removal, %	98%
Required SO ₂ removal, %	95%
Inlet SO ₂ , ppm _{dv}	3528
Gas Velocity, ft/sec	6.13
Liquid Flux, gpm/ft ²	6.0 ± 5%
pH	5.21
Ca/S stoichiometry	1.02
Gypsum Quality	Commercial
Reagent	Municipal Waste Sludge

The system will use municipal waste sludge with 91to 92% available CaCO₃. The sludge is delivered as slurry with 20 to 25% suspended fine solids. This is stored in the reagent feed tank and pumped directly to the absorber as required to control the SO₂ removal.

The dewatering system consists of primary hydroclones and horizontal belt filters. Secondary purge hydroclones are used to develop a purge stream for control of chlorides and fines.

Operation with oxygen firing recycles most of the water produced by combustion and the saturation of the gas in the wet scrubber. Unlike SO₂, water is not removed in the air quality control system. The steady state water content of the flue gas leaving the scrubber would be about 50% by weight if water is not removed. The saturation temperature would be about 180F. Therefore the flue gas leaving the scrubber is cooled to about 138F to remove water and provide a flue gas with about 18% moisture.

Oxygen Supply System (OSS)

The proposed demonstration project requires oxygen volumes ranging from 200 to 500 short tons per day with peak instantaneous flow rates equivalent to 660 short tons per day. During Phase I of the project, a customized and dedicated Air Separation Unit (ASU) was considered. Even though no technical barriers were found for construction and operation of the plant, alternate solutions for the oxygen supply were sought to minimize project cost without jeopardizing the technical value. These options included 1) a smaller ASU complemented by liquid oxygen (LOX), 2) relocation and adaptation of a used ASU, 3) coupling multiple standard ASUs, and 4) full oxygen supply by LOX. Considering the fact that the oxygen requirement for the proposed demonstration is for approximately six months, the most economical option is to use LOX.

Air Liquide's OSS is designed to provide gaseous oxygen at the pressures and flow rates specified for the process. The OSS will consist of the following:

- Multiple vacuum insulated LOX storage vessels will be installed, with a total capacity of approximately 180,000 gallons (856 tons).
- Unloading stations to allow for multiple transport deliveries simultaneously.
- Steam-heated vaporization equipment engineered to provide greater than 620 tons per day continuous flow rate.
- Pressure control systems designed to maintain process pressure settings at predetermined set points.

The Phase 1 design was completed in May and the costs assembled in early June 2006. As described above, the initial approach using a full capacity ASU was too costly and the supply was changed to LOX. Likewise, the cost of the initial design for the boiler island equipment is much higher than initially anticipated and includes several areas that are candidates for cost reduction. Consequently, B&W is working to significantly reduce cost while simultaneously seeking additional funding opportunities. In spite of this cost discrepancy, several funding proposals have been submitted in hopes that by the time of project selections in the fall of 2006, this cost/funding gap can be closed.

3.0 PROCESS

3.1 General Design Basis

All components including the OSS, wet scrubber, fans and flues are commercially available equipment but none have been operated in this mode or with this flue gas composition. The unique requirements for the boiler island equipment are related to the difference in flue gas composition and properties compared to air or flue gas from conventional combustion with air. Table 3.1 summarizes the specification for all major equipment and identifies any unique requirements.

Table 3.1 - Specification Of Major Equipment		
Equipment	Specification	Unique Requirements
Wet Scrubber	Inlet gas flow, temperature and gas composition is given in Table 3.2.2 stream 5.	Without a scrubber, use of high sulfur coal will result in excessive sulfur in the boiler under oxy-firing, which could cause severe corrosion. In addition, the plant will be unable to comply with the current emission limit. Integral to the scrubber is also moisture control. Moisture in the flue gas must be reduced to ensure reliable pulverizer operation and coal conveying.
Fans	Gas flows, temperatures and gas compositions entering the various fans are given in Table 3.2.2 streams 2 & 5.	Addition of a scrubber will impose several inches of water in additional pressure drop which requires a booster fan be installed after the existing ID fan. Since the flue gas is not diluted with nitrogen from air, the gas has a higher concentration of CO ₂ and contains moisture making the flue gas much denser. The Forced Draft and Primary 'Air' fans were checked during Phase 1 considering the higher temperature of the recycle gas, compared to ambient air, and the recycle mass flow required and found to be adequate.
Interconnecting Flues	Stream flows, and temperatures for the major flues are shown in Table 3.2.2 streams 2, 5, 6, 7 & 8.	New flues to connect the existing ID fan to the new booster fan, from the booster fan to the wet scrubber, from the wet scrubber to the steam coil gas reheater and from the gas reheater back to the FD fan or stack were designed. Dampers to control air and flue gas to permit both air and oxy firing were added.
Steam Coil Gas Reheater	A new heat exchanger will be used to heat the flue gas leaving the scrubber prior to entering the FD fan.	A steam coil heater was selected and sized to provide capability to compensate for the unique gas composition and heat transfer properties of the flue gas. Data acquired during the demonstration will provide confirmation of these assumptions.
Controls	Commercially available control system hardware will be utilized but the software logic will be unique.	Additional hardware to accommodate the larger number of inputs and outputs for the digital control system (DCS) is required and software modifications will be required to control oxygen to the various locations and the recycle flue gas flow.
Oxygen Supply System	The OSS equipment is commercially available	There are no unique requirements of the OSS except the oxygen delivery pressure is considerably lower than for a typical oxygen application. Due to the size and cost, it was decided that a full air separation plant is not practical so a liquid oxygen supply system will be provided.

3.2 Heat and Mass Balances

Figure 3.1 shows a process flow diagram of Unit 8 converted to oxygen (or air) firing. Table 3.2 provides a basic heat and mass balance for the process and Table 3.3 gives the assumed coal analysis. Figure 2.7 shows the equipment arrangement schematically. In this demonstration of oxygen combustion, the unit will be converted utilizing existing equipment to the greatest extent practical. This will reduce the cost of the retrofit and preserve the capability to operate at full load with air but it will also have some minor impacts on performance in the oxygen combustion mode. For example, some reduction in boiler efficiency is expected. However, the unit is expected to be operated within the limits of the existing heat transfer surfaces and materials with no reduction in steam production anticipated.

Due to the boiler size and steam conditions, steam cycle efficiency is not high. More importantly, the age and configuration of the equipment will introduce much more air infiltration than would occur in a modern plant design. In addition, firing high sulfur coal will require the addition of sulfur removal capability as well as dehumidification and temperature control of the recycle flue gas entering the FD fan. Both of these functions have been integrated into a wet scrubber in a novel manner. Figure 2.7 is a schematic of the plant showing both the existing and new equipment for this oxygen combustion conversion.

Figure 3.1 - City of Hamilton Retrofit Process Flow Diagram

A booster fan will be added after the existing induced draft (ID) fan and its outlet flue will be connected to the new wet scrubber. From the wet scrubber a flue will convey the gases to a steam coil heat exchanger to increase the gas temperature to ensure no moisture droplets enter the FD fan. After this fan, the concentrated CO₂ flue gas stream will be split, about 70% of the

flue gas will be conveyed to the FD fan as recycle gas and the other 30% will be extracted for “sequestration” which, for this project will be sent up the existing stack.

Table 3.2 - Heat and Mass Balance

Stream #	1	2	3	4	5	6	7	8	
	Oxidant	Recycle Gas	Flue Gas Lvg Boiler	Flue Gas Entering AH	Flue Gas Ent Scrubber	Flue Gas Lvg Scrubber	Flue Gas Lvg Reheater	Flue Gas to Sequestration	
% Volume wet								% Wet	% Dry
CO ₂	0.00	51.52	53.91	50.70	50.70	51.52	51.52	51.52	63.61
H ₂ O	0.00	19.01	21.34	20.18	20.18	19.01	19.01	19.01	N/A
SO ₂	0.00	0.01	0.27	0.25	0.25	0.01	0.01	0.01	0.01
O ₂	95.0	6.24	5.01	5.93	5.93	6.24	6.24	6.24	7.70
N ₂	1.50	21.03	17.18	20.80	20.80	21.03	21.03	21.03	25.97
Ar	3.5	2.19	2.30	2.15	2.15	2.19	2.19	2.19	2.70
SUM	100	100	100	100	100	100	100	100	100
% Mass									
CO ₂	0.00	65.03	67.65	64.32	64.32	65.03	65.03	65.03	
H ₂ O	0.00	9.82	10.96	10.48	10.48	9.82	9.82	9.82	
SO ₂	0.00	0.02	0.49	0.46	0.46	0.02	0.02	0.02	
O ₂	95.60	5.72	4.57	5.47	5.47	5.72	5.72	5.72	
N ₂	1.16	18.82	16.34	19.27	19.27	18.82	18.82	18.82	
Ar	3.24	0.59	0.62	0.58	0.58	0.59	0.59	0.59	
SUM	100	100	100	100	100	100	100	100	
Mass Flow klb/h	58.6	245	331.5	351.1	351.1	349.0	349.0	101.5	
Temperature F	59	138	612	585	352	138	150	138	

Table 3.3 - Coal Analysis

Stream 9	Ultimate Analysis
% Wt	
C	68.15
S	3.00
H2	4.67
H2O	5.00
N2	1.25
O2	6.43
Total	100.00
Mass Flow klb/h	28.07

3.3 Plant Requirements

Additional power will be required to operate the new fan and wet scrubber island. Table 3.4 provides a listing of all additional power needs. Tables 3.5 through 3.8 show the plant inputs and outputs for both oxygen and air firing.

Table 3.4 - Auxiliary Power Loads									
	Line No.	Item	Tot Qty Inst	Inst Hp Each	Inst Hp Total	Inst kW Each	Inst kW Total	Norm Qty Oper	Oper kW Average
REAGENT PREPARATION ISLAND									
2	1	REAGENT STORAGE TANK AGITATOR	1	15.0	15.0	11.2	11.2	1	8.2
1	2	REAGENT FEED PUMPS	2	5.0	10.0	3.7	7.5	1	3.0
0	3	REAGENT PREP AREA SUMP AGITATOR	1	3.0	3.0	2.2	2.2	1	1.9
1	4	REAGENT PREP AREA SUMP PUMPS	2	10.0	20.0	7.5	14.9	1	5.9
0	5	TRUCK UNLAODING PUMP	2	7.5	15.0	5.6	11.2	1	3.4
									22.4
ABSORBER ISLAND									
1	5	ABSORBER RECIRCULATION PUMP A	1	150.0	150.0	111.9	111.9	1	96.5
1	6	ABSORBER RECIRCULATION PUMP B	1	150.0	150.0	111.9	111.9	1	102.6
1	7	ABSORBER RECIRCULATION PUMP C	1	200.0	200.0	149.1	149.1	1	110.0
0	8	ABSORBER RECIRCULATION TANK AGITATOR	3	15.0	45.0	11.2	33.6	3	29.7
1	9	OXIDATION AIR COMPRESSOR (MAIN DRIVE)	2	150.0	300.0	111.9	223.7	1	99.7
0	10	OXIDATION AIR COMPRESSOR LUBE OIL HEATER	2	5.0	10.0	3.7	7.5	1	2.1
0	11	OXIDATION AIR COMPRESSOR LUBE OIL PUMP	2	2.0	4.0	1.5	3.0	1	1.4
1	12	ABSORBER BLEED PUMP	2	5.0	10.0	3.7	7.5	1	2.1
1	13	ME WASH PUMP	2	25.0	50.0	18.6	37.3	1	17.3
0	14	ABSORBER AREA SUMP AGITATOR	1	3.0	3.0	2.2	2.2	1	1.8
1	15	ABSORBER AREA SUMP PUMPS	2	20.0	40.0	14.9	29.8	1	11.8
1	16	CONDENSER COOLING WATER PUMPS	2	100	200.0	74.6	149.1	2	128.9
DEWATERING ISLAND									
0	17	SECONDARY HYDROCLONE FEED PUMP	2	5.0	10.0	3.7	7.5	1	2.9
1	18	HORIZONTAL BELT VACUUM FILTER DRIVE	2	5.0	10.0	3.7	7.5	1	3.6
1	19	FILTRATE PUMP	2	3.0	6.0	2.2	4.5	1	2.2
1	20	VACUUM PUMP	2	30.0	60.0	22.4	44.7	1	22.1
1	21	CAKE WASH WATER PUMP	2	3.0	6.0	2.2	4.5	1	2.2
1	22	CAKE WASH WATER HEATER (NOT NEEDED)	0	-	-	-	-	-	
1	23	CHLORIDE PURGE PUMPS	2	3.0	6.0	2.2	4.5	1	0.3
0	24	RECLAIM WATER TANK AGITATOR	1	5.0	5.0	3.7	3.7	1	3.7
1	25	RECLAIM WATER PUMPS	2	5.0	10.0	3.7	7.5	1	2.9
TOTAL POWER FOR COH UNIT #8					1,338.0		997.8		688.7

WFGD SYSTEM									
Boiler Island									
0	32	ID Booster Fan 2300 VAC [will change to 4160 VAC]	1	400.0	400.0	298.3	298.3	1	212.0
0	33	Space Heaters for ID Booster Fan Motor	0		-	-	-	0	
0	34	ID Booster Fan Damper Actuator	1	0.1	0.1	0.1	0.1	1	0.1
0	35	Guillotine Dampers	5	2.0	10.0	1.5	7.5	5	6.4
0	36	Dilution Air Fan	0	-	-	-	-	0	
0	37	Space Heaters for Fan Motor	0		-	-	-	0	
0	38	Fan Damper Actuator	1	0.1	0.1	0.1	0.1	1	0.1
0	39	Heating - Space heaters for Building [none expected]		Later					
0	40	Ventilation Fans - Louver Dampers w Fans		Later					
0	41	Lighting - Interior		Later					
0	42	Lighting – Exterior; Assume not required, except door lights		Later					
0	43	120 VAC Outlets		Later					
0	44	Maintenance, such as welding outlets		Later					
		Electrical Building [expect to add to Scrubber or De-watering]							
		Heating		Later					
		A/C		Later					
		Lighting		Later					
		120 VAC outlets		Later					
					-	-	-	-	
TOTAL POWER FOR COH UNIT #8 Boiler SYSTEM					410.3		305.9		218.6

Table 3.5 - Inputs Streams for Oxycombustion at Full Load		
Description	Flow (klb/h)	Comments
Fuel	26.1	Waterloo Coal
Oxygen to Boiler	58.6	@ 59 F and 8 psig, 30 trucks/day @21 tons each
Oxygen to Scrubber	0.98	@ 59 F and 30 psig
Limestone	2.7	Municipal Sludge, 10 trucks/day @ 5000 gal each
Scrubber Makeup Water	16.1	
Trona	0.20	1 Ton Supersacks, 1 truck/day
Steam for ASU	10.0	Saturated @ 160 psig
Steam for Steam Coil	11.8	@ 700 F and 400 psig

Table 3.6 - Outputs Streams for Oxycombustion at Full Load		
Description	Flow (klb/h)	Comments
Ash	3.0	
Gypsum	4.55	Commercial quality, 135 klb/day, CoH to determine # trucks
Stack Gas	101.5	
Water from Scrubber	30.0	

Unburned Carbon	0.067	
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Table 3.7 - Inputs Streams for Air Firing at Full Load		
Description	Flow (klb/h)	Comments
Fuel	25.6	Waterloo Coal
Limestone	3.2	Municipal Sludge
Scrubber Makeup Water	12.5	
Trona (if used)	0.20	1 Ton Supersacks, 1 truck/day

Table 3.8 - Output Streams for Air Firing at Full Load		
Description	Flow (klb/h)	Comments
Ash	2.9	
Gypsum	5.63	Commercial quality, 135 klb/day, CoH to determine # trucks
Stack Gas	323.3	
Water from Scrubber	2.14	
Unburned Carbon	0.065	

3.4 Predicted Emissions (with Air Firing and WFGD)

Since the wet scrubber will continue to operate after the oxycombustion testing, emissions performance firing the high sulfur Waterloo coal while firing with both oxygen and recycled flue gas and air as the oxidant were developed in order to assist the City of Hamilton in assessing whether any permitting issues might arise. The addition of the wet scrubber will substantially reduce SO₂ and particulate emissions but since it was decided that the burners will not be replaced, NO_x and CO emissions will be unchanged.

Table 3.9 shows the current emissions and those predicted for both oxygen firing and air firing after the retrofit with the WFGD in service. Table 3.10 provides predicted mercury removal. Though titled “Air Firing”, the mercury removal is expected to be achieved with oxycombustion as well.

Table 3.9 - Predicted Emissions with Air Firing				
Operating Condition >>>>>		Current Baseline	Oxy-Firing with WFGD	Air-Firing (future)
			(no dilution air)	with WFGD
Boiler Load	%	100	100	100
Boiler Heat Input (nominal)	MBtu/h	320	320	320
Fuel HHV	Btu/lb	12,320	12,320	12,320
Fuel Sulfur Content	%	0.9	3.0	3.0
Predicted Scrubber Removal Efficiency	%	N/A	Up to 98.0	Up to 98.0
Stack Gas Mass Flow	1000 lb/h	332	106	376
Gas Temperature to Stack	Degree F	325	250	132
Stack Gas Vol. Flow	ACFM	106,374	26,417	94,812

Stack Gas Emissions:				
NO_x	lb/h	120 (avg)	72 (avg) ⁽¹⁾	120 (avg) ⁽³⁾
NO_x	lb/MBtu	0.35 to 0.40	0.20 to 0.25	0.35 to 0.40
CO	lb/h	32	32	32
CO	lb/MBtu	0.10	0.10	0.10
SO₂	lb/h	464	< 77	77
SO₂	lb/MBtu	1.45 ⁽²⁾	< 0.24	0.24
Particulate	lb/h	44.8	< 22.4	< 22.4
Particulate	lb/MBtu	0.14 ⁽¹⁾	< 0.07	< 0.07
HCL	% Removal	N/A	> 95	> 95
Mercury	% Removal	N/A	> 80	> 80

Notes: ⁽¹⁾ - This is a conservative value for permitting which is expected to be lower in practice.
⁽²⁾ - Current Permit Limit
⁽³⁾ - Since over-fire air ports would remain following the oxycombustion testing, some NO_x reduction could be achieved. Firm predictions are difficult but depending on how much over-fire air is used, as much as a 20% NO_x reduction is possible. However, the reduction in air flow to the burner due to diversion to the over-fire air ports would reduce mixing in the burner somewhat and could produce a 25% to 50% increase in unburned carbon. It may be possible to improve UBC with burner modifications.

Table 3.10 - Predicted Mercury Emissions with Air Firing		
Mercury	Potential Guarantee (Permit)	Predicted Performance
Removal of Oxidized Hg	85	90
70% Oxidized	60	63
80% Oxidized	68	72
90% Oxidized	76	81

3.5 Technical Risks

Table 3.11 lists the primary project risks and shows that the most significant technical risks are related to operation and process control. This demonstration will require 1) control of the complex transition from air firing to gas recycle modes, 2) discharge of the flue gas with concentrated CO₂ up the existing stack, and 3) control of the oxygen supplied to at least three locations to ensure safe concentration results. The first one poses some technical risk but the other two are not considered significant challenges.

The unit will initially be fired with air until stable coal combustion has been established at 25% to 30% load based on minimum stable load with one of the three pulverizers in service. Operation at that load will be maintained as the damper controlling the flow from the wet scrubber outlet to the stack is gradually closed forcing gas flow into the recycle flue. Simultaneously the forced draft fan air intake damper will begin to close allowing flue gas to enter the inlet of the forced draft (FD) fan along with air.

As previously described, the volume of flue gas that will be released to the atmosphere for this large-scale pilot (which would be sequestered in a commercial application) will be about 32% of the volume flow leaving the wet scrubber at about 130°F to 140°F. In order to maintain the required temperature to the windbox for combustion after mixing the cold oxygen, the cool but saturated flue gas leaving the wet scrubber will be reheated to about 250°F. This is necessary to protect the FD fan since replacement of the existing airheater is not feasible. Since it is not diluted with nitrogen from air, the flue gas sent to the stack will contain a higher concentration of CO₂, which is heavier than air (at the same temperature). It will also contain slightly superheated water (this water would be removed for sequestration). Due to this temperature, stack calculations and dispersion modeling completed in Phase 1 have indicated sufficient buoyancy to release the gas and disperse it safely. During Phase 2, this information will be reviewed with the proper environmental authorities to ensure safe release during testing.

It should be noted that the amount of CO₂ emitted to the atmosphere is no greater than that associated with air firing (except a very small amount due to reduced fuel efficiency and reactions within the scrubber). Of importance is the fact that there will be no increase in release of any constituent under oxygen combustion conditions that will be greater than current levels. In fact, NO_x, SO₂, SO₃, mercury and particulate will all be reduced. Further heating and dispersion studies will be made to ensure that safe conditions are maintained under all circumstances. Dilution air may also be mixed with this gas to reduce the CO₂ concentration, density, and moisture levels significantly.

Oxygen is a colorless, odorless, tasteless, nonflammable gas. It makes up about 21% of the air in our atmosphere. The potential danger is that many substances that do not burn in air, may burn in an oxygen-enriched atmosphere (>23.5% O₂). Therefore, it is essential that oxygen be prudently used and controlled. Human exposure to gaseous oxygen at concentrations up to 21%, as in air, presents no risks but higher concentrations (>25%) present health risk ranging from inflammation of organic matter in the body with potential death. However, oxygen production is a commercial process operated safely in numerous locations throughout the world and as such, the OSS and oxygen control does not present any significant new safety or technical risk to the project.

Table 3.11 - Project Risk Table			
SYSTEM	DESCRIPTION	TYPE	LEVEL (H,M,L)
Boiler	CO ₂ leakage into plant	Safety	L
Boiler	Steam generation	Performance	L
Combustion	Achieve and maintain stable combustion during operation in either mode	Operational	L
Combustion	Achieve NO _x level desired	Performance	M
Wet Scrubber	Sulfur removal and gas cooling capacity	Performance	L
Stack	Discharge and dispersion of gas stream with concentrated CO ₂	Safety	L-M
Oxygen System	Proper mixing must be achieved to prevent local areas where the oxygen concentration >23.5%.	Safety	L (controlled)

Trona System	There is limited but successful experience with Trona for SO ₃ control with typical PC flue gas but not with the flue gas composition resulting from oxygen combustion with flue gas recycle.	Performance	M
Trona System	Particulate control impact on the Hot ESP. Data to date has shown the Trona improves removal.	Performance	L
Process Control	Ability to control transition from air to oxygen operation smoothly and automatically	Operational	M

Carbon dioxide is heavier than air at the same temperature so leakage from the boiler enclosure and flues is of some concern for safety. Fortunately, any leakage would be at much higher than the ambient temperature causing the gas to rise and mix with the air as it cools. In addition, most of the flues and equipment operate under negative draft meaning air would leak into the process rather than flue gas leaking out. This is true except for the very short ducts from the FD fan to the tubular airheater, the tubular airheater 'air' side casing, the ducts to the pulverizers and windbox, and the lower portion of the boiler enclosure, which is gas tight membrane wall construction.

The testing will help quantify the accuracy of heat transfer predictions in the boiler and other heat transfer equipment, hot ESP, and wet scrubber performance, Trona system, and other functional aspects affected by the unusual flue gas composition and high CO₂ concentration which for commercial application would represent a performance risk relative to guarantees.

4.0 BOILER ISLAND DESIGN SUMMARY

4.1 Background

Figure 4.1 shows the location in the process and basic arrangement of the equipment described in this section schematically. The City of Hamilton Municipal Electric Plant Unit 8 boiler in Hamilton Ohio is a balanced draft B&W Sterling Power Boiler design of 1963 vintage (National Board number A-21638) with three B&W EL-38 pulverizers and six modified 30 inch diameter B&W circular burners firing approximately 28,500 lb/h (350 MBtu/h) of an eastern bituminous (non-Ohio) compliance coal to produce about 250,000 lb/h of steam at 900F and 875 psig.

The boiler furnace is 20 ft. deep and 16 ft. wide with a 60 inch diameter steam drum and a 42 inch diameter mud drum. The boiler is followed by a hot electrostatic precipitator provided by Air Correction Division of Universal Oil Products Company and installed in the mid 1970's, a tubular airheater, primary air, forced draft and induced draft fans, and a roof mounted stack. Grade elevation is 580 ft and the top of the boiler steel is at 685 ft. The roof mounted stack extends to 757 ft. The fans are all located on the operating floor at elevation 596 ft.

Figure 4.1 – City Of Hamilton Oxygen Combustion Conversion Schematic

4.2 Boiler Design Basis

The CoH plans to return to air firing after the oxygen firing test period burning a lower cost high sulfur coal. Therefore, the new equipment must be designed to accommodate coal combustion using either air or oxygen as the oxidant. Except for the wet scrubber and

auxiliaries and the additional ID booster fan, the equipment that will only be used during the six month oxy-firing test will be removed following testing. The new equipment must not increase the plant's current emission levels despite burning the higher sulfur coal.

Design Coals

Table 4.1 shows the currently burned compliance coal (Eastern Kentucky) and an Ohio coal used for the analysis.

Table 4.1 - Coal Analyses			
Constituent		Eastern Kent.	Ohio
Ultimate Analysis	% By	Wt	Wt
	Ash	7.50	11.50
	S	1.00	3.00
	H2	4.85	4.67
	C	73.65	68.15
	H2O	5.00	5.00
	N2	1.40	1.25
	O2	6.60	6.43
		100.00	100.00
	BTU/LB	13,000	12,320
Proximate Anal %			
	Moisture	5.00	5.00
	Vol. Matter	35.35	36.00
	Fixed C	52.15	47.50
	Ash	7.50	11.50
	Total	100.00	100.00

Performance Testing

Tests were conducted on this unit during the week of November 28, 2005. The purpose of these tests were to establish boiler baseline values for calibrating boiler performance design tools for current operation with the compliance coal and air as the oxidant prior to using those tools to analyze the unit performance firing a high sulfur bituminous coal with oxygen as the oxidant and recirculated flue gas.

A total of two tests were conducted during this test program; the first at a reduced load condition, and the second at a higher load but not at full load. Full load could not be achieved due to large variations in feedwater flow and NOx emissions that would exceed permissible limits. The results of these tests were used to establish expected boiler operating conditions on air firing and oxygen firing. Table 4.2 contains a description of the tests conducted and a summary of the pertinent performance information.

Table 4.2 - Boiler Performance Summary from Performance Test		
Description	Test 1	Test 2

Date	11/30/05	11/30/05
Time	7:40-8:45	14:40-16:49
Gross Megawatts	14	22
HP Steam Flow, klb/h	121.4	194.6
Sec. SH Spray Flow, klb/h	10.2	19.8
Main Steam Temperature	931	940
Furnace Exit Gas Temperature (HVT) F	1550	1913
Gas Temperature Lvg Blr, F	581	595
Gas Temperature Lvg AH, F	324	343
Excess Air Lvg Blr, %	68.2	32.0
Excess O ₂ Lvg Blr, %	8.64	5.19
Excess O ₂ Lvg HESP, %	9.86	6.62
Average O ₂ Lvg AH, %	10.29	6.95
Average CO Lvg Blr, ppm	20.96	61.54
Average NOx Lvg Blr, ppm	244.00	200.93
Average NOx Lvg Blr, lb/MBtu	0.50	0.32

A detailed test report is contained in Appendix 2.

4.3 Boiler Performance

The boiler performance for oxygen firing was calculated using a spreadsheet created by B&W. The program calculates the heat and material balance of the boiler island when in the oxy combustion mode. Furnace Gas Temperature (FEGT) was calculated using a special version of B&W's furnace performance program. The heat and mass balances can be found in section 3.2.

The goal was not to make any significant changes to the boiler. This was accomplished by recirculating enough flue gas back through the system (70%) to maintain similar boiler outlet conditions. Table 4.3 shows a comparison of the original predicted boiler performance and current predictions with both air and oxygen as the oxidant (adjusted per performance test results).

Table 4.3 - Boiler Predicted Performance Comparison			
STEAM, ACTUAL, Mlb/h	250	250	250
	As Built Air	Air	O ₂
TYPE FUEL	PC	PC	PC
EXCESS AIR LVG BLR %	25	25	0

BURNERS, # IN USE	6	6	6	
				(M lb/hr)
O2 Trim to Burners	n/a	n/a	6	
O2 LVG BLR %	4.25	4.11	4.57	
CONT. BLOWDOWN	0	0	0	
FUEL	26.2	25.6	26.1	
FLUE GAS LVG BLR	332	323.3	331.5	
Flue Gas Entering Scrubber	n/a	342.4	351.1	
AIR LVG AH	269	273	306	
PRIM SPRAY	16.9	17.7	13.8	
O2 Added before Tubular AH	n/a	n/a	58.6	
STEAM AT SH OUTLET	875	875	875	
MIN OPER IN BLR DRUM	940	940	940	
DROP, DRUM TO SH OUTLET	65	65	65	
SUPERHEATED STEAM	900	900	900	
O2 Added before Tubular AH	n/a	n/a	59	
FLUE GAS LVG BLR	621	620	612	
FLUE GAS LVG AH	325	333	393	
FEED WATER ENTERING BLR	350	350	350	
RECYCLE/AIR ENTERING AH	80	80	215	
RECYCLE/AIR LVG AH	471	424	448	
Flue Gas Into Scrubber	n/a	346	406	
Flue Gas After Scrubber	n/a	133	138	

Metals

An analysis of the superheater metals was completed. The existing metals were found to be adequate for both the oxygen and air firing conditions.

4.4 Existing Hot Electrostatic Precipitator (HESP)

The HESP, supplied by Air Correction Division of Universal Oil Products Company, was added to the unit in the mid 1970's (B&W built the boiler in 1963). Very few drawings or technical information were available for this equipment, but it was not desired to make any modifications unless a significant problem was encountered.

During the November 2005 performance test, it was found that the oxygen across the HESP increased by approximately 1.5%. It was decided that during the conversion to oxygen firing, repairs would be made to the HESP casing to try and reduce this air infiltration as much as possible. This test was completed in late June of 2006 and the report is contained in Appendix 2.

4.5 Existing Tubular Airheater (TAH)

The tubular air heater offered similar challenges as the HESP. Due to the increased leakage across the TAH some of the oxygen input into the system would be lost to the surroundings. Since the power requirements for separating the oxygen from the nitrogen molecules used eight megawatts of power it is desired to lose as little oxygen from the system as possible. It was determined the leakage investigation would also include the TAH.

4.6 New Recycle Gas Reheater

A SCAH will be added after the wet scrubber in order to raise the temperature of the flue gas above saturation. This allows the flues between the scrubber and the boiler to be made out of carbon steel instead of the alloy that would have been needed to inhibit corrosion and to prevent water droplets from entering the FD fan.

During the study it was found that there was insufficient air heater surface to maintain the recirculated flue above a minimum temperature required for burner performance after mixing of cold oxygen (59°F) with the gas. It was decided to use the same SCAH to raise the flue gas temperature leaving the scrubber from 132°F to 250°F. This temperature increase offsets the input of the 59°F oxygen gas from the oxygen supply system provided by Air Liquide (AL) and helps maintain the desired tubular air heater air side exit gas temperature. The SCAH will not be used during normal air operation after the oxycombustion testing since the saturated WFGD exit gas will be sent to the stack and not recycled.

Table 4.4 - SCAH Design Conditions

Load lb/h	Wet Flue Gas Mlb/h	Ent. temp.	Lvg. temp. req'd	Steam psig	Steam psig req'd	Steam temp.	Cond.. lb/h	Draft Loss Inches H2O
250,000	375	138	250	400	400	700	11,817	1.5

4.7 Oxygen Injection Locations

The oxidant (95% O₂, 3.5% Ar, and 1.5% N₂) will be injected in three locations; the FD fan inlet, the FD fan outlet and directly into each burner. The injection points were based on the guidance from Air Liquide and equipment layout constraints. From pilot scale testing at the B&W Research Center with Air Liquide in the Small Boiler Simulator (SBS) at the Alliance Research Center (see Appendix 4), it was determined that a percentage of oxygen must be injected directly into the burners. This percentage will be determined during start-up and burner tuning during Phase 2. The Heat and Mass Balance Diagram in Section 3.2 shows the locations in the process and Figure 4.1 shows them schematically.

4.8 Fans

The current arrangement for Unit 8 is comprised of an induced (ID) and a forced (FD) draft fan, both originally supplied by American Standard. The addition of the wet scrubber a second ID fan (ID Booster) to compensate for the scrubber, SCAH, damper, and flue draft losses.

Under the oxygen firing mode the recirculated flue gas will be recycled and enter the FD fan through the existing air inlet duct. The gas flue is connected into the existing air inlet duct work which will use a guillotine damper on the air inlet side to keep the air from entering the system during recirculation. Barron industries was contacted to ensure the existing FD fan would be capable of the higher inlet temperature to the fan (250°F [oxygen] to 80°F [air]). They confirmed the fan should be adequate, but a technician would need to analyze the rotor prior to giving a final decision. This analysis would be accomplished in Phase 2 of the project.

Table 4.5 - FD Fan Design Conditions

Speed (RPM)	Flow (acfm)	Static Pressure (in. wg)	Inlet Temp (F)
1,180	58964	9.9	150

A booster ID fan will be added prior to the wet scrubber to overcome the additional pressure drop created by the wet scrubber, SCAH, and the additional flue work. The fan was designed to force the flue gas from the existing ID fan outlet flue into the wet flue gas desulfurization (WFGD) unit, SCAH and flue work prior to the FD fan. This fan will be used in both oxygen and air firing methods.

Table 4.6 - ID Booster Fan Design Conditions

Speed (RPM)	Flow (acfm)	Static Pressure (in. wg)	Temp (F)
1180	99461	10	338

The hot primary oxidant fans on this unit are integral with the pulverizers and are driven by the same motor as the pulverizer. Consequently they were checked with the pulverizers and found to be adequate for the expected oxycombustion conditions.

A stack dilution fan was also considered but not included. Since the CO₂ rich gas from oxycombustion has a higher molecular weight than a conventional flue gas, a dispersion fan for the stack gases was considered. This fan would take ambient air and mix it with the highly CO₂ concentrated flue gas stream prior to entering the base of the stack. The air would be introduced at an angle to help aid in the mixing of the two streams. The objective of this fan would be to make the concentration of the flue gas exiting the stack similar to the original buoyancy conditions in order to keep the CO₂ rich gas from spilling over the top of the stack and falling onto the plant and city; however, preliminary calculations have shown that this fan is not likely to be necessary. Final confirmation will be made during Phase 2.

4.9 Pulverizers

The pulverizers were reviewed by B&W's specialists and no concerns were identified for the high sulfur coal with oxygen as the oxidant. This will be confirmed in Phase 2. The main concern is the amount of moisture in the recycled gas which is significantly higher than occurs with air firing due to the concentrating effect. However, the outlet temperature of the pulverizer could be increased to offset the high moisture content. Sufficient Primary oxidant temperature must be achieved in the airheater to accomplish this, tempering oxidant will be provided by bypassing the airheater with some recycled flue gas. This will also be finalized in Phase 2 of the project.

4.10 Dampers

The new arrangement will include the addition of four tight shut-off guillotine type dampers, one louver shut-off, and one flow control damper. There will also be one flow measuring device for the recirculated gas flow.

Table 4.7 - Dampers			
	Style	Type	Comments
Guillotine 1	Guillotine	Shut-off	From Bottom
Guillotine 3	Guillotine	Shut-off	45° angle
Guillotine 4	Guillotine	Shut-off	
Guillotine 6	Guillotine	Shut-off	
Louver 5	Louver	Shut-off	
Flow Control 2	Louver	Flow Control	

4.11 Safety Requirements

From a safety perspective, oxycombustion presents two areas of concern; 1) oxygen concentration beyond about 25% by volume which could result in a fire and 2) carbon dioxide leakage into the plant building where it could be concentrated enough to be harmful if breathed.

Oxygen Concentration

Every day the use gaseous oxygen at 21% volumetric concentration is experienced in the air. Oxygen itself is a colorless, odorless, tasteless, nonflammable gas. A potential danger is that many substances which do not burn in air, may burn in an oxygen-enriched atmosphere (> 23.5% O₂). Liquid oxygen is extremely cold (-297 F) and oxygen as a liquid or cold dry gas may cause severe frostbite to the eye or skin.

In addition, higher concentrations of oxygen (25% to 75%) present a risk of inflammation of organic matter in the body. Elevated oxygen levels may result in cough and other pulmonary changes. High concentrations of oxygen, >75% under pressure, causes symptoms of hyperoxia which includes cramps, nausea, dizziness, hyperthermia, ambylopia, respiration difficulties, bradycardia, fainting spells, and convulsions capable of leading to death.

There is no concern about hazard due to breathing should oxygen leak from the supply pipes, the only concern might be combustion but the oxygen would most likely be diluted by the surrounding environment so rapidly no hazard would be presented. Oxygen monitoring is not recommended because oxygen will mix very rapidly with air and it is almost impossible to measure it. The oxygen transport lines should be leak checked periodically at 100 psig and adequate ventilation should be provided in areas where oxygen is being used. In the event of a significant oxygen leak, the action would be to clear the area and let the oxygen vaporize. More information is provided in Section 6.5.

Carbon Dioxide Leakage

The boiler house must be well ventilated. Ambient CO₂ monitoring will be performed continuously on the operating floor areas and in low or confined locations areas, especially where the boiler flues or equipment operate at positive internal gas pressure. Since the unit is balanced draft, the equipment operating at positive pressures consists of the flue from the booster ID fan outlet through the WFGD and SCAH to the inlet of the FD fan and the outlet of the FD Fan through the TAH to the boiler windbox. Only a very short portion of the FD fan inlet flue, the very short FD fan outlet flue to the TAH, the TAH, the flue to the windbox and the windbox itself operate at positive pressure and are inside the building. All of this equipment is on the lower floors and since carbon dioxide would tend to sink, monitoring of these lower floors would be provided.

If the concentration is in excess of the OSHA limit of 5000 ppm, additional ventilation must be provided to disperse the CO₂. CO₂ detectors will be installed at strategic locations, especially low areas where personnel may enter. If CO₂ levels are exceeded, action must be taken to prevent personnel exposure, determine the source, and improve ventilation. If the level cannot be decreased to a safe level, the unit must be shut down and repaired. CO₂ detection should be used before entering confined spaces even though normal shutdown or unit trips must be followed by an air purge of the system.

Gas Side Design Pressures

The gas side design pressures were investigated for this unit. No exact values could be found for the existing design. It was assumed that they were designed according to the standards of 1963 which were to design the furnace to +3/-7 inches of water with no transient consideration. The flues, ducts, and other equipment were assumed to be designed according to the static of the prior fan.

Current standards would require the furnace to be upgraded to withstand a design pressure of ± 10 inches of water and a ± 35 inches of water transient. The modification of the buckstays was assessed and due to tight clearances, it will need to be discussed with the City of Hamilton in Phase 2.

The current standards also recommend the air side of the tubular air heater and the flues from the FD fan to the TAH would also need to be upgraded. This should also be discussed with the

customer as part of Phase 2. The following summarizes the preliminary evaluation of the furnace structural design:

- The buckstay spacing is satisfactory; therefore, there would be no need for additional buckstay levels.
- The tie bars are attached to the wall tubes by welding rather than the current design which utilizes pins and clips. Although there have been some of the older units with this type of construction that have had weld cracking problems, there are others that have operated successfully for over 45 years without problems. As long as they have not had problems with this unit, there is probably no reason to change the construction. It is probably satisfactory for the transient pressure.
- Most of the tie bars are 6" x 1/2" bars with stand-offs on 40" centers. In order to make these good for the -35 in. wg transient pressure, two standoffs would have to be added between each existing pair of stand-offs to meet a maximum allowable spacing of about 14".
- Many of the existing buckstays would be over their allowable stress for the -35 in wg transient condition. Some of these could be made satisfactory by adding overturning posts. However, this could get rather difficult trying to clear existing platforms. Other buckstays would have to have reinforcement added. This could be difficult since most, if not all, of the reinforcement would have to be added to the outer flange which results in very inefficient use of material and presents the problem of clearances. An alternative to reinforcement would be replacement. Based on a quick review, it appears that all of the buckstays could be replaced with heavier members of the same depth, and they would satisfy the transient pressure condition. If this option is chosen, replacing the tie bars may be considered.

Table 4.8 shows the design pressures. Reinforcement of some existing flues and buckstays will be further investigated as part of Phase 2 but is expected and included in the cost estimate.

Table 4.8 - Gas Side Design Pressures

Babcock and Wilcox
Design Parameter Sheet
City of Hamilton, Hamilton, OH (Oxy-Firing)
MCR (250,000 lb/hr Steam Flow), 3% Sulfur Coal, 98% Sulfur Removal
PRELIMINARY PRESSURES (NO DAMPERS OR SEC AIR REHEAT)
February 21, 2006 Rev. D

Equipment
Flue / Duct

Equipment	Material Thk. IN.	Current Air- Operating Pressure (in H2O) [In/Out]	Predicted Oxy- Operating Pressure (in H2O) [In/Out]	Predicted Air- Operating Pressure (in H2O) [In/Out]	Design Pressure Range (in H2O)	Transient Design Pressure (in H2O)	Design Temp (F) [Outlet]	Predicted Original Design Pressure Range (in H2O)
Furnace	Existing	-0.1	-0.1	-0.1	+10/-10	+/-35	No Change	+3/-7
Conv. Pass	Existing	-1.2	-1.2	-1.2	+10/-10	+/-35	No Change	+3/-7
CP→HESP	Existing	-1.2/-1.9	-1.2/-1.9	-1.2/-1.9	+10/-10	+10/-35	No Change	+3/-7
HESP	Existing	-1.9/-3.4	-1.9/-3.4	-1.9/-3.4	+10/-10	+10/-35	No Change	+3/-7
HESP→TAH _{gas}	Existing	-3.4/-3.5	-3.4/-3.5	-3.4/-3.5	+10/-10	+10/-35	No Change	+3/-7
TAH _{gas}	Existing	-3.5/-6.7	-3.5/-6.8	-3.5/-6.6	+10/-10	+10/-35	No Change	+3/-7
TAH _{gas} →ID	Existing	-6.7/-6.9	-6.8/-7.0	-6.6/-6.8	+10/-10	+10/-35	No Change	+3/-7
ID	Existing	-6.9/+5.0	-7.0/+4.9	-6.8/+5.1				
ID→Booster	New 3/16"	n/a	+4.9/+4.2	+5.1/+4.4	+10/-10	+35/-10	No Change	n/a
ID Booster	New 3/16"	n/a	+4.2/+17.2	+4.4/17.4				
Booster→WFGD	New 3/16"	n/a	+17.2/+16.9	+17.4/+17.1	+20.7/-10	+35/-10	422	n/a
WFGD	New 3/16"	n/a	+16.9/+9.3	+17.1/+8.7	+20.4/-10	+35/-10	214	n/a
WFGD→Steam Coil	New 3/16"	n/a	+9.3/+9.3	+8.7/+8.7	+10/-10	+35/-10	214	n/a
Steam Coil	New 3/16"	n/a	+9.3/+7.8	+8.7/+7.2	+10/-10	+35/-10	250	n/a
Steam Coil→FD	New 3/16"	n/a	+7.8/+5.5	n/a	+10/-10	+35/-10	250	n/a
FD	Existing	+0/+9.9	+5.5/+14.7	+0/+9.9				
FD→TAH _{air}	Existing	+9.9/+9.4	+14.7/+14.2	+9.9/+9.4	+17.2/-10	+17.2/-10	250	+9.9/0
TAH _{air}	Existing	+9.4/+6.7	+14.2/+11.0	+9.4/+6.3	+16.7/-10	+16.7/-10	No Change	+9.4/0
TAH _{air} →WB	Existing	+6.7/+5.1	+11.0/+9.4	+6.3/+4.7	+11/-10	+11.0/-10	No Change	+6.72/-0
WB	Existing	+5.1/+3.1	+9.4/+7.4	+4.7/+2.7	+10/-10	+/-35	No Change	+3/-7
WB→PA Fan	Existing	+3.1/+2.9	+7.4/+7.2	+2.7/+2.5	+10/-10	+35/-10	No Change	+3/-7
PA Fan	Existing	+2.9/+33.5	+7.2/+37.8	+2.5/+33.1				
PA Fan→BLR	Existing	+33.5/+2.9	+37.8/+7.2	+33.1/+2.5	+37.8/-10	+37.8/-10	No Change	+3/-7
Bypass→TA to PA	Existing	+9.9/+9.8	+14.0/+13.0	+9.9/+9.8	+14.2/-10	+35/-10	No Change	+9.9/0
Penthouse	Existing / VENT (?)				+2/-2	+35/-35	No Change	+2/0
Roof Tubes	Existing				+10/-10	+35/-35	No Change	+3/-7
Steam Coil→Stack	New / Existing	+5.0/+4.0	+7.8/+7.5	+7.2/+6.2	+10/-10	+35/-10	No Change	-6.9/+4
Stack	Existing						No Change	

***The economics of the situation will push this to be vented although our standards would recommend the documented +/-35 transient stiffening.

** For Original Air Firing this is the flue from the existing ID fan to the stack

* Existing Furnace is believed to be +3/-7 in wg with no transient

* **NO TRANSIENTS WERE ORIGINALLY DESIGNED FOR ON THIS UNIT**

* These design pressures assume the isolated stack issue will be resolved via a bypass or some other means

* Air side of TAH will need to be stiffened even more if not system is not vented or a bypass created

4.12 Controls & Instrumentation

Gas Flow Measuring Device

Accurate air and gas flow measuring devices are available from a variety of vendors such as Phoenix, and Fox and Dole Technical Sales. Gas flow will be controlled from a measurement of boiler output and device will be used to measure the amount of recycle gas flow and the air flows entering the Primary and FD fans for control purposes and to verify our design parameters.

Flue Gas Flow

The boiler and auxiliary equipment are designed to start-up and shutdown using air firing. During this period flue gas flow will be routed to the stack.

During oxy-combustion approximately 70% of the flue gas leaving the unit is recycled back to the boiler. Pure oxygen is mixed with the recycle gas upstream of the PA and FD fan(s). 80% to 90% of the oxygen required for combustion will mix with the flue gas while 10-20% is fed directly to the burners.

The demand for recycle flue gas flow and pure oxygen flow varies with Btu input demand on the boiler. The control system generates the fuel/air demand using a lead/lag philosophy. During load increase, air and/or recycle gas/oxygen demand leads fuel demand. To avoid unacceptable oxygen concentration increases during load increases, recycle gas flow increase and oxygen flow increase can be simultaneous but recycle gas flow cannot lag oxygen flow. During load decrease, air and/or recycle gas/oxygen demand lags fuel demand. To avoid unacceptable oxygen concentration increases during load decrease, recycle gas flow decrease and oxygen flow decrease can be simultaneous but recycle gas flow cannot lead oxygen flow decrease.

During load increases the boiler requires an increase in flue gas recycle and oxygen. The control system generates an increase in the demand prior to increasing fuel flow. In other words, the boiler must receive increased flue gas recycle / oxygen flow prior to combusting the additional coal flow.

To improve the stability of the recycle flue gas/ oxygen flows and the furnace pressure, the CO₂ compression system must be on back pressure control. This will provide a relatively constant flue pressure for the FD and ID fans that is similar to that provided by the stack when on air firing. If the CO₂ compression system attempts to maintain constant mass flow, the recycle flue gas/ oxygen flow control stability will be reduced and could become unstable under certain operating conditions.

Oxygen System Purge

To maintain clean conditions in the oxygen injection lines into the oxygen mixer and to the individual burner oxygen lances, low pressure slightly superheated steam is supplied. Steam flow indication is provided for each burner for proof of purge in the DCS. As oxygen injection begins, the purge steam is gradually removed and as oxygen is decreased purge steam is increased in order to maintain a positive flow in the lines at all times.

The oxygen lances to the burners are retractable and steam purge is continuous when air firing or when a burner is out of service and is reduced during the transition until the flow is fully oxygen (Figure 4.2).

Figure 4.2 – Oxygen Supply to the Burners

The steam purge to the oxygen mixer can be shut down when air firing but a five minute steam purge is required before introducing oxygen into the mixer. Temperature measurements on the oxygen injection grid probes will indicate positive purge flow.

Oxygen Injection Valves

Oxygen from the oxygen supply system is supplied at 8 psig to the boiler injection points and is split into the burner supply headers, one per pulverizer group, with automated shut-off valves, and to the oxygen mixer. Individual burners are supplied from the header corresponding to their pulverizer.

The oxygen mixer injection header shut-off valve can be opened when the boiler O₂ supply pressure from the oxygen supply system is greater than the low pressure trip set-point of the pressure switch. Once the oxygen mixer injection header shut-off valve is opened, the oxygen mixer injection header control valve can be released to control the measured O₂ injection flow to match the O₂ flow demand from the boiler controls.

The burner O₂ injection header shut-off valve can be opened when 1) the boiler O₂ supply pressure from the oxygen supply system is greater than the low pressure trip set-point of the pressure switch and 2) the corresponding pulverizer feeder is started. Until the pulverizer feeder is started, a steam purge is maintained to the header to insure lance cleanliness.

Once the burner O₂ injection header shut-off valve is opened, the header supply control valve controls oxygen flow based on pulverizer demand. As the oxygen flow is increased, the purge steam is decreased.

Transitioning From Air Firing To Oxygen Firing

The boiler is started up in the usual manner on air firing and brought up to approximately 40% stable load. The oxygen supply system is started and ready for service prior to initiating the flue

gas recycle mode (see Figure 4.3). Once stable operation is achieved at minimum coal firing rate (lighters may be left in until the transition is complete as a precaution), the following is the procedure for the transition:

1. With the FD fan inlet air isolation damper is fully open and the flue gas recycle guillotine damper (tight shut-off) is fully opened allowing flue gas to enter the FD and PA fan inlet flue.
2. The FD fan inlet air supply damper is gradually closed drawing more flue gas into the FD fan inlet flue instead of the stack. If the desired recycle flue gas flow is not achieved when the FD Fan inlet air supply damper is fully closed and the recycle guillotine damper is fully open, the stack inlet damper can be gradually closed to force additional flue gas to the FD fan inlet flue. The stack inlet damper may not completely close since flue gas flow must be maintained to the stack at all times. When operating in equilibrium, the flue gas flow to the stack is equal to the sum of the excess oxidant (air and/or oxygen) added, air infiltration, and the products of combustion. The boiler is in full flue gas recycle mode once the FD fan inlet air supply duct damper has fully closed. Boiler output controls the recycle demand. Recycle gas flow can be used to trim furnace absorption. Increasing recycle decreases absorption and vice-versa.

The FD flow is measured and temperature compensated based on the densities of the air and oxygen/recycle gas flow streams. An approximated density vs. air flow transition curve, calibrated during startup, is used during transition between air and oxygen firing modes.

Figure 4.3 – Oxygen Supply to the Recycle Flue

3. Oxygen is injected into the oxygen mixer and into the burners in service to make up for the reducing oxygen content in the flue gas/air mixture. The total oxygen demand is based on the stoichiometric oxygen requirement corresponding to the total BTU input plus the excess oxygen lost to the CO₂ compression system/stack, less the amount of oxygen entering with any air flow. The oxygen flow is trimmed based on the desired versus measured excess oxygen at the economizer outlet.

The total O₂ to the in-service burner lances is a preset percentage of the total oxygen flow. The O₂ flow to an individual burner level is a function of the individual pulverizer demand. Oxygen

is supplied to all burners associated with a given pulverizer. Distribution between burners is pre-tuned during commissioning using manual valves for each burner lance. The total quantity of O₂ to the in-service burners as a percentage of the total oxygen demand is tuned during start-up (lances are designed for up to 20% of full load total oxygen flow). Oxygen to the burners is temporarily increased to a higher flow than set point during the transient to ensure flame stability and returned to its set point once stable.

During the transition, the oxygen flow to the in-service burners is established first and ramped to the higher set point percentage of the total oxygen demand to the unit to ensure flame stability. Once the preset flow is injected in the burners in service, the remaining oxygen required to meet the total oxygen demand is fed to the oxygen mixer. The total oxygen demand is trimmed to maintain the desired excess O₂ leaving the boiler as previously described. The local concentration of oxygen in the recycle flue gas downstream of the mixer must also remain below maximum oxygen concentration limit (23.5%) throughout the system. This is achieved by measuring the local concentration at an appropriate distance downstream of the oxygen mixing location (a grid) and diverting oxygen to the burners if the measured local concentration exceeds about 21 percent there is an alarm.

The demand for total O₂ flow is coordinated between the boiler and the oxygen supply system. The boiler control loop forwards the O₂ flow demand to the oxygen supply system and the oxygen supply system supplies the required O₂ flow.

4. Starting an additional pulverizer to increase boiler load is achieved by adjusting the burner air registers from the cooling to the light-off position. The igniters will then be placed in service on the pulverizer to be started. Oxygen to the in-service burners is ramped to the high set point during the transient to ensure flame stability and decrease the risk of high O₂ concentration in the recycle flow as oxygen is added. PA flow to the pulverizer is increased as quickly as practical when the burner line shutoff valves are opened, which increases recycle flue gas flow. PA flow increase to the pulverizer is controlled by the primary air flow control damper. The ID fan will compensate and the CO₂ compression system will control to backpressure resulting in a temporary decrease in flow to the compression plant.

The total oxygen demand is increased leading fuel flow demand. Oxygen to the Oxygen Mixer may temporarily drop as additional oxygen is diverted to the in-service burners. When the feeder is started, additional oxygen is demanded and the oxygen supply the burner lances coming into service is initiated. The pulverizer is started followed by the feeder and coal flow is increased. This will automatically back the other pulverizers down to maintain heat input. Once stable conditions are achieved at the new total heat input, oxygen to the burners is returned to its normal set point.

5. Load increase is accomplished by increasing the heat input load demand which will increase the recycle flue gas demand and oxygen demand. Again, oxygen to the burners is temporarily increased to its high set point during the transient and returned to normal set point once stable conditions are achieved. Fuel flow increase lags flue gas (and oxygen) flow. The process is opposite for a load decrease; recycle and oxygen decrease lags fuel flow decrease.

6. As boiler load is increased or decreased, some adjustment of the backpressure from the CO₂ compression system will be required to maintain the proper flue gas recycle flow.

Transitioning From Oxygen Firing to Air Firing

The boiler load is reduced to the selected transition load. The oxygen shut off valves to the burner injection headers are closed as each pulverizer is shut down and purge steam is introduced. Flue gas recycle is reduced by:

1. Gradually opening the FD fan inlet air supply duct dampers and the stack inlet damper. As air displaces the flue gas, the oxygen demand decreases reducing the injection rate to the in-service burners and the oxygen mixer in the recycle flue.
2. As the oxygen injection rates are reduced based on the total oxygen demand, the oxygen supply system output will be reduced and excess oxygen will be vented as necessary when the boiler demand is less than the output of the oxygen supply system.
3. Once the FD fan inlet air supply duct damper and the stack inlet damper are fully open, the flue gas recycle guillotine damper is closed and the unit is in air-fired mode.
4. When the oxygen is no longer being demanded, the shut-off valves for the oxygen mixer and to the burners are closed and minimum purge steam is fed to all burners.

4.13 Restoration

Upon completion of the test program, the oxygen lances will be removed from the burners, the recycle flue, steam coil heater, and the recycle and stack control dampers will be removed. This will result in the original gas path with the addition of the WFGD systems. The booster ID fan will continue to be necessary to overcome the pressure loss through the absorber. The control system will be returned to the air firing mode and the oxy-firing control logics will be either disabled or removed.

5.0 FLUE GAS DESULFURIZATION DESIGN SUMMARY

5.1 Background

Based on the composition of the flue gas resulting from oxygen combustion with flue gas recycle, and an assumed composition of the municipal water treatment plant (WWTP) byproduct being used instead of limestone as a reagent (later verified by sample analysis), a mass balance was performed around the absorber tower. The number of absorber recirculation pumps was based on the flow required for 97% removal efficiency of SO₂ and the number of absorber trays was determined by minimizing the power consumption.

General Description

The following is a general description of the Wet Flue Gas Desulfurization (WFGD) system for the City of Hamilton. The waste water treatment plant (WWTP) byproduct based, flue gas scrubbing system is designed to remove 97 % of the SO₂ while the unit is running 3.0 % Sulfur coal, equating to 0.14 lb SO₂/MBtu emissions – which meets the 1.45 lb SO₂/MBtu emissions limit from the flue gas produced over the performance and range conditions contained in the design basis. The WFGD System design for this application incorporates features which have been developed over many years of commercial operation and pilot work in utility flue gas cleaning. Babcock & Wilcox has more than one hundred years of service to the power industry, with extensive experience in utility boiler air pollution control systems and is a recognized leader in FGD.

The proposed system will consist of three major subsystems: the wet absorber system, reagent preparation system and the gypsum dewatering system. Listed below are the major items that comprise the proposed Wet Flue Gas Desulfurization System:

- a. Two (2) 100% stationary transport pumps will transport limestone reagent/WWTP byproduct from a truck to a reagent storage tank. Two (2) 100% reagent feed pumps will be provided to feed the reagent to the one (1) absorber tower.
- b. One (1) 100% absorber tower with integral reaction tank, three (3) installed 100% absorber recycle pumps, three (3) standard spray headers and two (2) trays. SO₂ removal occurs in the absorber.
- c. Two (2) absorber bleed pumps to discharge the absorber blowdown.
- d. Two (2) 100% oxidation air compressors. Oxidation of the absorber slurry will be carried out in-situ with the compressors used to supply air to the absorber reaction tank located integrally to the proposed absorber.
- e. One (1) primary dewatering hydroclone cluster to meet the requirements of the WFGD absorber systems for gypsum slurry primary dewatering.

- f. Two (2) 100% vacuum belt filter systems, complete with associated tanks and pumps to provide 94.5% purity gypsum product. Each filter system is sized to dewater the gypsum slurry produced by the unit.
- g. One (1) secondary hydroclone cluster to purge chlorides and fine particulates from the absorber system.
- h. Motors for equipment supplied.
- i. Instrumentation.
- j. Process piping, valves and fittings.

Design Factors

Before all of the equipment could be designed, several unconventional conditions had to be taken into account while performing the mass balance including the impact of the different gas compositions on sulfur removal, moisture in the flue gas leaving the absorber, and an unconventional reagent.

Gas Composition

The main issue concerned the change in the composition of the flue gas entering the absorber under oxygen combustion with flue gas recycle conditions (as shown in Table 5.1).

Table 5.1 - Differences in Inlet Flue Gas Composition		
Flue Gas Composition (Weight %)		
Constituent	Air-firing	Oxy-firing
CO ₂	18.75	64.34
H ₂ O	4.65	10.41
SO ₂	0.44	0.45
N ₂	71.09	19.42
O ₂	5.05	5.36

As it can be seen, a large portion of the flue gas during oxy-firing is CO₂. This increase in CO₂ concentration will not adversely affect the chemistry in the scrubber due to the fact that SO₂ is a much stronger acid gas. Since there is enough SO₂ in the outlet flue gas to ensure that the partial pressure of CO₂ remains at a safe level, the SO₂ is not hindered in its reaction with the calcium carbonate to form gypsum and CO₂. In contrast, if the SO₂ levels were decreased to the point where the partial pressure of CO₂ would rise above 2 atm, the gypsum forming reaction would shift to the reactants and the SO₂ would cease to react.

Moisture Level Leaving the Absorber

When the nitrogen in the flue gas is replaced with CO₂, there is a concentrating effect which increases the moisture level entering the absorber in the flue gas. When this happens, the increased partial pressure due to moisture also increases the saturation temperature achievable

by evaporation inside the scrubber from 138F in air-firing mode to approximately 175 °F in oxy-firing mode. This increase in moisture and temperature would not only increase the forced draft and primary ‘air’ fan requirements but jeopardizes the ability to dry and convey coal in the existing pulverizers. It was therefore decided that this excess moisture needed to be removed, and a process was developed to dehumidify the flue gas leaving the scrubber in order to reach a reasonable saturation temperature of 138 °F. The resulting percent weight composition of water in the flue gas can be seen in Table 1 as 10.41%, which is down from ~26.5% without the moisture removal process.

Use of Municipal Water Treatment Byproduct

Another unique aspect of this WFGD system is that the reagent will not be the typical limestone, but a calcium-rich byproduct from a wastewater treatment plant located near the City of Hamilton Municipal Power Plant. Samples of the byproduct were obtained and tested at B&W’s research center in Alliance, OH and was found to have the following properties after normalization of the analysis: (91.7 – 92.3% CaCO_3 , ~7.16% MgCO_3 , 1.09% SiO_2 and 0.32 – 1.01% inerts). The analysis also yielded that the slurry will contain ~13% suspended solids, although the byproduct provider stated that the solids could be 25 – 35% of the slurry.

This byproduct has many advantages over the normal limestone reagent. First, it is more reactive than typical limestone – which is 90% minimum CaCO_3 and up to 2.0% MgCO_3 . This increase in reactivity equates to a lower amount of required reagent than would normally be needed to be pumped into the absorber to achieve the desired removal.

Another advantage is that, since limestone will not be used in the WFGD system, the corresponding limestone preparation system (which includes a day bin and feeder, ball mill, ball mill slurry tank and classifier) will not be needed. Instead, this equipment will be replaced with two byproduct transport pumps (one operating, one spare). The byproduct, which will be transported via 5,000 gallon capacity trucks, will be pumped directly to the reagent feed tank, and from there, into the absorber module. The absence of the ball mill reduces the power requirements dramatically, since the ball mill is typically a major source of absorbed power in a WFGD system. In addition, since this reagent already comes in the form of slurry, none of the typical water requirements to the mill (grinding water and cooling water) are needed.

Since Unit 8 is small, relative to a typical utility plant, the scrubber is also very small. Where the absorber vessel for a typical 600 MW unit would be on the order of 50 - 65 ft. in diameter, the absorber for this project is only 16 ft. in diameter as a result of the much lower volume of flue gas / SO_2 . This created some difficulty in obtaining a suitable oxidation air compressor. Some vendors could not provide a compressor that would provide such low volumes of air at such a high discharge pressure (4,847 lbs/h of dry air @ 22.8 psig). Normally, blowers or centrifugal compressors are used in the WFGD application, but screw compressors had to be used for this special circumstance. The small size also affected the cost of the horizontal belt vacuum filter. According to one vendor, the belt filter becomes relatively expensive on a dollar per square foot basis for such low filter feed flows.

Two Flue Gas Compositions

Another unusual design criterion for the absorber tower was the need to accommodate two different compositions of flue gas (as noted in Table 1). The reaction tank in the absorber had to be sized to maintain proper residence and retention times for both types of flue gas. Additionally, during oxygen-firing, the nearly pure oxygen stream is being used to supply the oxidation air for the absorber. This stream is split into three pipes, with each pipe connecting to one of the three oxidation air lances. The connection is made in between the absorber vessel and the air lance isolation valve in order to minimize the amount of lance pipe that will be exposed to both the oxygen and air streams. Accompanying the oxygen piping will be steam purge lines that are connected to the lances prior to the location where the oxygen streams enter. This low pressure steam will purge the lances right before oxygen is introduced in order to eliminate the risk of any particles that may be present in the lance from combusting with the oxygen.

5.2 Sulfur Trioxide Control

Sulfur trioxide (SO_3) is a product of the combustion process which when exposed to condensing water is converted to sulfuric acid (H_2SO_4) that can result in deleterious corrosion of flues and equipment and form a 'blue plume' at the stack. Typically about 1-2% of the SO_2 produced in the boiler is oxidized to SO_3 within the boiler itself. In an oxycombustion process the gas constituents would concentrate by about a factor of 3.5 if none were removed from the recycle stream. In spite of the modest SO_3 removal provided by the WFGD, the potential for sulfuric acid formation and corrosion when firing the high sulfur coal would be high. Even though the test period is relatively brief, Trona injection was selected to provide a means of removing SO_3 .

Trona is a naturally occurring sodium based dry sorbent that is pneumatically conveyed and injected upstream of the particulate control device, in this case the HESP, where it is collected subsequent to absorption of the SO_3 from the flue gas. Successful testing of Trona with air fired boilers burning high sulfur coals has been conducted by American Electric Power. B&W has obtained license rights to apply this new technology and this project would not only provide an opportunity for testing but also characterize it with the oxycombustion gas composition. A very small amount of Trona, about 150 lb/h, is required to achieve at least 80% removal of SO_3 at a minimal cost. The Trona will be injected using a temporary skid for the oxycombustion tests. A more permanent system may be considered by the City of Hamilton following the oxycombustion testing to prevent 'blue plume' when operating with air firing and the WFGD.

5.3 Wet Absorber

As the hot flue gas enters the absorber, it is saturated and cooled. The gas turns upward and is evenly distributed across the absorber cross-section by the absorber tray. Experience has shown that such gas distribution devices are required to optimize SO_2 removal. In addition to providing an even gas flow for the main spray zone, the absorber tray also provides an area of intimate contact between the flue gas and limestone.

The gas leaving the absorber trays passes through two (2) operating standard spray headers (with one (1) spare standard spray header), which are supplied with recirculating slurry from the

absorber's integral reaction tank, and continues through two (2) layers of chevron type mist eliminators for water droplet removal.

Absorber Tower

One (1) 16-foot diameter absorber tower, with an integral reaction tank will be provided. The absorber shell is constructed of Alloy 2205. A complete system of internal supports and external stiffeners brace the walls and shell against pressure and other loads. The weight of the absorber and its internals are supported through the absorber shell and structural system.

The integral reaction tank provides a minimum of 24 hours of solids residence time at MCR load conditions with the specified inlet SO₂ loading. The reaction tank shall be equipped with a drain line and valve to empty the reaction tank during maintenance periods.

As flue gas passes up through the absorber, it is quenched by absorber slurry falling from the sprays/trays and then passes through the perforated absorber trays. The trays, constructed of Alloy 2205, are sectioned into compartments by baffles. This allows for a more even distribution of liquid on top of the trays. The tray level also serves as an excellent platform for slurry spray header/nozzle inspection and maintenance. The tray supports are also constructed of Alloy 2205.

After the flue gas passes through the tray, it encounters the absorber spray zone. The absorber spray headers are fiber-reinforced plastic (FRP) and the supports are constructed of Alloy 2205. The absorber slurry is sprayed from silicon carbide spray nozzles, which are bolted to the spray headers.

The absorber module is equipped with two stages of mist eliminators which remove carryover mist by inertial contact. The primary stage captures large particles and the secondary stage captures wash water droplets and finer particles. The two-stage mist eliminator is kept free of slurry deposits by using a water wash system. Make-up water is directed to both the upstream and downstream faces of the first stage mist eliminator by an array of spray headers and spray nozzles. The upstream face of the second stage mist eliminator is also water washed during operation by an array of spray headers and spray nozzles. The mist eliminators shall be washed sequentially by section to optimize the wash flow rate. The mist eliminator spray headers shall be constructed of FRP. The mist eliminator blades and spray nozzles shall be constructed of polypropylene.

Absorber Agitators

Each integral reaction tank shall be equipped with three (3) Ekato side-entry agitators to provide the required mixing and suspension of solids in the tank at the design conditions. The agitator design is such that operation of the absorber system will not be adversely affected if one of the agitators is out of service. All wetted components of the agitators shall be constructed of a corrosion / abrasion resistant alloy (the agitator shaft shall be constructed of 6% moly stainless steel and the blade shall be constructed of cast duplex stainless steel). The side-entry agitators

employ the use of a flushless mechanical seal. The mechanical seal can be replaced while the absorber is on-line.

Absorber Recirculation Pumps and Piping

The absorber recirculation pumps are used to supply the absorber spray headers with slurry from the absorber reaction tank. Three (3) (two (2) operating, one (1) spare) 100% recirculation pumps will be supplied for each absorber. The absorber recirculation suction and discharge piping shall be 16" diameter fiber reinforced plastic (FRP). A pneumatic actuated isolation knife-gate valve shall be located on the suction piping. Bladder connections shall be located on the discharge piping for on-line pump isolation for maintenance and inspection.

Oxidation Air System

An in-situ oxidation system will provide oxidation air to the absorber reaction tank. This oxidation system forces calcium sulfite ($\text{CaSO}_3 \cdot \frac{1}{2} \text{H}_2\text{O}$), formed by the SO_2 removal process, to be oxidized to calcium sulfate ($\text{CaSO}_4 \cdot 2 \text{H}_2\text{O}$). The two (2) oxidation air compressors (one operating and one installed spare) supply the air used for oxidizing the recirculated slurry. The air is introduced into the reaction tank via the oxidation air lance agitators. The lance agitators provide dispersion of the oxidation air, as well as agitating the absorber slurry.

Make-up Water System

The make-up water will be supplied via stub-ups in each area for the types of water required in each area. The mist eliminator wash, oxidation air humidification, vacuum filter cake wash, vacuum pump seal water, reclaim tank, and flushing and area hose stations will use service water.

The absorber reaction tank make-up will use Reclaim water.

The mist eliminators are automatically washed with 100% service water based on a programmed wash cycle to remove any deposits to assure proper performance. The mist eliminator wash water valve sequencing minimizes the demand for make-up water.

The plant fire protection water system will be tied in for use as the emergency quench water supply source. The emergency quench water piping system will contain a grid of spray nozzles inside the flues to efficiently distribute water to quench the flue gas in the event of loss of power to the absorber recycle pumps.

Absorber Area Sumps

One (1) absorber area sump will be provided. The sump is equipped with two (2) sump pumps (one operating and one spare) and one (1) agitator (sump concrete, linings and civil works are to be provided by others). The absorber area sump collects all drains within the absorber island. The sump pumps discharge back to the absorber tower.

Absorber Operating Philosophy

The absorber operates on a continuous basis. Limestone slurry feed rate to the absorber depends upon absorber slurry pH. Absorber slurry level is controlled by adding reclaim/make-up water to the absorber. Absorber slurry density is controlled by bleeding absorber slurry to the dewatering system. Oxidation air flow is provided at the design compressor capacity.

The most important control parameter in a wet absorber system is pH control. Proper design of this loop provides maximum limestone reagent/WWTP byproduct utilization and the flexibility to adjust to changes in load. The limestone slurry feed flow is adjusted to control absorber slurry pH. Two pH instruments in the absorber bleed pump suction piping are used to measure pH in the absorber. Both signals are sent to the DCS and the operator must select which of the two signals to use for control. The pH signal is compared to a pH set-point. This signal is conditioned via proportional plus integral action. The signal is then used to position the feed slurry valve, which feeds limestone slurry into the absorber reaction tank.

In order to optimize FGD system performance, absorber slurry density is continuously monitored and used to bleed slurry from the reaction tank. The density of the absorber slurry is measured in the absorber bleed pump discharge piping by a density meter. The signal is sent to the control system and compared to a density control limit set point and used to open or close the dewatering hydroclone feed valve. The underflow from the hydroclone is directed to the vacuum belt filters. The overflow from the hydroclone is directed to either the absorber or the purge system.

Two instruments constantly monitor the absorber reaction tank level. The reaction tank level is controlled based on a continuous level control. The reclaim water pump will direct reclaim water to the absorber as necessary to maintain the desired absorber liquid level.

5.4 Reagent System

There is no limestone delivered to the plant. Instead, a limestone reagent/WWTP byproduct will be transported by truck from a nearby wastewater treatment plant to the City of Hamilton Municipal Power Plant. There is one (1) distinct reagent train for the plant. The train consists of two (2) stationary reagent transport pumps, one (1) reagent storage tank with agitator, and two (2) reagent feed pumps.

On average, ten (10) trucks, each holding 5000 gallons of 13% total suspended solids (TSS), will transport the reagent to the plant per weekday. From the trucks, the reagent is pumped by means of 2 x 100% stationary reagent transport pumps (one (1) operating, one (1) spare) to one (1) 62-hour capacity reagent storage tank equipped with agitator. The feed line recycles back to the reagent storage tanks and permits a continuous flow of slurry at various operating loads.

Stationary Sludge Pumps

Two (2) 100% stationary transport pumps (one (1) operating, one (1) spare) will be used to transport the sludge from the trucks to the storage tank. The tubing for the pump inlet will be

sized so that it will accommodate the 3" cam-lock fittings on the truck. The pumps will transport the sludge directly into the top of the storage tank.

Reagent Storage Tank and Pumps

One (1) reagent storage tank is provided to serve the absorber module. The reagent storage tank is equipped with one (1) agitator and has a capacity of 62 hours at 100% performance conditions based on the slurry demands for one unit. A total of two (2) 100% reagent feed pumps (one (1) operating and one (1) installed spare) are provided to feed the reagent slurry to the absorber. The slurry feed rate is continuous and maintained at a high level demand, with recycle back to the reagent storage tank, in order to maintain the necessary flow velocities required to prevent plugging in the piping system.

5.5 Gypsum Dewatering System

The gypsum dewatering area includes the primary dewatering hydroclones, secondary hydroclones, secondary dewatering system (vacuum belt filters) and reclaim water system. The following equipment descriptions indicate the function of the various system components.

Primary Dewatering System

Gypsum slurry from the absorber is fed to the primary dewatering hydroclone cluster by two (2) x 100% capacity absorber bleed pumps (one (1) operating and one (1) spare). The purpose of the primary dewatering hydroclone cluster is to concentrate the solids of the gypsum slurry stream. Flow is added involutively to produce a swirling motion inside the hydroclone. This swirling motion produces a centrifugal force and affects solids separation.

The primary dewatering hydroclone overflow is fed to the secondary hydroclone head pipe / tank. The secondary hydroclone head pipe / tank gravity flows to the reclaim water tank with a small portion feeding the secondary hydroclone. Primary dewatering hydroclone underflow is gravity fed to the vacuum belt filters. Each hydroclone cluster contains two hydroclones (one (1) operating and one (1) spare) and is supplied with a valve network which can isolate flow to either of the hydroclones.

Purge System

In order to limit build-up of chlorides and fine particulate in the absorber system, a small portion of the absorber liquor is purged from the WFGD system. To accomplish this, the purge stream could be taken from the primary hydroclone. However, by purging at this location, the purge stream would contain up to 5% suspended solids. In order to reduce the suspended solids in the purge stream, a secondary hydroclone cluster is employed to process a portion of the primary hydroclone overflow, so that a purge stream of less than 2% suspended solids is produced in the secondary hydroclone overflow.

For the unit, two (2) 100% secondary hydroclone feed pumps and one (1) secondary hydroclone cluster has been provided. As the primary hydroclone overflow gravity flows to the reclaim water

tank, it passes through a head pipe / tank. This head pipe / tank provide positive suction pressure to the secondary hydroclone feed pumps. The secondary hydroclone feed pumps draw off a portion of the slurry and deliver it to the secondary hydroclone cluster consisting of four hydroclones (three (3) operating and one (1) spare). The underflow from the secondary hydroclones is gravity fed to the reclaim water tank, while the overflow flows to the (Owner's) waste water treatment system, with an option for a portion of the overflow to be gravity fed to the reclaim water tank.

Secondary Dewatering System

From the underflow of the primary dewatering hydroclones, the gypsum slurry flows by gravity to either of the two (2) 100% capacity horizontal belt vacuum filters. Each vacuum filter is based on dewatering gypsum from one (1) absorber unit. The gypsum is dewatered via the belt filters to 90% solids and washed to produce a 94.5% purity gypsum cake, which drops onto a gypsum conveying system that transports it to a building set aside for storage of a gypsum pile.

Reclaim Water System

Each vacuum receiver filtrate pump (one per filter) directs the primary filtrate from the belt vacuum filter to the reclaim water tank (1 x 100% capacity per unit). The reclaim water tank also collects reclaim water from each secondary hydroclone head pipe / tank and secondary hydroclone underflow and overflow. The reclaim water is recycled from the four-hour capacity reclaim water tank back to the absorber by either of the two (2) reclaim water pumps (one (1) operating and one (1) spare).

After sizing was completed, the process and instrument diagrams (P&ID's) were developed. Each piece of equipment, valve and pipe that would be needed was included on the drawings along with supplemental information such as pipe diameters and valve instrumentation. With this information, the equipment was located and the relative length of pipes determined. This information was vital for calculating the suction and discharge pressures, total developed head, and net positive suction head (NPSH) for each pump in the system along with the discharge pressure for the oxidation air compressor. This information was used to generate requests for quotes (RFQ's) for the pumps and compressor. Additional RFQ's for the vacuum filters, hydroclones, mist eliminators, strainers, and agitators (for both the absorber and liquid holding tanks) were also generated to determine the sizing and/or quantity of the equipment needed.

Once the quotes were received and evaluated, line and valve lists were produced and P&ID's were finalized. An electric motor list and utility consumption list (service water and air usage) were also composed based on equipment and pipe requirements (cooling water and air, quench water, seal water, flush water etc.)

5.6 Controls & Instrumentation

The control of the FGD will be through the Babcock & Wilcox supplied distributed control system (DCS) along with vendor supplied proprietary control packages. The scope of instrumentation will generally be as depicted on the process & instrumentation diagrams.

Process Control and Monitoring

All analog inputs and outputs will be indicated and trended on the display console.

All motors and automatic valves shall have status indication. All motors and automatic valves shall have Fail to Respond alarms. These shall consist of Failed to Start and Failed to Stop for the motors and Failed to Open and Failed to Close for the automatic valves. They also shall have Command disagreement alarms. These shall consist of motor started and not running and motor stopped and running for the motors and valve commanded open and valve not open and valve commanded closed and not closed for the valves where feedback is available.

Analog transfer functions shall be “bumpless”; at the time of transfer, the output signal shall remain unchanged and then gradually change to the new value.

Where redundant transmitters are used, a DCS algorithm shall be used to select the signal to be used for control. If only one of the signals is of “good” quality, that signal shall be selected. If both are of “good” quality, the Operator shall select the signal that will be used for control (via the operator’s console). A high deviation between input signals shall be alarmed. Criteria for signal selection will be provided in the FGD Operations Manual.

For any equipment with a local control panel, the DCS will receive a local/remote signal from the panel. Starting the equipment will be allowed only from the location selected. Stopping will always be allowed from both the local panel and the DCS. The start will be executed through the DCS, whether local or remote, in order to ensure that the appropriate permissives are met. Stopping will also be through the DCS, but a local stop will also have a contact connected directly to the MCC for equipment requiring a safety stop for personnel protection.

Signal-averaging circuits shall include an algorithm that eliminates any input signals from the average when “bad quality” is detected. High deviation between input signals shall be alarmed.

All three-input auctioneering circuits shall include high and low deviation alarms for each input signal as they compare to the median selected signal.

Absorbent Preparation Control Philosophy

The Absorbent Preparation System is the first part of three in the WFGD functionality. This system is responsible for the entire reagent preparation process. With that in mind, and as mentioned in the General Description of the WFGD, the reagent slurry will either be pumped from a 5000 gallon truck at a frequency of approximately ten times per day or a WWT calcium-rich byproduct will be pumped from a local wastewater treatment facility to the Reagent Feed Tank.

Reagent Feed Tank

The Reagent Feed Tank provides a temporary storage location of reagent slurry prior to transfer to the absorber module. It is a tank in which the liquid slurry level varies from the low level setpoint to the high level setpoint.

The Reagent Feed Tank Agitators keep the slurry solids in suspension.

The tank is equipped with one level transmitter. When the level of the tank reaches a preset low/high level, the operator will be prompted.

The control of the Reagent Feed Tank Agitator is through the DCS. When a tank level is above minimum, the agitator is permitted to start. When the tank level is below minimum the agitator is commanded to stop.

Reagent Feed Pumps

Two Reagent Pumps are provided with one normally operating and one as standby spare. The operating Reagent Feed Pump delivers reagent to the absorber unit. The slurry feed rate is continuous with recycle back to the Reagent Feed Tank, maintaining reagent feed line velocities within the proper range. As the slurry demand changes per the absorber pH control loop, the absorber reagent feed control valve is adjusted, allowing slurry feed to be regulated from the recirculated slurry feed loop to the absorber.

Reagent is pumped from the Reagent Feed Tank by the Reagent Feed Pumps. As the Reagent Feed Tank slurry level reaches the high slurry level, the DCS will alarm the operator. When the high-high slurry level is reached, the Reagent Feed Tank inlet valve for that tank will automatically close.

Start-up and shutdown of the Reagent Feed Pump is manually initiated from the DCS. Operation of the Reagent Feed Pump is interlocked with the Reagent Feed Tank slurry level, i.e., if the Reagent Feed Tank slurry level is equal to or below the minimum level, the Reagent Feed Pump will not be permitted to start.

The DCS will automatically shutdown the Reagent Feed Pump if the Reagent Feed Tank slurry level reaches the low-low liquid trip setpoint.

The Reagent Feed Pump normally stays in operation recirculating slurry to the Reagent Feed Tank and Absorber Tower.

Oxidation Air Blowers

The oxidation air blowers are supplied as a vendor package. The oxidation air blowers will be controlled by this local control package with interface to the DCS. The local control panel supervises the starts and stops and oversees the proper operation of the blowers. Remote

interface will also be provided for remote starting, stopping, and status monitoring of the blower.

Though specifics will vary with the final vendor selected to supply the oxidation air blowers, each blower will be provided with a local microprocessor based control package with the following typical Control Panel Buttons:

ACKNOWLEDGE/RESET
START
STOP
CONTROL MODE LOCAL/REMOTE SWITCH

Oxidation Air Flow

An in-situ forced oxidation system provides oxidation air to the absorber for oxidizing the absorber slurry in the absorber reaction tank.

Oxidation air mass flow is set to a default value. The operator can manually input the mass flow to DCS.

The absorber tower utilizes the oxidation air blower to supply oxidation air to the absorber reaction tank for the purpose of converting the calcium sulfite produced in the sulfur dioxide removal process to calcium sulfate. The flow of oxidation air to the absorber reaction tank can be regulated in response to both boiler load and inlet sulfur dioxide concentration.

Oxidation Air Humidification water is injected into the oxidation air supply header to lower the oxidation air temperature for more efficient transport to the absorbers. The flow rate of the water is such that the air entering the absorber reaction tank is maintained at a saturated state over the entire range of design conditions.

Start-up and shutdown of each oxidation air blower is manually initiated by the operator from the DCS. During start-up, oxidation air flow into the absorber can be started at any time the reaction tank begins to fill with water but before the level reaches a pre-determined elevation. The control of the blowers is handled through a vendor supplied control package.

Spraying water into the discharge piping of the oxidation air blowers cools and saturates the oxidation air. The water flows at a constant flow based on maximum oxidation air flow demand. The humidification water valve opens when the oxidation blower discharge valve opens and closes when the discharge valve closes.

Absorber Recirculation Pump

Three (3) absorber recirculation pumps are provided, two (2) operating and one (1) spare at 100% load design range flue gas condition. Each pump feeds a single absorber spray header. The number of pumps operating is set by boiler load and inlet sulfur dioxide load.

The number of absorber recirculation pumps in service at any one time is set by unit load and absorber inlet flue gas sulfur dioxide concentration.

Start-up and shutdown of each absorber recirculation pump is controlled through the DCS by the operator and is interlocked with the absorber reaction tank liquid level.

One absorber recirculation pump must be in service to maintain operation of the absorber unit.

The Absorber Recirculation Pump runs continuously.

Absorber Agitator

The absorber tower reaction tank utilizes three (3) constant-speed side-entering agitators to maintain solids suspension off-bottom of absorber slurry.

The absorber agitators are operated from the DCS console in the FGD control room. Operation of the agitators shall be interlocked with absorber level. Above minimum level in the reaction tank automatically starts the agitator if auto is selected. Below minimum level will automatically stop the agitator regardless of control mode selected.

Absorber Reaction Tank Level

Two (2) flange-mounted hydrostatic pressure transmitters continuously monitor the gypsum slurry level in the absorber reaction tank. Fresh slurry and reclaim water are continuously fed to the absorber tower which increases the gypsum slurry level in the reaction tank. To maintain the level in the reaction tank between a high level setpoint and a low level setpoint, the slurry is pumped by the Bleed Pumps to the Primary Dewatering area, i.e., the Primary Hydroclone.

The hydrostatic pressure measurement is a differential pressure signal measuring between tank levels (head of slurry) versus atmospheric pressure in order to achieve the reaction tank level reading. When the reaction tank level falls to the low setting, blowdown is recycled back to the absorber from the Reclaim Water Tank.

Note that an absorber bleed pump must be in operation when an absorber is in service. The pump does not shut down upon the absorber reaction tank level reaching the low level setpoint. The pump remains in operation in the recirculation mode. The absorber bleed pumps continue to operate unless the reaction tank level drops below the minimum level to operate the pump. A high reaction tank alarm is generated to warn the operator of an overflow condition.

Absorber Mist Eliminator Wash Pumps

Two (2) Mist Eliminator Wash Pumps are provided, one (1) operating and one (1) stand-by. The Mist Eliminator Wash Pumps supply wash water from the Mist Eliminator Wash Water Tanks to the First and Second Stage Mist Eliminator Spray Headers.

The Mist Eliminator Wash Supply System is designed for continuous operation.

Start-up and Shutdown of the Mist Eliminator Wash Pumps is manually initiated by the Operator from the DCS. Operation of the Mist Eliminator Wash Pumps is interlocked with the Mist Eliminator Wash Water Tank level. Each Mist Eliminator Wash Pump is provided with automatic suction and discharge isolation valves which can be closed to isolate the off-line pump.

Absorber Mist Eliminator

The mist eliminators are automatically washed with service water based on a programmed wash cycle to remove any deposits to assure proper performance.

The mist eliminators are washed according to a wash sequence. Wash sprays clean the lower and upper surfaces of the first stage mist eliminator, the lower surface of the second stage mist eliminator. Each of the three surfaces is washed via wash spray headers. Headers are supplied with on/off isolation valves to divide the mist eliminator surfaces into wash quadrants. The on/off valves control flow to one of the mist eliminator quadrants on each spray level. The system may be operated in either an automatic or manual mode. Washing in quadrants minimizes the instantaneous demand for water.

Absorber Bleed Density

This section covers the control of the Absorber Reaction Tank Bleed as a means of maintaining Absorber Bleed density.

In order to optimize the FGD system performance and maintain the overall system water balance, Absorber Bleed density is continuously monitored. The normal density of the Absorber Bleed Slurry is 20 wt. % total suspended solids.

The density of the recirculated slurry is measured at the common Absorber Bleed Pump discharge line by a density meter. The density meter signal is compared to an operator adjustable density setpoint. A density measurement greater than the setpoint value (%TSS) (indicating a higher density) will increase the Reclaim Water Pump to Absorber Reaction Tank reclaim water flow and slows the Reclaim Water Pump recycle sending reclaim water back to the Reclaim Water Tank. A density measurement less than the setpoint dead band value (%TSS) does just the opposite. The reclaim water flow from the Reclaim Water Pumps will decrease the flow to the Absorber Reaction Tank. Flow increase and decrease is controlled by a modulating control valve on the discharge of the reclaim water recycle line.

Absorber Recirculation Slurry pH

Two (2) pH instruments are located in the common discharge line from the Absorber Bleed Pumps. This provides for a continuous pH sampling loop. Absorber Bleed Slurry pH signals are sent to the DCS where a DCS algorithm compares the signals from both probes. A difference of 10% or more will initiate an ALARM. The operator will SELECT the signal that is determined to be correct. The selected signal will be linearized

through a logarithmic function and compared to an operator adjustable logarithmic characterized pH setpoint. The derivative of Inlet SO₂ is used to modify the reagent demand.

Absorber Dewatering Control Philosophy

One (1) Primary Hydroclone is provided. The purpose of the Primary Hydroclone is to concentrate the solids in the gypsum slurry stream from the Absorber bleed. The primary hydroclone battery consists of multiple cyclones. A swirling motion produces a centrifugal force that encourages solids separation. Each hydroclone cluster is supplied with a valve network which can isolate flow to the individual cyclones. The underflow stream from the dewatering hydroclone is fed by gravity to the secondary dewatering (vacuum filter) systems.

Gypsum slurry is fed to the Primary Hydroclone via the Absorber Bleed Pump. The underflow of the Primary Hydroclone, or the heavier dewatered gypsum, is then sent to secondary dewatering at the vacuum belt filter. Dart valves are supplied at the underflow of the Primary Hydroclone so that any of the two (2) supplied 100% capacity Vacuum Belt Filters can be used for secondary dewatering. The underflow may also be directed to the reclaim water tanks via the dart valves.

The Primary Hydroclone overflow (chlorides and fines) is purged from the system by the Secondary Hydroclone Feed Pumps.

One (1) Secondary Hydroclone Head Pipe is provided. The purpose of the Head Pipe is to ensure that the pump suctions for the Secondary Hydroclone Feed Pumps remain flooded during operation.

The Secondary Hydroclone Head Pipe Level is maintained by the combined action of the Secondary Hydroclone Head Pipe Underflow Isolation Valve, located at the bottom outlet of the pipe, and the Secondary Hydroclone Head Pipe Overflow connection located at the top of the pipe. During periods of operation, the Secondary Hydroclone Head Pipe Isolation Valve will close allowing the pipe to fill with slurry. The pipe will continue to fill until the level reaches the overflow connection. Slurry will then continuously overflow from the Head Pipe ensuring that the Secondary Hydroclone Feed Pumps are provided with a flooded suction line during operation.

Once the Secondary Hydroclone Feed Pump has completed the purge cycle, the pumps are taken out of service and the Secondary Hydroclone Head Pipe Underflow Isolation Valve opened and the pipe is allowed to drain.

Secondary Hydroclone Feed Pumps

Two (2) 100% capacity Absorber Secondary Hydroclone Feed Pumps, one (1) operating and one (1) stand-by, are provided for the Absorber Secondary Hydroclone. Under normal operation, the Absorber Secondary Hydroclone Feed Pumps feed slurry from the Absorber Primary Hydroclone to the Secondary Hydroclone.

Start-up and shutdown of the Absorber Secondary Hydroclone Feed Pump is initiated through the DCS by the operator. Operation of the Absorber Secondary Hydroclone Feed Pump is interlocked with the Secondary Hydroclone Head Pipe level. Each Absorber Secondary Hydroclone Feed Pump is provided with suction and discharge isolation valves which automatically close to isolate the off-line pump. When in automatic, the stand-by pump will automatically START upon a TRIP of the operating pump.

Placement of the stand-by pump in the manual mode will disable the automatic START upon loss of the operating pump. The stand-by pump should only be placed in the manual mode for maintenance purposes.

Vacuum Filters

Two (2) Vacuum Belt Filters are provided, one (1) operating and one (1) spare. The spare filter is common to both Absorbers. The Belt Filters are used to dewater the Primary Hydroclone Underflow slurry to form a gypsum cake containing less than 10% moisture. During the dewatering process, the filter cake is washed to remove chlorides and other dissolved solids from the cake.

The Vacuum Belt Filter is provided with a local control panel to assist in the maintenance of the equipment. The local control panel includes a LOCAL/REMOTE selector that must be in the LOCAL position for the local controls to operate, or in the REMOTE position for the DCS controls to operate. The DCS has similar LOCAL/REMOTE selectors that must agree with the local switches before the LOCAL or REMOTE mode is active. The local control panel also includes an emergency stop (E-STOP) pushbutton. This is a maintained pushbutton and must be manually reset after activation. The E-STOP is active in both the LOCAL and REMOTE modes. Activation of the E-STOP will cause an immediate shutdown of the Vacuum Belt Filter drive components.

Vacuum Filters Auxiliary Skid

One (1) Vacuum Pump is provided per Vacuum Belt Filter. The Vacuum Pump is used to provide vacuum for the Vacuum Belt Filter to remove excess water (filtrate).

The Vacuum Pump startup and shutdown sequence is either a manually initiated operation by the FGD operator from the DCS interface terminals or an automatically initiated operation by the DCS as part of the Belt Filter startup sequence. Operation of the Vacuum Pump is interlocked with the Filtrate Receiver level and the Vacuum Pump Seal Water Flow switch.

Reclaim Water Tank

The Reclaim Water Tank is designed to return the recovered liquids from the primary Filtrate Pump and Service Water to the Absorber Reaction Tank for level control.

The Reclaim Water Tank system is designed for continuous operation. The agitator is started and stopped from the DCS automatically or operator initiated and runs continuously unless low level trips the agitator. The Agitator is permitted to start when the level is above minimum.

Reclaim Water Pumps

Two Reclaim Water Pumps are furnished. One pump operates, while the second pump remains in automatic standby mode, based on absorber area sump level.

Startup of the primary pump is initiated when Reclaim Water Tank Level is high. Shutdown of the primary pump is initiated when Reclaim Water Tank Level is at minimum. Startup of the Standby pump is initiated upon failure of the primary pump or low discharge pressure at the pump.

Chloride Purge / Waste Water Pump

Two (2) Chloride Purge/ Waste Water Pumps are provided with one normally operating and one as standby spare. The operating Chloride Purge Pump delivers product to the Waste Water Treatment System (by CoH).

5.7 Restoration

Since the CoH plans to continue to operate the unit with the wet scrubber after completion of the oxycombustion demonstration, restoration will simply consist of removing the recycle flue and damper, steam coil airheater after the scrubber, and FD and Primary fan air inlet dampers, connecting the scrubber to the existing stack, removing the stack damper, and disabling or removing oxycombustion control logics from the control system (air firing capability will remain available during the demonstration). Burner oxygen lances and oxygen piping will also be removed from the plant. The entire wet scrubber system including ID booster fan, absorber, reagent preparation, dewatering, and controls will remain in service.

6.0 OXYGEN SUPPLY SYSTEM

6.1 Background

In order to operate the City of Hamilton boiler in oxycombustion mode, oxygen volumes ranging from 200 to 500 short tons per day with peak instantaneous flow rates equivalent to 660 short tons per day are required. To obtain the required flow rates, during Phase I of the project, a customized and dedicated Air Separation Unit (ASU) was considered. This section describes the process of oxygen production for the retrofit.

6.2 Air Separation Plant

Process description, plant performance, design basis and cost estimates are detailed in Appendix 5 of this report.

6.3 Liquid Oxygen System

Even though no technical barriers were found for construction and operation of the air separation plant, alternate solutions for the oxygen supply were sought to minimize project cost without jeopardizing the technical value. These options included 1) a smaller ASU complemented by liquid oxygen (LOX), 2) relocation and adaptation of a used ASU, 3) coupling multiple standard ASUs, and 4) full oxygen supply by LOX. Considering the fact that the oxygen requirement for the proposed demonstration is for approximately six months, the most economical option found out is to use LOX.

The liquid OSS, as illustrated in Figure 6.3 is designed to provide gaseous oxygen at the pressures and flow rates specified for the process. LOX will be hauled from various ASU plants in the area by trailers. The OSS will consist of the following:

- 12 insulated 15,000 gallons LOX storage vessels will be installed, with a total capacity of approximately 180,000 gallons (856 tons). These LOX tanks are commercially available.
- Unloading stations to allow up to three transport deliveries simultaneously.
- Steam-heated vaporization equipment engineered to provide greater than 620 tons per day continuous flow rate.
- Pressure control systems designed to maintain process pressure settings at predetermined set points.

6.4 Oxygen Distribution System

The oxygen distribution system (ODS) distributes the gaseous oxygen produced to various points of use. In the proposed oxycombustion technology, oxygen is primarily used to mix with recycled flue gas to generate the oxidant required for combustion. In addition, pure oxygen is also injected into burners and wet FGD system through oxygen lances. To avoid contamination of pure oxygen piping with any impurities, steam will be purged through the oxygen lances. The ODS controls the oxygen flow to the set point as required by the process and also monitors

safety (pressure, temperature, flow) in different branches of the ODS. The ODS is also designed to communicate with the burner management system for safe operation of the boiler.

One of the key challenges in the oxycombustion technology implementation at the CoH is safe and efficient mixing of oxygen with flue gas. Mixing has to be obtained with the available mixing length or real estate. The concentration of oxygen has to be maintained lower than 23.5% on the wall of mixing duct to avoid special attention and materials. Air Liquide has vast experience in handling pure oxygen and also mixing gases with pure oxygen. Air Liquide patented a static mixing device known as Oxynator™ primarily used for mixing oxygen with air for other industrial applications. Numerical simulations were performed with this existing design to validate the applicability for mixing with flue gas. Figure 6.1 shows the computational domain of the numerical simulations performed.

Numerical simulations performed showed that the current design of the Oxynator™ does not give satisfactory mixing performances with the available mixing length. As shown in Figure 6.2, the oxygen injected in the centre of the duct is not dispersed completely before the end of mixing zone. As a result, Air Liquide investigated alternative designs (beyond the scope of this project) and found few designs that can give satisfactory mixing with the given safety and real estate constraints.

Figure 6.1 - Computational domain of numerical simulations performed

Figure 6.2 - Oxygen concentration profiles at various downstream cross sections

6.5 Restoration

Upon completion of the test program, the LOX vessels/OSS, ODS will be removed from the test site. The control system will be returned to the air firing mode and the oxy-firing control logics will be either disabled or removed.

Figure 6.3 - Schematic of the liquid OSS

Figure 6.4 - P&ID of Oxygen Distribution System

7.0 PHASE 2 TEST PLAN

The following describes the test program and testing standards employed for the project. The applicable test codes are listed in Table 7.1.

Gaseous and Particulate Emissions Sampling and Monitoring System

Flue gas will be sampled continuously from the convection pass outlet as well as several other locations. A multi-point sampling grid will be setup to extract the flue gas for analysis. After filtering and drying, CO, CO₂, O₂, and NO_x concentrations will be measured continuously by calibrated analyzers. SO₂ will be monitored through a heated sample line. Existing CEMs will be used to monitor the O₂, SO₂ and NO_x concentrations discharged through the stack.

Fly ash will be sampled upstream and downstream of the hot ESP and downstream of the wet FGD systems isokinetically. Representative samples will be analyzed for unburned carbon content and particulate loading.

Mercury Sampling & Measurement System - Coal-mercury transformation will be quantitatively determined by isokinetic sampling of flue gas according to the Ontario Hydro procedure (modified EPA Method 29) at the convection pass outlet. Cumulative batch samples will be obtained downstream of the hot ESP and the wet FGD systems by radial traverses of an Anderson Universal Stack Sampler. Mercury contents of the collected particulate matter and the absorbed gas phase in impinger solutions, including elemental and oxide fractions are measured by cold vapor atomic absorption spectroscopy (CVAAS). The fate of coal-mercury and its subsequent transformation is determined from material balance calculations.

Temperature and Heat Absorption Measurement Equipment - Furnace exit gas temperatures will be measured by a K-type, water-cooled, high-velocity-thermocouple (HVT) probe.

In-duct O₂ Measurement - Rapid response, in-situ, oxygen sensors will be installed along the primary and secondary flow paths to monitor the oxygen concentration in the gas recycle streams and ensure that safe operation limits are not exceeded. Ambient air quality at various locations around the OSS and in the vicinity of the stack base will be checked regularly with hand-held atmospheric monitors.

Procedures for Establishing Baseline Data

Two baseline performance tests will be conducted to establish the impact of oxygen combustion. They will be conducted after the retrofit to establish the boiler performance with air firing prior to operation with oxygen. Since the plant will be capable of operating up to full load with either air or oxygen, one baseline with air-firing will evaluate performance of the boiler at full load and the other at a selected part load. Air-in leakage and removal efficiency for the hot ESP, air-in leakage for the tubular airheater, performance of the new wet scrubber, and emissions including mercury will be measured. This data will later be compared to data taken during oxygen firing with flue gas recycle to identify changes in heat transfer, emissions, boiler efficiency, scrubber performance, and emissions including mercury removal. In addition,

operating differences between oxygen firing and air firing will be identified, optimized, and assessed to the extent practical.

Table 7.1 - Applicable Test Codes and Procedures	
Test Activity/Measurement	Test Code
Boiler Efficiency (Heat Loss Method)	¹ PTC 4
Air Heater Leakage	PTC 4.3-1968(R1991)
Flue Gas Sampling (O ₂ , CO, NO _x)	PTC 19.10-1981
Raw Coal Sampling	PTC 3.2-1990
Pressure Measurement	PTC 19.2-1987
Temperature Measurement	PTC 19.3-1974(R1986)
Water Flow	PTC 19.5-1972, PTC 6-1976(R1991)
Steam/Water Pressure Drop	PTC 4.1-1964(R1991)
Air/Gas Pressure Drop	PTC 4.1-1964(R1991)
Power Consumption	² IEEE-120
Steam Quality	PTC 19.11-1970
Sodium Concentration	³ ASTM D2791
Silica Concentration	ASTM D859
Steam Caution Conductivity	ASTM D1125
Steam Purity	ASTM D2186 Method C
Unburned Carbon in Refuse	ASTM D-6316
Coal Ultimate Analyses and HHV	ASTM D-3176 & ASTM D-5865
SO ₂ Analysis	⁴ EPA Method 6C
NO _x Analysis	EPA Method 7E
Raw Coal Sampling	PTC 3.2-1990
Particulate Emissions	EPA Method 5B ASME PTC-38
Ash Mineral Composition	ASTM D-2795 & D-3682
Limestone Slurry	Fineness per B&W Wet Sieve Method
Limestone Analysis	ASTM C-25, EPRI N3, EPRI J1, EPRI B6/FC bond lab test procedure, TGA Method, B&W Reactivity Procedure
Sulfuric Acid Mist (SO ₃)	EPA Method 8
Mercury	Modified EPA Method 29
¹ Power Test Code ² Institute of Electrical and Electronic Engineers ³ American Society for Testing and Materials ⁴ Environmental Protection Agency	

Two series of 14 tests each are planned to document the oxy-firing (Table 7.2). The initial ‘A’ series will be performed just after startup commissioning once the controls are tuned for oxygen firing. The initial ‘A’ series can be further broken down into three categories: baseline on air, baseline on oxygen, and operation variation on oxygen. The first category, baseline on air, is to

baseline how the unit operates when firing air as the oxidant. The second category baseline on oxygen is to determine the operational and performance impact that oxy-firing has on a standard boiler design relative to air. This second category additionally becomes the baseline when firing oxygen. The third category contains the parametric studies; varying several of the key operational parameters.

The second ‘B’ series of tests will document the long term impact of firing oxygen. A boiler’s thermal performance can change significantly during the initial several months of operation. This phenomenon is commonly referred to as “seasoning.” It is therefore conceivable that similar “oxygen seasoning” could occur even on a unit that has been conventionally seasoned for many years.

To adequately document the impact of oxy-firing, numerous parameters must be measured. Figure 7.1 is a graphical representation of the boiler and air quality control equipment showing the test data to be collected at the various locations. Table 7.2 is a tabular summary of the general data to be collected, Table 7.3 shows the data by location, and Table 7.4 provides a detailed description of what data will be collected for each test.

Table 7.2 - Test Matrix					
Series	TEST #	TYPE	OXIDANT	DESCRIPTION	LOAD
A: Initial Test Series after Startup	A0	Prelim	Air	Fech Flow & Coal Fineness	High
	A1	Data	Air	High Load Baseline – Air	MCR
	A2	Data	Air	Mid load Baseline – Air	67%
	A3	Data	Air	Low Load Baseline – Air	35%
	A4	Data/Opr	Air/O ₂	Transition	TL (≈35%)
	A5	Data	O ₂	Low load Baseline – O ₂	35%
	A6	Data	O ₂	Mid Load Baseline – O ₂	67%
	A7	Data	O ₂	High Load Baseline – O ₂	MCR
	A8	Data	O ₂	High Load- #1 mill out	MCR
	A9	Data	O ₂	High Load- #2 mill out	MCR
	A10	Data	O ₂	High Load- #3 mill out	MCR
	A11	Data	O ₂	Min Recirculated Gas Flow	MCR
	A12	Data	O ₂	Lower Recirculated Gas Flow	MCR
	A13	Data	O ₂	125 F Scrubber Outlet Temp	MCR
	A14	Data	O ₂	140 F Scrubber Outlet Temp	MCR
B: Aged Test Series after 3 months O ₂ firing	B1	Data	Air	High Load Baseline – Air	MCR
	B2	Data	Air	Mid load Baseline – Air	67%
	B3	Data	Air	Low Load Baseline – Air	35%
	B4	Data/Opr	Air/O ₂	Transition	TL (≈35%)
	B5	Data	O ₂	Low load Baseline – O ₂	35%
	B6	Data	O ₂	Mid Load Baseline – O ₂	67%
	B7	Data	O ₂	High Load Baseline – O ₂	MCR
	B8	Data	O ₂	High Load- #1 mill out	MCR
	B9	Data	O ₂	High Load- #2 mill out	MCR
	B10	Data	O ₂	High Load- #3 mill out	MCR
	B11	Data	O ₂	Min Recirculated Gas Flow	MCR
	B12	Data	O ₂	Lower Recirculated Gas Flow	MCR
	B13	Data	O ₂	125 F Scrubber Outlet Temp	MCR
	B14	Data	O ₂	140 F Scrubber Outlet Temp	MCR

TL = transition load, Fech = Fechheimer probe

Table 7.3 - Summary of Data by Location

	Temp.	Flow	O2	CO	CO2	NOx	SO2	SO3	Hg	Cl	Part. Loading	UBC	Moist.	Grab Sample
0: Ambient Air Inlet	2Pt													
1: FD Fan Inlet (down stream of O2 injection)			PI											
2: AH "Air" Inlet (down stream of O2 injection)	Grid		PI											
3: AH "Air" Outlet (combustion "air")	Grid		Grid		Grid								Calc	
5: Furnace Exit	HVT, 2CP		PI, HVT		HVT									
7: Boiler Outlet (before Trona Injection)	Grid	Calc	Grid	Grid	Grid	Grid	2Pt, MT	MT	MT		MT	MT	MT	
8: HESP Exit (AH Gas Inlet)	Grid	Calc	Grid		Calc				MT	MT	MT	MT	MT	
9: AH Gas Exit (ID Inlet)	Grid	Calc	Grid		Grid		2Pt, MT	MT					MT	
10: ID Booster Exit (WFGD Inlet)	Grid	Calc												
11: WFGD Exit (SCAH Inlet)	Grid	Calc	Grid	Grid	Grid		2Pt, MT	MT					MT	
12: SCAH Exit	Grid													
13: Recycle (FD In)	MT	PI, MT												
14: Stack	MT	MT							MT		MT		MT	
A1: Total - LP Oxygen	PI	PI												
A2: FD Inlet - LP Oxygen	PI													
A3: FD Outlet - LP Oxygen	PI													
A4: Burner - LP Oxygen	PI	PI												
A5: WFGD - HP Oxygen	PI	PI												
TI - Trona		PI												
PC - Pulverized Coal	PI													
Coal - Feeder Inlet														Grab
Slurry - WFGD														Grab
Plant Instrumentation - General	All Available Boiler Side													

2CP = Two color pyrometer

2Pt = Two point grid

Calc = Calculated

Grab = Grab sample of stream

Grid = Multi point grid, continuously monitored

HVT = Manual traverse using a High Velocity Temperature probe

MT = Manual traverse

PI = Plant Instrumentation

8.0 PHASE 2 STATEMENT OF WORK

The objective of the project is to convert and operate a plant of significant size to oxy-firing with flue gas recycle, producing a concentrated CO₂ stream suitable for future processing and permanent storage. The project will take place at a 25 MWe pulverized coal-fired unit located at the City of Hamilton Municipal Electric Plant, Hamilton, Ohio (near Cincinnati). The boiler is a B&W Stirling Power Boiler of 1963 vintage with three pulverizers and six burners that will fire a Waterloo Ohio high sulfur bituminous coal. Once the project is completed, the unit will be restored for air firing with the added benefit of allowing the utility to continue to burn high sulfur (e.g. Ohio) coal on an on-going basis while meeting SO₂ emission limits.

This project will provide extremely valuable data for use in developing and confirming design assumptions and predictions as well as directly applicable experience in controlling and operating an oxy-fired power plant under all typical conditions in a utility dispatch mode. Since the magnitude of the investment inherent in a new large utility plant is on the order of a billion dollars and the associated operational risks would be untenable without prior experience at a significant scale, this project is essential to the commercialization of the oxy-combustion option for carbon management.

Due to the size (economy of scale) and extensive use of original equipment in order to minimize project cost while maximizing the technical value, this project will not reflect the optimization or economics of a large scale new oxy-fired plant with a high efficiency supercritical steam cycle, large air separation unit, and CO₂ compression system. Though the impact on performance can be quantified and taken into account, the age and design of the existing equipment will result in somewhat higher air infiltration than would be achievable with a modern plant design. Performance results obtained at this project scale should permit prediction of the critical performance parameters at modern combustion power plants.

This Project is intended to be accomplished in two Phases. In Phase 1, the process was established, heat and mass balances for various load conditions were made and all equipment was designed (sized). In addition, equipment layout and general arrangement drawings were made (not fabrication details) and equipment lists were assembled. Equipment loads were determined and vendor quotes were obtained for all major equipment. The cost of all Phase 2 activities was estimated and a schedule was generated. This work was performed by The Babcock & Wilcox Company (B&W) located in Barberton, Ohio and American Air Liquide (AL) located in Countryside, Illinois in cooperation with the personnel at the City of Hamilton plant. Almost a million dollars has been spent so far to develop a design that will provide the technical and operational information needed to allow progression to commercial application. Although Phase 1 is nearing completion, addition work to minimize cost and obtain funding is ongoing with the objective of closing the gap between them.

Phase 2 will consist of the completion of the remaining detail engineering, equipment fabrication and procurement, installation, commissioning, testing, and restoration. A Final report will summarize all Phase 2 activities and detailed results of testing. The Phase 2 equipment will be manufactured at a variety of U.S. vendor and fabrication shops under the management of B&W and AL, and by the B&W Construction Company. A Gantt chart showing the schedule for the tasks and subtasks is provided in Section II.C Technical

Approach and Work Plan Definition. The estimated costs are provided in the DOE required budget forms. The estimated cost for major equipment is provided below (Table 8.1).

Table 8.1 - List of Major Equipment, Estimated Cost and Justification (does not include erection, balance of plant (BOP), or consumables (oxygen etc.))		
Equipment	Estimated Cost	Justification
Wet Scrubber	\$6.0M	Without a wet scrubber, use of high sulfur coal will result in excessive sulfur in the boiler under oxy-firing, which could cause severe corrosion. In addition, the plant will be unable to comply with the current emission limit.
New Induced Draft Booster Fan	\$0.4M	Addition of a scrubber will impose several inches of water in additional pressure drop (the existing fan will be inadequate). The ID fan will be located before the scrubber.
Existing Forced Draft Fan	\$0	Phase 1 work determined replacement is not necessary.
Interconnecting Flues	\$1.0M	New flues will be necessary to connect the ID Fan exit to the booster fan, the booster fan to the new scrubber, the new scrubber to the existing stack inlet and to the existing FD Fan inlet (recycle). Dampers will be needed to control air and flue gas to permit both air and oxy firing.
Controls	\$1.1M	This complex process will require some control hardware upgrades and extensive software modifications.
Oxygen Supply System (OSS) & Oxygen Piping to the Boiler	\$1.2M	To provide the oxygen for combustion, an oxygen supply and delivery system must be provided.

Tasks to be Performed

Task 1.0: Project Coordination - Costs associated with coordinating Phase 2 efforts by B&W, AL and the City of Hamilton including meetings with the DOE and other co-funders e.g. OCDO and monthly reports will be collected in this task.

Task 2.0: Boiler Island Detail Engineering - Although all equipment was sized, weights and loads determined, and major structural issues assessed in Phase 1, final structural and stress analysis of equipment components remains to be completed. In this Task, the detailed design of any reinforcement of the boiler buckstays and existing flues required to accommodate the higher gas side pressures and applicable National Fire Protection Association (NFPA) requirements will be completed. The stiffening and support components of the new flues as well as the detailed design of the wet scrubber internal components will also be accomplished. Final purchasing specifications for all vendor supplied equipment will be generated (the control system will be addressed under Task 4).

Subtask 2.1: Structural Review of Boiler & Flues – Due to the increased gas side pressure resulting from the added booster fan, the boiler enclosure and existing flues must be checked and possibly reinforced to satisfy NFPA requirements.

Subtask 2.2: WFGD Absorber Structural Design – Structural design of the major components was completed in Phase 1 in order to determine an accurate cost. These must be reconfirmed and secondary supports must be designed to support component fabrication.

Subtask 2.3: Fans – Final fan specifications will be prepared for procurement.

Task 3.0: Boiler Island Equipment Fabrication Details - In concert with Task 2, the equipment arrangement drawings from Phase 1 will be updated following a final approval by the City of Hamilton. In addition, any drawings needed for the burners or boiler stiffening will be generated and the Phase 1 flue drawings will be reconfirmed. Once this work is completed, component detail drawings for all parts being fabricated for the boiler, flues, and wet scrubber equipment will be developed.

Task 4.0: Control System Detail Design – System Requirement Specifications will be finalized and the control logics will be completed and the final specifications for instruments and controls will be generated for purchasing the control system hardware, software, and all new and replacement instrumentation.

Task 5.0: Oxygen Supply System (OSS) Detail Design - Detailed component design and selection of the oxygen supply equipment and component specifications for all vendor supplied components will be completed in this task. The final design of the oxygen piping and mixing devices will also be completed along with any mixing evaluation required.

Subtask 5.1: Oxygen Supply System Design – The design of the OSS will be finalized and engineering required to support material procurement will be completed.

Subtask 5.2: Design of Oxygen Piping & Supports - The piping requirements between the OSS and the boiler, scrubber and burner front will be finalized and structural supports designed.

Subtask 5.3: Oxygen Distribution System Design: Design of the equipment for oxygen distribution, injection, and mixing with the flue gas will be finalized and confirmed by numerical modeling as needed.

Task 6.0 Equipment Procurement and Fabrication - In this task, all vendor supplied equipment and materials will be either fabricated in B&W or AL manufacturing facilities or procured in accordance with the drawings and specifications generated in Tasks 2 through 5 and delivered to the site.

Subtask 6.1: Boiler, Flues, and Fans

Subtask 6.2: Oxygen Supply and Distribution Equipment

Subtask 6.3: Wet Scrubber, Dewatering, and Reagent Systems

Subtask 6.4: Controls

Task 7.0 Equipment Installation - All new boiler island equipment including burners, fans, flues, over-fire air/oxygen ports will be installed and any necessary modifications to the existing equipment will be completed. In addition, the boiler outlet casing, tubular airheater, and hot electrostatic precipitator casings will be refurbished to minimize air in-leakage. The Foxboro boiler and burner management control system will be upgraded and reprogrammed for air firing with the new wet scrubber and oxygen combustion modes. The OSS and local controls, oxygen piping, oxygen injection and distribution equipment will be installed. The wet scrubber island including reagent preparation, gypsum dewatering systems, and local controls will be installed.

Subtask 7.1: Boiler, Flues, Fans and Controls

Subtask 7.2: Oxygen Supply and Distribution Equipment

Subtask 7.3: Wet FGD, Dewatering, and Reagent Systems

Subtask 7.4: Topical Report: Activities through Installation & Tie-In

Task 8.0 Commissioning, Shakedown and Tuning - B&W and AL will provide commissioning service support to assist the plant personnel in shakedown and commissioning of the equipment and the overall process. Following this process there will be a brief period of tuning to optimize the combustion process to achieve optimum performance for the key parameters (NO_x, CO, unburned carbon, CO₂ purity, etc.) prior to system testing. Controls tuning for key conditions such as transition from air to oxygen while increasing or decreasing load, load changing, various pulverizer combinations, as well as pulverizer, fans, main fuel, and Oxygen Supply System trips will also be accomplished. This activity is expected to require about two to three months.

Task 9.0 Operation and Testing - Following commissioning, shakedown, and tuning, the unit will be operated in the oxy-fired mode for approximately three (3) months. During that period, both operational and performance tests will be made. Operational tests will consider unit startup, load changing and shutdown approaches to understand system dynamics and response and improve on the initial control philosophies. Following baseline testing with air firing and a baseline full load test with oxygen firing, performance testing will occur at defined part loads and conditions to establish both equipment and process performance parameters. These will include steam generator performance, NO_x, SO₂, SO₃, particulate, moisture, and CO₂ in the flue gas at various locations in the process and in the exit stream, wet scrubber performance, scrubber exit temperature control capability, combustion performance, and boiler and plant impacts such as assessment of heat transfer and combustion control and response. Power requirements will be assessed but due to the size of the plant and the configuration of the Oxygen Supply System, they will not be representative of a new large utility reference plant. B&W will utilize its 3-D numerical models for optimum baseline and oxy-combustion conditions to address scale-up in future applications, including alternate boiler designs, different coals, etc.

Task 10.0 Final Report - The Phase 2 Final Report will contain a thorough description of all Phase 2 activities and the results of testing. Testing results will include the matrix of tests conducted, data analyses and plots of key parameters for those tests, equipment and process performance, and flue gas composition produced at various loads.

Task 11.0 Site Restoration - Upon completion of the test program and at a time convenient with the plant, the site restoration requirements for the OSS and boiler island will be carried out. Equipment unnecessary for air firing will either be removed entirely or modified to minimize or eliminate any impact on air firing. The control system will also be returned to air firing logic.

Project Milestones

1. Release of vendor equipment for purchasing
2. Release of all drawings for fabrication
3. Completion of equipment installation and tie-in
4. Submittal of Topical Report on Activities through Installation & Tie-In
5. Completion of commissioning of all equipment with air-firing
6. First operation in full oxygen combustion mode

7. Completion of Transition and tuning in oxygen combustion mode (beginning of test period)
8. Completion of Testing
9. Submittal of draft Final Report

9.0 PHASE 2 COSTS AND SCHEDULE

The primary objective of Phase 1 was to develop the design sufficiently to accurately determine the cost. Due to the first-of-a-kind nature of the project, the design was developed sufficiently to ‘freeze’ the arrangement and essentially finalize the major equipment specifications. Some engineering was expected to remain for Phase 2 including final confirmation of all equipment, fabrication detail drawings, operating instructions, control logics, remaining equipment specifications and purchasing functions.

Prior to beginning Phase 1, a preliminary estimate of the Phase 2 cost and schedule were assembled to support pursuit of external funding.

Preliminary Phase 2 Cost and Schedule

Prior to initiating Phase 1, a preliminary estimate of the costs was made. Air Liquide supplied an estimate of \$15M for an ASU. B&W estimated the wet scrubber system using a ‘rule of thumb’ cost based on \$/kW and added a rough order of magnitude for flues, a fan, and burner modifications of \$12M plus \$3.4M for the test program (including expendables) for a total of about \$31M. The initial schedule for both phases is shown in Figure 9.1.

Figure 9.1 – Preliminary Project Schedule

Final Phase 2 Costs

As described in Section 6.0, Air Liquide designed an Air Separation Unit that resulted in a cost almost double the preliminary estimate. As a result, other approaches to provide oxygen including used equipment and liquid oxygen were considered. In addition, the test plan was more carefully scrutinized to more accurately define how much oxygen would be required and at what timing. After more thorough consideration, liquid oxygen (LOX) supply and storage

was selected. The main challenge was the logistics of delivery of sufficient oxygen to the site. The final cost of the LOX system and the oxygen required for startup, shakedown and testing was just over \$15M. In addition, Air Liquide agreed to provide \$3.0 M of in-kind cost sharing to the project.

Table 9.1 shows the final cost breakdowns for Phase 2. The first row titled Baseline (proposal) totaling \$55.88M are the costs at the time proposals were submitted for funding. Since those costs were substantially higher than the preliminary \$31M estimate, they were studied to identify 1) what reductions could be made without sacrificing the value of the program and 2) why they were so much higher than the preliminary estimate.

Cost Reductions

Since the City of Hamilton was simultaneously considering, how best to meet future emissions limits for the entire plant, replacement of the wet scrubber with a cooler to control gas temperature and humidity only was also considered to reduce cost. Such a cooler would be similar to an absorber tower but would not require as large the WFGD slurry preparation or dewatering. Some limestone would be needed to control pH in the vessel and some dewatering would be required in order to dispose of the significant amount of water with some solids being removed from the process. The concept was that the oxycombustion testing would be carried out burning the current compliance coal. However, since the ability to switch to a lower cost high sulfur coal while reducing SO₂ and particulate emissions was a major attraction to the project for the City of Hamilton, they did not favor this approach. In addition, the magnitude of the cost savings was not expected to bring the total back to the \$30M range so it was not further pursued.

Meanwhile, several areas of reductions in the costs were implemented. Some administrative costs were consolidated and reduced, the cost of the purchased items were reviewed and reductions identified, the six existing burners will be modified rather than replacing them with new burners and the balance of plant and installation costs were adjusted for the oxygen supply system based on the final information. After implementing these cost reductions the total was reduced to \$51.58M, still \$20.5 M higher than the preliminary estimate.

The need to add a full wet scrubber system to prevent undesirable concentration of SO₂ and SO₃ in the recycle loop, cool and dehumidify the recycle gas represents a total of \$28.33 M including the absorber, limestone preparation and dewatering to provide wallboard grade gypsum. The oxygen supply system is \$15.1 M, testing and reporting were \$1.16M and the remaining boiler system including flues, dampers, burner modifications, boiler modifications, Trona, and controls and instrumentation totaled \$6.99 M. The total installation cost was about \$11.3 M and the balance of plant costs totaled \$8.25M. These costs are included in the respective system totals.

Table 9.2 shows the potential funding sources identified at the time proposals were being submitted. The unfunded \$12.33 M presented a significant challenge even if all of the proposals that were submitted were awarded. Work to fill the funding gap progressed and other funding sources were investigated as B&W and Air Liquide awaited selections.

However, when the key proposals were not selected it was decided not to proceed with Phase 2.

Reasons for Higher Cost

Due to the large discrepancy between the preliminary estimate and the final costs, the preliminary estimate was revisited to determine why. Two major contributors to the difference were identified, 1) the wet scrubber system cost and 2) the balance of plant work required.

As Table 9.1 shows, the basic wet scrubber system cost excluding BOP and the integrated cooler totaled \$20.6M which is \$824/kW based on 25MWe. This is almost six times higher than would be expected for a large plant in the 600 MWe for which published installed costs were on the order of \$200/kW all in at the time. Though the preliminary estimate used a much higher cost per kilowatt, it was far from the factor of six. The impact of SO₃ on the system was not fully understood prior to the engineering effort. Though not a major cost, a Trona system was included.

The balance of plant costs of \$8.25M was also significantly underestimated. Buildings and foundations had not been considered initially and the extent of electrical and piping was not well understood and was also significantly underestimated.

Preliminary hopes were to use the existing fans but it was determined after more detailed engineering that an additional ID booster fan was required. In addition, the potential for higher gas side pressure forced reinforcement of the furnace and flues.

These added components also increased the installation costs aggregately resulting in the significantly higher cost.

Table 9.1 – Phase 2 Cost Breakdowns (1000\$)										
CASE	REAGENT PREP	ABSORBER	DEWATERING	FLUES & FANS	BURNERS	OXY PROCESS	OPERATION & TESTING	TRONA	OXYGEN SUPPLY	TOTAL
Baseline (proposal)	\$1,835	\$17,905	\$10,617	\$4,692	\$1,476	\$2,025	\$11,182	\$221	\$5,928	\$55,881
Known Cost Reductions	\$1,695	\$16,782	\$9,850	\$4,366	\$210	\$1,921	\$11,500	\$110	\$5,144	\$51,579
Note 1: The testing cost includes oxygen and a portion of the Air Liquide cost share.										

Table 9.2 – Potential Phase 2 Funding Sources			
SOURCE	POTENTIAL FUNDING Scrubber (\$1000)		COMMENT
DOE	\$15,000		
Ohio Third Frontier	\$8,000 - \$150 = \$7,850		
OCDO	\$5,000		Must burn Ohio high sulfur coal (scrubber required)
Potential Utility Sponsors	\$2,000		Not started until selected
B&W	\$6,400		Cash
Air Liquide	\$3,000		In-kind Cost Share (split between engineering and oxygen supply)
TOTAL	\$39,250		
SHORTFALL	\$12,329		

Final Phase 2 Schedule

The initial start date was late July 2005 but as the Program Timeline in Appendix 1 indicates, work was not initiated until October of 2005. A proposal for funding of Phase 1 was submitted to the Ohio Coal Development Office (OCDO) on March 18, 2005. On April 28, 2005 B&W was notified that the proposal had received very favorable reviews and a presentation at their Technical Advisory Committee (TAC) meeting on May 12, 2005 was requested and made. The result was a recommendation for funding to the Ohio Air Quality Authority (OAQA). On June 14, 2005 the project was approved and the contract negotiation process began. It was initially intended that work would start shortly after OAQA approval since project costs could be counted for 90 days with OCDO approval.

Subsequent to the transfer of OCDO reporting from the Ohio Department of Development to the OAQA on July 1, 2003, the standard grant agreement had been rewritten and some of the modified terms and conditions were unacceptable to B&W. Extended negotiations resulted in a delay in starting Phase 1 until October 2005 when B&W decided to initiate work without coming under contract with OCDO. This four month delay delayed the forecasted completion of Phase 1 until May 2006. Due to longer than expected time to obtain some vendor quotes, the Phase 1 work was actually complete in June 2006.

Following completion of Phase 1, proposals for Phase 2 funding were submitted to the OCDO, the Ohio Third Frontier, and the DOE in hopes of obtaining funding to initiate Phase 2 by November 2006. Unfortunately, the project was not selected for funding was not awarded so Phase 2 can not proceed.

Figure 9.3 shows the proposed Phase 2 schedule based on a November 1, 2006 start date. The project schedule is shown in figure II.C.4. The duration is approximately thirty-six (36) months in total. The schedule for completion of design and installation is estimated at 24 months followed by about six (6) months of commissioning and testing and six (6) months for data reduction and evaluation and completion of the Final Report.

Figure 9.3 – Phase 2 Project Schedule

10.0 CONCLUSIONS

The Phase 1 project to retrofit the City of Hamilton Unit 8 for oxygen combustion was an extremely valuable experience for B&W and Air Liquide. It improved understanding of the technical issues and equipment requirements as well as the challenges and considerations of retrofitting this new technology to an existing plant. The City of Hamilton site offered an excellent opportunity with the desired size and existing equipment configuration for a small-scale demonstration. However, the need for a complete wet scrubbing system and the high relative cost at 25 MWe scale resulted in a much higher project cost than initially anticipated and caused that system to become a disproportionate portion of the total cost.

Although the inability to proceed with the City of Hamilton project Phase 2 was disappointing, B&W and Air Liquide continue to be optimistic about oxygen combustion as both a retrofit and Greenfield solution that provides carbon concentration for sequestration or sale as a byproduct. In fact, recent internal and external economic studies are showing oxycombustion to be very competitive with other forms of CO₂ capture, the most prominent being solvents such as MEA and IGCC. Only oxycombustion and Amine scrubbing provide can be retrofit to the existing fleet, about half of which is pulverized coal fired. It is also our expectation that as IGCC develops and costs, performance, reliability, and emissions becomes available from field demonstrations in combination with more accurate estimates of the cost of a carbon-capture ready IGCC plant (water shift reactor, hydrogen burning gas turbine etc.), pulverized coal with oxycombustion will become even more attractive for new installations as well.

In the absence of Phase 2 of this project, B&W and Air Liquide continue to pursue oxycombustion. For several months SaskPower has been studying solvents and oxycombustion as carbon capture options for a new 300 MWe net power plant planned in Saskatchewan, Canada. On October 30, 2006 they announced a decision to build a near-zero emissions oxycombustion power plant with B&W and Air Liquide that will capture about 8000 tonnes of CO₂ per day.

In support of that and other projects B&W and Air Liquide are in the process of converting B&W's existing 100 MBtu/h (30MWth or 10MWe equivalent) Clean Environment Development Facility (CEDF), into an oxygen combustion test facility (OCTF). The facility was built in 1994 with funding from the DOE, OCDO, and B&W as a combustion facility for burner development and specially designed to also serve as a test facility for a variety of environmental control devices including ESP, baghouse, dry and wet scrubbing. The modifications in progress will permit testing of a wide range of fuels, and equipment combinations under oxygen combustion conditions at a 10 MWe equivalent scale and is expected to be in service summer of 2007.

11.0 DRAWINGS

The following drawings are contained in appendix 6:

864503-SSP-006 Rev. B marked Plot Plan	554984E-01
P&ID-01-0	554985E-01
F-101	554986E-01
F-501	554987E-01
20466F-3	554988E-01
EP-601-A	554989E-01
EP-602-A	554990E-01
EP-2100-B	554991E-01
78725J-01	554992E-01
78726J-01	554993E-01
78727J-01	554994E-01
81817J-00	555210E-01
81818J-00	556466E-01
81819J-00	3777-M02-0
81820J-00	AMT01E-00
554977E-00	AMT02E-00
554978E-00	AMT03E-00
554979E-00	377-M05-2-A
554980E-01	377-M05-1-A
554981E-01	
554982E-01	
554983E-01	

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APPENDIX 1: PROGRAM TIMELINE October 2005 – October 2006

The following is a chronological summary of project activities and progress by month.

October 2005: An internal B&W kick-off meeting was held on October 6 and efforts began to identify the required resources with the key difficulty being a scrubber process engineer. A meeting was held with OCDO on October 11 to discuss the grant agreement terms and documentation needed from B&W. The documents (statement of work, schedule, costs by month etc.) were updated. A site kick-off meeting was scheduled for November 9 to gather information needed to begin design. A visit was made to the site on October 11 to walk down the unit in preparation for performance testing which was scheduled for the week of November 14.

November 2005: A site Kick-off meeting was held on November 9th. Attendees included several people from B&W representing performance engineering, controls, scrubber, graphics, our local sales representative, Contract Manager, Technical Manager, and Project Manager, a representative from Air Liquide, a representative of OCDO, and from the City of Hamilton; the Deputy City Manager, Electric Director, the Electric Production Superintendent, and various plant personnel. The meeting went well and the City expressed much interest in the project and committed to provide the technical information to support Phase 1.

On November 30th a visit was made to the plant to establish equipment locations and verify the location of existing equipment and underground obstructions. Resource allocation was completed with the most difficult being for the FGD systems.

Process Engineering is progressing well. Initial mass balances for the oxycombustion mode were completed. The results of the performance testing completed at month's end will be used to finalize the heat and mass balances and boiler performance predictions.

Boiler performance testing was completed at month's end.

December 2005: Contact was made with a firm to provide a system to control SO₃ to avoid corrosion in the airheater and flues after the scrubber. Burner design was initiated. Stack dispersion modeling was initiated. The oxygen firing combustion program was revised to include Argon in the gas species output and checking is in progress. The impact on boiler performance with oxygen combustion was established. The scrubber island equipment sizing was essentially completed. Arrangement of equipment continued. Burner selection and design, and review of the stack and dilution requirements were in progress.

Efforts to further involve Air Liquide were initiated. Data reduction of the boiler performance test data began.

January 2006: A visit was made to the site on January 18 to obtain measurements and information for the equipment arrangement drawings. A status meeting was held on January 31 at the site. Air Liquide attended the meeting, though they just started work on the Air Separation Unit at mid-month. During work on another oxy-combustion project, an error in the heat and mass balance was discovered that also impacts this project. By month's end a solution was developed and equipment sizing was again in progress.

Photo 1: New Ash Silo and its Control room

Engineering tasks are being completed to support the initial equipment arrangement drawings (proposal type). One complication is that the City of Hamilton recently built a new ash removal system and silo at the site and installed the ash silo in the area B&W expected to use for the wet scrubber (see Photo 1). Equipment arrangement drawings are expected to be finalized by the end of March. The existing boiler fans were reviewed and an additional ID booster fan was found to be needed. A review of damper arrangement and control requirements as related to furnace implosion protection was initiated. Equipment sizing was essentially completed as was equipment layout. Requests for Quotes were in progress. Secondary issues including a Trona system for SO₃, increasing gas temperature to the windbox and final burner selection continued. A meeting with AEP to begin development of the Trona system was arranged. Design work was expected to be completed in February and A/E involvement initiated. A preliminary cost for Phase 2 was targeted for March/April.

The preliminary stack dispersion study using EPA software was completed to evaluate dispersions of the effluent from the stack with a wet scrubber in the system. Burner selection was in progress.

February 2006: Lewis Perry Associates was provided information and a scope description for quote prior to initiating their work, which was expected to start early in March. Discussions with the City of Hamilton on emissions were held February 22 to assess how to approach the EPA early in Phase 2. The SO₃ system vendor was provided information to begin understanding the requirements for a system and a meeting is being arranged. A preliminary scheme for oxygen control and burner management was developed and a meeting with Air Liquide was planned for March 6. Arrangements were made to test the unit for particulate and SO₃ in May while burning their compliance coal. The mass balance issue reported last month was resolved and H&M balance work completed. Based on the final mass balances, the process design was completed (the design is frozen).

Air Liquide work on the air separation unit continued.

The boiler and scrubber design effort was approximately 70% completed; the balance of plant effort was just starting. Process design was completed, and preliminary P&IDs for the scrubber were released and commented upon. Release of the final P&IDs for the scrubber was expected in mid-March. A preliminary scheme for oxygen control and burner management was developed, and a controls meeting with Air Liquide was planned for March 6. Boiler control work was just starting. Functional engineering work on the boiler and scrubber should be expected to be completed in March, and the control and instrument work on both systems were expected to be finished in April. The cost estimate was expected to be initiated in April.

March 2006: Meetings were held at the City of Hamilton site on March 16 to discuss the wet scrubber and on March 22 with Lewis Perry Associates and AEP to discuss equipment arrangement and the SO₃ system requirements. A meeting was held in Barberton with Air Liquide on March 6 to discuss controls. Lewis Perry Associates' work is underway. B&W engineering will be completed in April, with the possible exception of oxygen-combustion control philosophy. By mid-May the balance of plant engineering and the cost estimate should be completed. Process engineering was completed. Air Liquide work on the air separation unit continued.

A meeting with City of Hamilton personnel on March 16 resulted in a decision to move the dewatering building and to add a new single story electrical building. These decisions have allowed us to complete the scrubber layout; in turn Lewis Perry Associates were released to start balance of plant engineering for the scrubber building. The dewatering building layout is advancing. B&W Mechanical Engineering completed sizing of the scrubber alloy plate and internal beams; this will allow us to finish and issue the final RFQ for the scrubber fabrication.

On March 16 B&W met with the vendor for the SO₃ system in their offices. A vendor representative visited the site on March 22 to assess the arrangement and they were asked to design and quote the system.

Scrubber and boiler P&IDs were completed. Interfaces with flues, fans and oxygen piping were initiated. A meeting was held with Air Liquide on March 6 and they subsequently

provided a first draft P&ID for the oxygen system and B&W reviewed the control requirements. A composite P&ID was in progress which will be used to develop the control and trip philosophies by mid-April. Once drafted, a meeting will be arranged with Air Liquide to finalize controls for this Phase and proceed to get a controls cost estimate from Foxboro for Phase 2.

Lewis Perry Associates was provided information and a scope description; they visited the site on March 22 and have initiated their work. They had already begun electrical design, the meeting and subsequent walk down were very beneficial.

During the March 22 visit, additional test ports were located in preparation for testing scheduled for May. Requests for Quotes for most equipment have been sent out.

April 2006: Meetings were held at the City of Hamilton site on April 17 to inspect the hot electrostatic precipitator relative to ability to inject Trona. A meeting was scheduled in Barberton with Air Liquide on May 4 to finalize controls. Lewis Perry Associates' work continues on schedule for completion in May. B&W engineering was completed and cost estimates are in progress with completion expected in late May.

Air Liquide provided a first cost for an air separation unit designed to provide full load capability continuously. It was substantially higher than the initial expectation. To reduce the cost the operating requirement were reduced to 14 hours at full load dropping to 50% load at night or perhaps 33% and Air Liquide engineering was requested be creative in reducing the cost.

Equipment arrangement drawings for the dewatering building were sent to City of Hamilton for comment. System design reports for the boiler island and the wet scrubber are in progress. All equipment is out for quotes and an engineering estimate for Phase 2 will be initiated first week in May.

The Phase 2 test matrix was drafted and a cost estimate was initiated. A textual description of the test plan was also initiated.

Major system descriptions and boiler performance reports were drafted and will be finalized in May.

May 2006: Lewis Perry Associates' work was completed. B&W engineering was completed and cost estimates were completed. The total B&W portion of the costs was more than twice the \$15M target so work to find ways to significantly reduce costs to approach that number were initiated. However, consideration of the basis for the initial target has indicated some large scope items (such as BOP) were not considered and the wet scrubber was based on a 'typical \$/kW' which resulted in significant underestimate for this small unit.

Air Liquide completed their cost reduction effort and was able to significantly reduce the cost of oxygen supply by providing liquid oxygen instead of building an air separation plant on site. As mentioned last month, the operating capability was reduced from full load capability round the clock seven days per week to 14 hours at full load during the five week days dropping to 50% load at night and operating with air on weekends. This approach made it possible for the oxygen supply cost to fall at about the original target of \$15M.

The final controls meeting was held with Air Liquide on May 4 and it is believed all major issues are understood (some minor items will need to be worked out in Phase 2). All equipment design, P&ID and Arrangement drawing work was completed. The system descriptions were drafted and reviewed.

The test matrix for Phase 2 was reviewed and text written describing the plan for the Final Report. The cost estimate was also completed. B&W construction was provided information for cost estimating.

June 2006: Engineering and cost estimating are essentially completed. Some drawing cleanup and report cleanup remains. The total costs for B&W's supply are 80% higher than the initial project guesstimate anticipated. The discrepancy is now understood but it is also believed that significant cost reduction is possible and an effort has been initiated to do so. In

addition, extension of the project to the end of November has been requested to allow reporting to include the cost reduction effort. A check and scrub of the costs is in progress and documents and drawings for a Final report are being updated.

Hot electrostatic precipitator efficiency and airheater heat balance baseline testing was completed at month's end.

Assembly of the Final Report for Phase 1 to date will begin when time permits. Proposals are being written to obtain funding for Phase 2. A proposal was submitted to DOE on June 19 for \$15M of Phase 2 funding. On June 21 a proposal was submitted to Ohio Third Frontier through Energy Industries of Ohio (EIO) for \$8M for Phase 2. It was through EIO because a Letter of Intent had to be submitted by June 10 and B&W had not submitted one but EIO had and the proposal they had planned to submit did not work out so ours was substituted. A proposal for \$5M to OCDO for Phase 2 was written and will be submitted by July 10. We will continue to pursue additional funding sources over the summer in hopes of closing the gap between cost and funding from both ends.

July 2006: Final costs have been reviewed and the numbers are being verified but significant reductions unrelated to the scrubber have been found. Cost reduction options being pursued include 1) cost reductions on the scrubber equipment and construction, 2) identifying the reagent prep and dewatering systems and several items included at City of Hamilton's request to approach the City of Hamilton to fund, and 3) attempting to design an inexpensive cooler for the oxycombustion testing and run with compliance coal. The third option would not provide the City with sulfur removal capability and may only be tolerable if they choose to build a combined scrubber for Units 8 and 9 and we would connect to it once our testing was completed. Efforts to firm up costs for these options are underway and a meeting with City of Hamilton to discuss options is being scheduled for late August. Sufficient potential cost reductions and funding sources appear to be available to make the project viable if they can all be realized. A proposal was submitted to OCDO on July 7 for \$5M of Phase 2 funding. This is the third proposal submitted in the past month totaling \$28M in potential funding (not including B&W and AL cost shares). Additional options are being explored including the State of Illinois.

August 2006: Final costs have been broken down into considerable detail and known cost reductions applied reducing the total from \$55.9M to \$49.4M. Additional cost reductions are possible but are not expected to amount to more than \$2M without significantly changing the design. On August 25 a meeting was held with City of Hamilton and they expressed continued interest but their primary desire is to have a scrubber for Unit 8 and they indicated no interest in a cooler with compliance coal approach so that work was discontinued. They were not discouraged by the higher cost but committed to try to help find a way to bridge the gap between cost and potential funding, including cost sharing. A review of the Phase 1 design with City of Hamilton is being scheduled of the week of September 18.

Proposals for a potential of \$37.95M in funding if Ohio coal is used, \$32.95M if not, have been submitted (includes B&W and AL contributions). Additional funding sources are being sought and a project summary has been provided to City of Hamilton to help organize a Utility sponsor consortium. Due to testing complications and the relatively small amount of potential funding, it was decided not to submit a proposal to Illinois at this time.

September 2006: Unfortunately, City of Hamilton was unable to meet that week of September 18 so a notebook containing design reports, performance predictions, power and electrical requirements and all drawings was prepared and sent on September 28. If they have further questions, a follow on meeting may be scheduled. Based on an announcement that selections have been made (no projects were mentioned) by the Ohio Third Frontier and we were not notified, it appears we will not receive that funding of almost \$8M. OCDO has finally posted their new solicitation and have sent us a letter indicating we will be required to resubmit our proposal again with comments on their new terms and conditions.

October 2006: Final report writing was initiated. On October 23, NETL announced the selections for the solicitation we had submitted our \$15M proposal for and this project was not among them. It was subsequently confirmed that the Phase 2 project was not selected.

**CITY OF HAMILTON
UNIT #8
TEST REPORT**

For

BASELINE TEST DATA

Prepared By: Douglas J. DeVault
Principal Engineer
Babcock & Wilcox Co.

EXECUTIVE SUMMARY

Tests were conducted on this unit during the week of November 28, 2005. The purpose of these tests was to establish boiler baseline values for an upcoming demonstration of firing a high sulfur bituminous coal with oxygen as the oxidant and recycled flue gas.

A total of two tests were conducted during this test program. The first test was conducted at reduced load condition. The second test was conducted at higher load but not at full load. Full load could not be conducted due to large swings in feed water flow and exceeding plant NOx limitations. The results of the tests were used to establish expected boiler operating conditions on air firing and oxygen firing. Table A2.1 contains a description of the tests conducted and a summary of the pertinent performance information.

TABLE A2.1 – PERFORMANCE SUMMARY

Description	Test 1	Test 2
Date	11/30/05	11/30/05
Time	7:40-8:45	14:40-16:49
Gross Megawatts	14	22
HP Steam Flow, klb/h	121.4	194.6
Sec. SH Spray Flow, klb/h	10.2	19.8
Main Steam Temperature	931	940
Furnace Exit Gas Temperature (MHVT) F	1550	1913
Gas Temperature Lvg Blr, F	581	595
Gas Temperature Lvg AH, F	324	343
Excess Air Lvg Blr, %	68.2	32.0
Excess O ₂ Lvg Blr, %	8.64	5.19
Excess O ₂ Lvg HESP, %	9.86	6.62
Average O ₂ Lvg AH, %	10.29	6.95
Average CO Lvg Blr, ppm	20.96	61.54
Average NOx Lvg Blr, ppm	244.00	200.93
Average NOx Lvg Blr, lb/MBtu	0.50	0.32

DISCUSSION OF RESULTS

All tests were conducted, and all operational performance of the unit was determined, in accordance with the guidelines set forth in the following ASME and EPA publications:

ASME Power Test Code PTC.4.1-1967, Steam Generators

ASME Power Test Code PTC 4.3-1968, Air Heaters

ASME Instruments and Apparatus, PTC 19.10-1985, Part 10 Flue & Exhaust Gas Analysis

EPA Method 7E, Determination of Nitrogen Oxide Emissions (Instrumental Analyzer Procedure)

Air and gas weights were calculated stoichiometrically, using measured O₂ and fuel analysis in accordance with PTC 19.10. Table A2.2 contains a summary of the actual fuel analyses obtained for these tests.

NO_x in pounds per million Btu was calculated using the EPA F factor derived from the following equations:

$$F_x = \frac{10^6 \times (1.53 \times C + 3.64 \times H + 0.57 \times S + 0.14 \times N_2 - 0.46 \times O_2)}{HHV}$$

$$NO_x lb / MBtu = NO_x ppm \times 1.194 \times 10^{-7} \times F_x \times \left(\frac{20.95}{20.95 - O_2} \right)$$

Where:

C = % Carbon in Fuel

H = % Hydrogen in Fuel

S = % Sulfur in Fuel

N₂ = % Nitrogen in Fuel

O₂ = % Oxygen in Fuel For first equation

HHV = Higher Heating Value of Fuel

O₂ = % O₂ in flue gas at the NO_x measurement location For the second equation

TABLE A2.2 – Fuel Analysis		
Constituent	As Received	Dry
HHV Btu/lb	11262	12169
Carbon, %	64.87	70.10
Hydrogen, %	4.20	4.54
Nitrogen, %	1.08	1.17
Sulfur, %	0.79	0.86
Oxygen, %	5.68	6.12
Moisture, %	7.46	N/A
Ash, %	15.92	17.21
Fixed Carbon, %	47.86	51.71

Volatile matter, %	28.76	31.08
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Temporary test instrumentation consisted of gas analysis measurements at the boiler gas outlet flue, the hot precipitator outlet flue, and the airheater outlet flue. These measurements were obtained from side to side gas grids consisting of multi-point probes. A High Velocity Thermocouple (HVT) probe was used for measuring Furnace Exit Gas Temperature (FEGT). All other data was obtained using permanent plant instrumentation. Flue gas analysis data was obtained at ten second intervals and averaged and stored to disc every five minutes. Temperature data was obtained at one minute intervals and averaged and stored to disc every five minutes. All plant data was obtained from the plant control system at one minute intervals.

Unburned carbon samples were obtained from the precipitator hoppers. Analysis showed the total unburned carbon to be 12.52%. Carbon in Ash values were corrected for carbonates by multiplying the percent CO₂ in the ash by the ratio of the molecular weight of Carbon (12) to the molecular weight of Carbon Dioxide (44) and subtracting this value from the Carbon in the ash. The carbon in ash corrected for carbonates was 12.46%.

FEGT was determined by using a 12' HVT probe to traverse the furnace exit plane. One observation door on each side of the unit was used for traversing. Although this was not as many points for traversing as desired, it was considered adequate for this testing.

All of the raw data is stored on disc and available upon request. Appendix A contains a summary printout of the combustion and efficiency calculations, and boiler performance. Appendix B contains a copy of our analyzer calibration sheets and a spread sheet for each test showing the corrections made to the data based on instrument drift. Appendix C contains all of the coal and ash analysis reports.

CONCLUSIONS

Testing went smoothly and no significant items of concern were encountered during the testing. The Unburned Carbon analyses yielded an average baseline Carbon in Ash value of 12.46 %. In a utility boiler this is generally considered very high, however, it has been B&W's experience that small furnaces such as this tend to run colder than the a large ones and this unburned carbon is in line with past experience. There is sufficient doubt, however, as to the accuracy of the number and it may be due to taking ash samples from the precipitator hoppers which may not be a true representation of actual performance. In subsequent testing fly ash samples will be taken by isokinetic sampling for a true representation. Measured O₂ was also higher at the boiler outlet. A heat balance around the airheater would tell us if all this air is going through the burners or is infiltrating through leaks in the casing.

I would like to take this opportunity to thank the people at The City of Hamilton for their cooperation in conducting these tests. It was a pleasure working with all of them.

LOAD IDENT	002	T2	T1
RUN NO.	1	2	3
UNITS			
LOAD	Base	22 MW	14 MW

STEAM AND WATER FLOWS

H.P. STEAM FLOW	MLB/H	250.00	194.64	121.45
SH SPRAY		26.39	19.80	10.17

AIR AND GAS FLOWS

XS AIR LVG ECO	PERCENT	27.0	45.0	86.6
XS AIR TO BURNERS		25.0	30.1	66.3

GAS LVG FURNACE	MLB/H	338.1	296.1	241.4
GAS ENT AIR HEATERS		340.6	325.4	267.8
AIR LVG AIR HEATERS		312.5	273.9	227.0

GAS OR AIR TEMPERING		5.0	0.0	0.0
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FLUID TEMPERATURES DEG F

SEC SH OUTLET		900.	940.	931.
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PRI-1 SH OUT		879.	879.	862.
ATTEMP-1 OUT		687.	687.	702.

ECO WATER IN		350.	324.	306.
SH SPRAY WATER		350.	324.	306.
RH-1 SPRAY WATER		350.	324.	306.

DRUM SAT. TEMP		539.	533.	530.
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PRESSURES PSIG

DRUM		940.	896.	870.
SEC SH OUT		875.	855.	855.

AIR AND GAS TEMPS DEG F

FURNACE EXIT (MEAS) MHVT		1944	1913	1550
GAS LVG ECO DEG F		570.	595.	581.
GAS LVG AIR HEATERS		325.	343.	324.
AVG AIR ENT AIR HEATERS		80.	60.	60.
AVG AIR LVG AIR HEATERS		469.	380.	381.

LOAD IDENT	002	T2	T1
RUN NO.	1	2	3
UNITS			
LOAD	Base	22 MW	14 MW
ABSORPTIONS	MKB		
TOTAL SH	87.6	72.4	42.7
PRI-1 SH	54.8	42.8	26.1
SEC SH	32.8	29.6	16.6
TOTAL OUTPUT	282.4	229.7	145.0
BURN ZONE HT REL RATE	MB/FT2-H	198.7	166.4 109.4
SUMMATION OF LOSSES	PERCENT		
DRY GAS	5.87	7.60	9.06
WATER FROM FUEL		4.31	4.74 4.71
MOISTURE IN AIR		0.14	0.18 0.22
UNB COMBUSTIBLE		0.30	2.92 2.92
RADIATION	1.04	1.28	2.03
UNACC AND MFRS MARGIN		1.50	0.50 0.50
TOTAL LOSSES	PERCENT	13.16	17.22 19.43
EFFICIENCY	PERCENT	86.84	82.78 80.57

PERFORMANCE ANALYSIS SUMMARY

DATE 02/16/06

LD ID	LOAD MW	HP STM FLOW	SH SPRAY	XAIR ECO %	DATE	TIME STRT	TIME END	BURNERS OUT
2	Base	250	26.4	27	E. KENT	UCKY		
T2	22 MW	194.6	19.8	45	11/30/2005	1440	1649	
T1	14 MW	121.4	10.2	86.5	11/30/2005	740	845	

A/E VALUES

LD ID	ECON	PRI-1	PRI-2	SEC	SEC-1	SEC-2	BLR BANK	TOT SH	ECO-1	ECO-2
2	0	1.172	0	0.927	0	0	1.045	1.067	0	0
T2	0	1.111	0	1.047	0	0	1.115	1.084	0	0
T1	0	1.09	0	0.983	0	0	0.904	1.046	0	0

KF VALUES

LD ID	ECO-1	ECO-2	ECO-3	PRI-1	PRI-2	SEC	SEC-1	SEC-2	BBNK-1	BBNK-2
2	0	0	0	1.44	0	0.86	0	0	1.92	0
T2	0	0	0	1.07	0	0.91	0	0	1.45	0
T1	0	0	0	1.27	0	1	0	0	1.25	0

GAS TEMPERATURES F MHVT

LD ID	ENT ACT	PLATENS EXP	FURN ACT	EXIT EXP	ENT ACT	GEN BK EXP	LVG ACT	GEN BK EXP	LVG L ACT	UNIT EXP-KF
2	0	0	1944	1944	1072	1121	570	638	570	638
T2	0	0	1913	1829	1106	1079	595	622	595	622
T1	0	0	1550	1560	934	979	581	594	581	594

ABSORPTIONS MKB (INPUT 8-3 CARD)

LD ID	SH REQ,	SH ACT	BLR BNK	ECON	OUTPUT TOT KF
2	64.6	87.6	46.4	0	282.7
T2	54.6	72.4	42	0	229.9

The Babcock & Wilcox Company

Generating Powerful SolutionsSM

a McDermott company

Calibration Data Sheet

Plant:	City of Hamilton		Test#:	1	
Date:	11/30/2005		Load:	13.5 MW	
Grid Location	Analyzer:	Cal Gas:		Cal In	Cal Out
Boiler Gas out	O ₂ %	N ₂	Zero	0.00	0.00
	4900	7.48	Span	7.49	7.50
Boiler Gas out	CO PPM	N ₂	Zero	0.01	0.02
	4900	480.4	Span	480.90	481.50
Boiler Gas out	NO _x PPM	N ₂	Zero	-0.17	1.80
	42C	990.9	Span	991.00	973.30
Air Heater Out	O ₂ %	N ₂	Zero	0.00	-0.01
	1440	7.48	Span	7.48	7.49
Precip Out	O ₂ %	N ₂	Zero	-0.01	-0.01
	1440	7.48	Span	7.48	7.49
Cal'd by:			Time:	7:21	9:54

Calibration gas cylinder numbers

N ₂	ALM031225	O ₂ /CO	ALM066742	NO _x	ALM066915
----------------	-----------	--------------------	-----------	-----------------	-----------

The Babcock & Wilcox Company

Generating Powerful SolutionsSM

a McDermott company

Calibration Data Sheet

Plant:	City of Hamilton		Test#:	2	
Date:	11/30/2005		Load:	Full Load	
Grid Location	Analyzer:	Cal Gas:		Cal In	Cal Out
Boiler Gas out	O ₂ %	N ₂	Zero	0.00	0.00
	4900	7.48	Span	7.48	7.50
Boiler Gas out	CO PPM	N ₂	Zero	0.01	0.26
	4900	480.4	Span	480.40	482.00
Boiler Gas out	NO _x PPM	N ₂	Zero	1.41	2.22
	42C	990.9	Span	991.00	979.80
Air Heater Out	O ₂ %	N ₂	Zero	-0.01	0.00
	1440	7.48	Span	7.49	7.49
Precip Out	O ₂ %	N ₂	Zero	0.00	0.01
	1440	7.48	Span	7.49	7.51
Cal'd by:			Time:	12:44	17:04

Calibration gas cylinder numbers

N ₂	ALM031225	O ₂ /CO	ALM066742	NO _x	ALM066915
----------------	-----------	--------------------	-----------	-----------------	-----------

Customer: City of Hamilton
Plant & Unit: Unit 8

Test ID: T1
Run #: 1

Date: 11/30/2005

Analyzer Drift and Bias Correction						
	Blr Out O ₂	Blr Out CO	Blr Out NO _x	HESP Out O ₂	AH OUT O ₂	
Test Avg (raw)	8.65	21.00	242.93	9.87	10.30	
Test Start	11/30/05 07:40	11/30/05 07:40	11/30/05 07:40	11/30/05 07:40	11/30/05 07:40	
Test Stop	11/30/05 08:45	11/30/05 08:45	11/30/05 08:45	11/30/05 08:45	11/30/05 08:45	
Span Gas	7.48	480.40	990.90	7.48	7.48	
Zero Gas	0	0	0.00	0.00	0.00	
Cal In						
Cal Time	11/30/05 07:21	11/30/05 07:21	11/30/05 07:21	11/30/05 07:21	11/30/05 07:21	
Zero	0.00	0.01	-0.2	0.0	0.0	
Span	7.49	480.90	991.0	7.5	7.5	
Cal Out						
Chk Time	11/30/05 09:54	11/30/05 09:54	11/30/05 09:54	11/30/05 09:54	11/30/05 09:54	
Zero	0.02	0.02	1.80	-0.01	-0.01	
Span	7.50	481.50	973.30	7.49	7.49	
Span, Test Mid Pt.	7.493	481.102	985.042	7.483	7.483	
Zero, Test Mid Pt.	0.007	0.013	0.493	-0.010	-0.003	
M, Test Mid Pt.	0.999	0.999	1.006	0.998	0.999	
B, Test Mid Pt.	-0.007	-0.013	-0.496	0.010	0.003	
Corrected Avg	8.64	20.96	244.00	9.86	10.29	

Customer: City of Hamilton
Plant & Unit: Unit 8

Test ID: T2
Run #: 1

Date: 11/30/2005

Analyzer Drift and Bias Correction						
	Blr Out O ₂	Blr Out CO	Blr Out NO _x	HESP Out O ₂	AH OUT O ₂	
Test Avg (raw)	5.20	61.55	202.07	6.63	6.96	
Test Start	11/30/2005 14:40	11/30/05 14:40	11/30/05 14:40	11/30/05 14:40	11/30/05 14:40	
Test Stop	11/30/2005 16:49	11/30/05 16:49	11/30/05 16:49	11/30/05 16:49	11/30/05 16:49	
Span Gas	7.48	480.40	990.90	7.48	7.48	
Zero Gas	0	0	0.00	0.00	0.00	
Cal In						
Cal Time	11/30/05 12:44	11/30/05 12:44	11/30/05 12:44	11/30/05 12:44	11/30/05 12:44	
Zero	0.00	0.01	1.41	0.00	-0.01	
Span	7.48	480.40	991.00	7.49	7.49	
Cal Out						
Chk Time	11/30/2005 17:04	11/30/2005 17:0	11/30/2005 17:0	11/30/2005 17:0	11/30/2005 17:04	
Zero	0.00	0.26	2.22	0.01	0.00	
Span	7.50	482.00	979.80	7.51	7.49	
Span, Test Mid Pt.	7.500	480.400	991.000	7.490	7.490	
Zero, Test Mid Pt.	0.000	0.010	1.410	0.000	-0.010	
M, Test Mid Pt.	0.997	1.000	1.001	0.999	0.997	
B, Test Mid Pt.	0.000	-0.010	-1.412	0.000	0.010	
Corrected Avg	5.19	61.54	200.93	6.62	6.95

STANDARD LABORATORIES INC.
NORTHERN DIVISION
P.O. BOX 5214
CRENSHAW, PA 16638
(814) 886-7400

STANDARD LABORATORIES, INC.

DATE: 1-30-2008
SAMPLE NO. 882808

BABCOCK & WILCOX
20 SOUTH VAN BUREN AVENUE
PO BOX 351
BARBERSVILLE, OH. 44203-0351

SAMPLE ID: HAMILTON HIGH LOAD COAL SAMPLE

OPERATING CO.:
SAMPLED BY: CUSTOMER PROVIDED
MINE:
LOCATION:

DATE SAMPLED: 11/30/05
WEATHER:
GROSS WEIGHT:

DATE RECEIVED: 1/18/06

OTHER ID: FROM THE CITY OF HAMILTON UNIT 8 - CHARGE #1688ENG148MA - REQ:
DOUGLAS J. DEVAULT

CERTIFICATE OF ANALYSIS

	ASTM METHOD	AS RECEIVED	DMY BASIS
MOISTURE	D2961 D3382 D3173	7.46%	
VOLATILE	D3175M	28.76%	31.08%
FIXED CARBON	D3172	47.86%	51.71%
ASH	D3174	15.82%	17.21%
SULFUR	D4239 METHOD B	0.79%	0.86%
CARBON	D3178 D5373	64.87%	70.18%
HYDROGEN	D3178 D5373	4.20%	4.54%
NITROGEN	D3179 D5373	1.08%	1.17%
OXYGEN	D3176	9.68%	6.12%
BTU/LB	D5865	11262	12169
MAF BTU/LB			14699
LBS OF SO2 PER MILLION BTU			1.41
LBS OF SULFUR PER MILLION BTU		0.781	
FREE SWELLING INDEX	D270	2.5	
GRINDABILITY INDEX	D409	49	

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NORTHEAST DIVISION
P.O. BOX 214
GREYSBURG, PA 16630
(814) 886-7480

STANDARD LABORATORIES, INC.

DATE: 1-30-2006
SAMPLE NO. 082805

BABCOCK & WILCOX
20 SOUTH VAN BUREN AVENUE
PO BOX 351
BARRERTON, OH. 44203-WJ51

SAMPLE ID: HAMILTON HIGH LOAD COAL SAMPLE

OPERATING CO.:
SAMPLED BY: CUSTOMER PROVIDED
MINE:
LOCATION:

DATE SAMPLED: 11/30/05
WEATHER:
GROSS WEIGHT:

DATE RECEIVED: 1/18/06

OTHER ID: FROM THE CITY OF HAMILTON UNIT 8 - CHARGE #160KENG140NA - HQ2:
DOUGLAS J. DEVAULT

CERTIFICATE OF ANALYSIS

ASH FUSION TEMPERATURE(S)	
D1857	REDUCING ATMOSPHERE
INITIAL DEFORMATION	OVER 2800
SOFTENING TEMPERATURE	OVER 2800
HEMISPHERICAL TEMPERATURE	OVER 2800
FLUID TEMPERATURE	OVER 2800

ASH MINERAL
D2795 D3682

SILICON DIOXIDE	58.25 %
ALUMINUM OXIDE	28.34 %
FERRIC OXIDE	5.80 %
TITANIUM DIOXIDE	1.65 %
PHOSPHOROUS PENTOXIDE	0.83 %
CALCIUM OXIDE	1.15 %
MAGNESIUM OXIDE	0.92 %
SODIUM OXIDE	0.43 %
POTASSIUM OXIDE	3.89 %
SULFUR TRIOXIDE	0.48 %

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DATE: 2-1-2006
SAMPLE NO. 882889BABCOCK & WILCOX
20 SOUTH VAN BUREN AVENUE
PO BOX 351
BARBERTON, OH. 44203-0351

SAMPLE ID: HAMILTON - ASH SAMPLE

OPERATING CO.:
SAMPLED BY: CUSTOMER PROVIDED
MINE:
LOCATION:DATE SAMPLED: 11/30/85
WEATHER:
GROSS WEIGHT:

DATE RECEIVED: 1/18/86

OTHER ID: FROM THE CITY OF HAMILTON UNIT 8 - CHARGE #16888881408A - RES:
DOUGLAS J. DEVAULT

CERTIFICATE OF ANALYSIS

	ASTE METHOD	AS DETERMINED
CARBON	03178 05373	12.32%
LOSS ON IGNITION		13.43%
CARBON DIOXIDE	01736	0.21%

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APPENDIX 3: BOILER ISLAND PERFORMANCE TEST REPORT

**CITY OF HAMILTON
UNIT #8
TEST REPORT**

For
ADDITIONAL BASELINE TEST

Prepared By: Douglas J. DeVault
Principal Engineer
Babcock & Wilcox Co.

EXECUTIVE SUMMARY

Tests were conducted on this unit during the week of June 29, 2006. The purpose was to obtain additional information to the original baseline tests for an upcoming demonstration of firing a high sulfur bituminous coal with oxygen as the oxidant and recycled flue gas. The additional information to be obtained was;

- Determine source high air infiltration between the airheater and boiler outlet
- Traverse the boiler outlet for particulate for more accurate unburned carbon values
- Determine hot electrostatic precipitator (HESP) capture efficiency

One test was conducted during this test program. The test was conducted at full load condition. The results of the test were used to supplement the original baseline test conducted on November 30, 2005. Table A3.1 contains a description of the test conducted and a summary of the pertinent performance information.

TABLE A3.1 – PERFORMANCE SUMMARY

Description	Test 1
Date	6/30/06
Time	15:00- 17:00
Gross Megawatts	N/A
HP Steam Flow, klb/hr	219.9
Sec. SH Spray Flow, klb/hr	24.7
Main Steam Temperature	927
Furnace Exit Gas Temperature (MHVT) F	1903
Gas Temperature Lvg Blr, F	554
Gas Temperature Lvg AH, F	347
Excess Air Lvg Blr, %	27.8
Excess O2 Lvg Blr, %	4.66
Excess O2 Lvg HESP, %	6.59
Average O2 Lvg AH, %	6.87
Average CO Lvg Blr, ppm	147.58
Average NOx Lvg Blr, ppm	227.55
Average NOx Lvg Blr, lb/MBtu	0.34

DISCUSSION OF RESULTS

All tests were conducted, and all operational performance of the unit was determined, in accordance with the guidelines set forth in the following ASME and EPA publications:

ASME Power Test Code PTC.4.1-1967, Steam Generators

ASME Power Test Code PTC 4.3-1968, Air Heaters

*ASME Instruments and Apparatus, PTC 19.10-1985, Part 10 Flue & Exhaust Gas Analysis
EPA Method 7E, Determination of Nitrogen Oxide Emissions (Instrumental Analyzer Procedure)*

Air and gas weights were calculated stoichiometrically, using measured O₂ and fuel analysis in accordance with PTC 19.10. Table A3.2 contains a summary of the actual fuel analyses obtained for these tests.

NO_x in pounds per million Btu was calculated using the EPA F factor derived from the following equations:

$$F_x = \frac{10^6 \times (1.53 \times C + 3.64 \times H + 0.57 \times S + 0.14 \times N_2 - 0.46 \times O_2)}{HHV}$$

$$NO_x lb / MBtu = NO_{xppm} \times 1.194 \times 10^{-7} \times F_x \times \left(\frac{20.95}{20.95 - O_2} \right)$$

Where:

C = % Carbon in Fuel

H = % Hydrogen in Fuel

S = % Sulfur in Fuel

N₂ = % Nitrogen in Fuel

O₂ = % Oxygen in Fuel For first equation

HHV = Higher Heating Value of Fuel

O₂ = % O₂ in flue gas at the NO_x measurement location For the second equation

TABLE A3.2 – Fuel Analysis		
Constituent	As Received	Dry
HHV Btu/lb	12481	13351
Carbon, %	70.62	75.55
Hydrogen, %	4.43	4.74
Nitrogen, %	1.31	1.40
Sulfur, %	0.67	0.72
Oxygen, %	6.42	6.85
Moisture, %	6.51	-
Ash, %	10.04	10.74
Fixed Carbon, %	51.23	54.79
Volatile matter, %	32.22	34.47

Temporary test instrumentation consisted of gas analysis measurements at the boiler gas outlet flue, the hot precipitator outlet flue, and the airheater outlet flue. These measurements were obtained from side to side gas grids consisting of multi-point probes. A High Velocity Thermocouple (HVT) probe was used for measuring Furnace Exit Gas Temperature (FEGT). All other data was obtained using permanent plant instrumentation. Flue gas analysis data was obtained at ten second intervals and averaged and stored to disc every five minutes. Temperature

data was obtained at one minute intervals and averaged and stored to disc every five minutes. All plant data was obtained from the plant control system at one minute intervals.

Unburned carbon samples were obtained by traversing the boiler outlet. Analysis showed the total unburned carbon to be 21.71%. Carbon in Ash values were corrected for carbonates by multiplying the percent CO₂ in the ash by the ratio of the molecular weight of Carbon (12) to the molecular weight of Carbon Dioxide (44) and subtracting this value from the Carbon in the ash. The carbon in ash corrected for carbonates was 21.54%.

FEGT was determined by using a 12' HVT probe to traverse the furnace exit plane. One observation door on each side of the unit was used for traversing. Although this was not as many points for traversing as desired, it was considered adequate for this testing. Ash particles were gathered by isokinetic sampling at the inlet and outlet of the HESP for determination of the collection efficiency. The collection efficiency was in excess of 98%. All of the raw data is stored on disc and available upon request. Summary printouts of the combustion and efficiency calculations, and boiler performance as well as the coal and ash analysis reports follow this report.

CONCLUSIONS

For the most part testing went smoothly and no significant items of concern were encountered during the testing. Although temperatures were gathered around the airheater and a heat balance was performed, the actual air going to the boiler could not be determined. This is due to the lack of available data to determine the amount of air being bypassed around the airheater for tempering. Because of this the goal of determining the source high air infiltration between the airheater and boiler outlet was not realized.

The Unburned Carbon analyses yielded an average baseline Carbon in Ash value of 21.71 %. In a utility boiler this is generally considered very high, however, it has been B&W's experience that small furnaces such as this tend to run colder than the a large ones and this unburned carbon is in line with past experience.

I would like to take this opportunity to thank the people at The City of Hamilton for their cooperation in conducting these tests. It was a pleasure working with all of them.

LOAD IDENT		002	T1
RUN NO.		1	2
	UNITS		
LOAD		Base	MW
STEAM AND WATER FLOWS			
H.P. STEAM FLOW	MLB/HR	250.00	219.90
SH SPRAY		26.39	24.70
AIR AND GAS FLOWS			
XS AIR LVG ECO	PERCENT	27.0	44.6
XS AIR TO BURNERS		25.0	25.8
GAS LVG FURNACE	MLB/HR	338.1	305.1
GAS ENT AIR HEATERS		340.6	345.0
AIR LVG AIR HEATERS		312.5	281.6
GAS OR AIR TEMPERING		5.0	0.0
FLUID TEMPERATURES		DEG F	
SEC SH OUTLET		900.	927.
PRI-1 SH OUT		879.	880.
ATTEMP-1 OUT		687.	672.
ECO WATER IN		350.	338.
SH SPRAY WATER		350.	338.
RH-1 SPRAY WATER		350.	338.
DRUM SAT. TEMP		539.	534.
PRESSURES		PSIG	
DRUM		940.	900.
SEC SH OUT		875.	855.
AIR AND GAS TEMPS		DEG F	
FURNACE EXIT (MEAS)	MHVT	1944	1903
GAS LVG ECO	DEG F	570.	554.
GAS LVG AIR HEATERS		325.	347.
AVG AIR ENT AIR HEATERS		80.	85.
AVG AIR LVG AIR HEATERS		469.	353.

DATA SUMMARY	B+W CONTR. NO. JEB	P08475 17.04	TIME
1025 DATE 11/30/06		PAGE 2	

LOAD IDENT		002	T1
RUN NO.		1	2
	UNITS		
LOAD		Base	MW
ABSORPTIONS	MKB		
TOTAL SH		87.6	81.9
PRI-1 SH		54.8	48.0
SEC SH		32.8	34.0
TOTAL OUTPUT		282.4	254.7
BURN ZONE HT REL RATE	MB/FT2-H	198.7	180.7
SUMMATION OF LOSSES	PERCENT		
DRY GAS		5.87	6.82
WATER AND WV FROM FUEL		4.31	4.30
MOISTURE IN AIR		0.14	0.16
UNB COMBUSTIBLE		0.30	3.20
RADIATION		1.04	1.15
UNMEAS AND MFRS MARGIN		1.50	0.50
OTHER LOSSES		0.00	0.00
TOTAL LOSSES	PERCENT	13.16	16.14
TOTAL CREDITS	PERCENT	0.00	0.00
EFFICIENCY	PERCENT	86.84	83.86

PERFORMANCE ANALYSIS SUMMARY

DATE 11/30/06

LD	LOAD	HP STM	SH	XAIR		TIME	TIME	BURNERS
ID	MW	FLOW	SPRAY	ECO %	DATE	STRT	END	OUT
2	Base	250	26.4	27	E. KENT	UCKY		
T1	MW	219.9	24.7	44.5	6/29/2006	1440	1649	

A/E VALUES

LD	ECON	PRI-1	PRI-2	SEC	SEC-1	SEC-2	BLR	TOT			
ID							BANK	SH	ECO-1	ECO-2	ECO-3
2	0	1.172	0	0.927	0	0	1.045	1.067	0	0	0
T1	0	1.155	0	1.107	0	0	0.935	1.133	0	0	0

KF VALUES

LD	ECO-1	ECO-2	ECO-3	PRI-1	PRI-2	SEC	SEC-1	SEC-2	BBNK-1	BBNK-2	CMB SH
ID											
2	0	0	0	1.44	0	0.86	0	0	1.92	0	0
T1	0	0	0	1.36	0	1.18	0	0	2.02	0	0

GAS TEMPERATURES F MHVT

LD	ENT	PLATENS	FURN	EXIT	ENT	GEN BK	LVG	GEN BK	LVG L	UNIT
ID	ACT	EXP	ACT	EXP	ACT	EXP	ACT	EXP	ACT	EXP-KF
2	0	0	1944	1944	1072	1121	570	638	570	638
T1	0	0	1903	1883	995	1090	554	624	554	624

ABSORPTIONS MKB (INPUT 8-3 CARD)

LD					OUTPUT
ID	SH REQ,	SH ACT	BLR BNK	ECON	TOT KF
2	64.6	87.6	46.4	0	282.7
T1	60.1	81.9	36.6	0	254.3

08/04/2006 13:24 FAX 8148862842

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STANDARD LABORATORIES INC.
NORTHERN DIVISION
P.O. BOX 214
CRESSON, PA 16630
(814) 886-7400

STANDARD LABORATORIES, INC.

DATE: 7-25-2006

SAMPLE NO. 901099

BABCOCK & WILCOX
20 SOUTH VAN BUREN AVENUE
PO BOX 351
BARBERTON, OH. 44203-0351

SAMPLE ID: RAW COAL FROM PILE 1 (SMALL
NEAREST #9) -- 17:30

OPERATING CO.:
SAMPLED BY: CUSTOMER PROVIDED
MINE:
LOCATION:

DATE SAMPLED: 6/29/06

DATE RECEIVED: 7/24/06

WEATHER:

GROSS WEIGHT:

OTHER ID: CITY OF HAMILTON UNIT 8 TESTS -- ORDER #168EENG140EN -- REQ:
EDWARD COLLET

CERTIFICATE OF ANALYSIS

	ASTM METHOD	AS RECEIVED	DRY BASIS
MOISTURE	D2961 D3302 D3173	9.52%	
VOLATILE	D3175M	33.25%	36.75%
FIXED CARBON	D3172	49.64%	54.86%
ASH	D3174	7.59%	8.39%
SULFUR	D4239 METHOD B	0.95%	1.05%
CARBON	D3178 D5373	69.28%	76.57%
HYDROGEN	D3178 D5373	4.46%	4.93%
NITROGEN	D3179 D5373	1.41%	1.56%
OXYGEN	D3176	6.79%	7.50%
BTU/LB	D5865	12206	13490
MAF BTU/LB			14726
LBS OF SO2 PER MILLION BTU			1.56
LBS OF SULFUR PER MILLION BTU		0.778	

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CRESSON, PA 16630
(814) 886-7400

STANDARD LABORATORIES, INC.

DATE: 7-25-2006

SAMPLE NO. 901098

BABCOCK & WILCOX
20 SOUTH VAN BUREN AVENUE
PO BOX 351
BARBERTON, OH. 44203-0351

SAMPLE ID: RAW COAL FROM PILE 1 (LARGE) --
17:30

OPERATING CO.:
SAMPLED BY: CUSTOMER PROVIDED
MINE:
LOCATION:

DATE SAMPLED: 6/29/06
WEATHER:
GROSS WEIGHT:

DATE RECEIVED: 7/24/06

OTHER ID: CITY OF HAMILTON UNIT 8 TESTS -- ORDER #168EENG140EN -- REQ:
EDWARD COLLET

CERTIFICATE OF ANALYSIS

	ASTM METHOD	AS RECEIVED	DRY BASIS
MOISTURE	D2961 D3302 D3173	6.51%	
VOLATILE	D3175M	32.22%	34.47%
FIXED CARBON	D3172	51.23%	54.79%
ASH	D3174	10.04%	10.74%
SULFUR	D4239 METHOD B	0.67%	0.72%
CARBON	D3178 D5373	70.62%	75.55%
HYDROGEN	D3178 D5373	4.43%	4.74%
NITROGEN	D3179 D5373	1.31%	1.40%
OXYGEN	D3176	6.42%	6.85%
BTU/LB	D5865	12481	13351
MAF BTU/LB			14957
LBS OF SO2 PER MILLION BTU			1.08
LBS OF SULFUR PER MILLION BTU		0.537	

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DATE: 7-26-2006

SAMPLE NO. 901100

BABCOCK & WILCOX
20 SOUTH VAN BUREN AVENUE
PO BOX 351
BARBERTON, OH. 44203-0351

SAMPLE ID: PC COAL - PULVERIZED COAL SAMPLE

OPERATING CO.:
SAMPLED BY: CUSTOMER PROVIDED
MINE:
LOCATION:

DATE SAMPLED: 6/29/06
WEATHER:
GROSS WEIGHT:

DATE RECEIVED: 7/24/06

OTHER ID: CITY OF HAMILTON UNIT 8 TESTS -- ORDER #168EENG140EN -- REQ:
EDWARD COLLET

CERTIFICATE OF ANALYSIS

	ASTM METHOD	AS RECEIVED	DRY BASIS
MOISTURE	D2961 D3302 D3173	2.36%	
VOLATILE	D3175M	33.52%	34.33%
FIXED CARBON	D3172	53.13%	54.42%
ASH	D3174	10.99%	11.25%
SULFUR	D4239 METHOD B	0.80%	0.82%
CARBON	D3178 D5373	72.48%	74.23%
HYDROGEN	D3178 D5373	4.20%	4.30%
NITROGEN	D3179 D5373	1.37%	1.40%
OXYGEN	D3176	7.80%	8.00%
BTU/LB	D5865	12795	13104
MAF BTU/LB			14765
LBS OF SO2 PER MILLION BTU			1.25
LBS OF SULFUR PER MILLION BTU		0.625	

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STANDARD LABORATORIES, INC.

DATE: 8- 4-2006
SAMPLE NO. 901101

BABCOCK & WILCOX
20 SOUTH VAN BUREN AVENUE
PO BOX 351
BARBERTON, OH. 44203-0351

SAMPLE ID: FULL LOAD ALL MILLS METHOD 5 RUN 1
BOILER OUT

OPERATING CO.:
SAMPLED BY: CUSTOMER PROVIDED
MINE:
LOCATION:

DATE SAMPLED: 6/28/06
WEATHER:
GROSS WEIGHT:

DATE RECEIVED: 7/24/06

OTHER ID: CITY OF HAMILTON UNIT 8 TESTS -- ORDER #168EENG140EN -- REQ:
EDWARD COLLET

CERTIFICATE OF ANALYSIS

	ASTM METHOD	AS DETERMINED
CARBON	D3178 D5373	22.14%
LOSS ON IGNITION		26.37%
CARBON DIOXIDE	D1756	0.69%

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N/A

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STANDARD LABORATORIES, INC.

DATE: 8- 4-2006
SAMPLE NO. 901103

BABCOCK & WILCOX
20 SOUTH VAN BUREN AVENUE
PO BOX 351
BARBERTON, OH. 44203-0351

SAMPLE ID: FULL LOAD ALL MILLS METHOD 5 RUN 3
BOILER OUT

OPERATING CO.:
SAMPLED BY: CUSTOMER PROVIDED
MINE:
LOCATION:

DATE SAMPLED: 6/28/06
WEATHER:
GROSS WEIGHT:

DATE RECEIVED: 7/24/06

OTHER ID: CITY OF HAMILTON UNIT 8 TESTS -- ORDER #168EENG140EN -- REQ:
EDWARD COLLET

CERTIFICATE OF ANALYSIS

	ASTM METHOD	AS DETERMINED
CARBON	D3178 D5373	21.28%
LOSS ON IGNITION		26.17%
CARBON DIOXIDE	D1756	0.54%

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N/A

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APPENDIX 4: BACKGROUND AND PRIOR PILOT TESTING

Oxy-combustion for enhanced oil recovery was evaluated by B&W initially in 1979 per request of a major oil company. Since the late 1990's, B&W has been a member of the Canmet oxy-combustion consortium and participated in 1 MBtu/h (million Btu per hour) tests in Canada. Pilot-scale development of the technology for commercial applications began several years ago at B&W in collaboration with Air Liquide. Highlights of the pilot-scale oxy-combustion experience are described below.

Pilot-scale Oxy-Combustion Facility

Figure A.4.1 shows B&W's 5 MBtu/h Small Boiler Simulator (SBS) facility used in oxy-combustion development work. The water-cooled SBS furnace measures 4.5 ft wide by 7 ft deep by 14 ft high and it has insulated walls. Pulverized coal (PC) flow from a storage bin is measured and controlled by a calibrated weigh feeder. The coal is then transported by heated primary oxidant (PO) at about 150°F to the burner. Secondary oxidant (SO) flow is preheated indirectly to 600°F by a gas-fired heater. For staged combustion, the 600°F tertiary oxidant (TO) is directed to over-fire oxidant ports located above the main combustion zone.

Figure A.4.1 - B&W Small Boiler Simulator (SBS)

Combustion products leave the furnace and enter a convection pass section that consists of four different sections simulating secondary superheater, reheater, primary superheater, and economizer as shown in Figure A.4.1. After leaving the convection pass section, the flue gas temperature is controlled further by a heat exchanger to values suitable for dry scrubber and baghouse operation. Collected particulate matter in the baghouse modules is dislodged from the bags by a pulse jet cleaning cycle for subsequent removal and disposal. A wet flue gas desulfurization system can also be connected and operated when the dry scrubber is not in service. Furnace and convective pass time-temperature profiles, SO₂ and NO_x concentrations,

fly ash particle size, and unburned carbon levels of this facility accurately resemble full-scale coal-firing behavior.

Figure A.4.2 shows the overall facility layout especially configured for combustion of pulverized coal with oxygen and recycled flue gas. A condensing heat exchanger (CHX) is used to cool down and dehumidify the recycled flue gas before it is mixed with oxygen and recycled to the primary and secondary oxidant streams. The oxygen levels in the primary and secondary streams, convection pass exit gas, and CHX exit flue gas, are measured with oxygen sensors. For safety purposes, boiler permissives and interlocks are set to limit the oxygen concentration in the flue gas at 25% for secondary and 21% for the primary zones using the oxygen sensors and the oxygen flow controls.

Figure A.4.2 - Overall Schematic of the SBS in Oxy-combustion Mode

A scaled-down version of the B&W[®] DRB-4Z[®] low-NO_x burner as shown in Figure A.4.3 was modified for oxygen firing and used in these tests. The burner was designed for staged combustion operation. This burner has been installed in numerous commercial units with excellent results. Liquid oxygen from an onsite 9,000-gallon oxygen tank was vaporized and regulated to an appropriate pressure for delivery to the test area inside a copper line. The oxygen delivery control system was integrated into the boiler safety interlock system. Flue gas was sampled continuously from the convection pass exit and the concentrations of O₂, NO_x, CO, CO₂, and SO₂ were measured on a dry-basis by calibrated gas analyzers. The stack fly ash was

isokinetically sampled and analyzed for carbon content to determine carbon utilization in the system. Tests were performed with a low-sulfur sub-bituminous coal. Proximate and Ultimate analyses on an as-received basis are listed in Table A.4.1.

Figure A.4.3 - B&W's DRB-4Z[®] Low-NO_x Burner

Table A.4.1 - Pilot-scale Coal Analysis	
Parameter	Value
Moisture, %	19.95
Ash, %	5.10
Volatile Matter, %	35.04
Carbon, %	55.87
Hydrogen, %	6.05
Nitrogen, %	0.78
Sulfur, %	0.28
Heating Value, Btu/lb	9452
Mercury, ppm	0.07
Chlorine, ppm	118

Relevant Pilot-scale Results

Combustion and emissions characteristics of oxy-combustion with flue gas recycle were measured and compared to air-fired operation. Optimization of the boiler parameters for NO_x reduction was carried out without altering the heat transfer in the furnace and convection pass. Effects of flue gas recycle flow rate, burner zone stoichiometry, and oxygen distribution into different zones of the burner and furnace were investigated.

Oxidant Switching and Impact on Flame Shape - Switching from air to O₂-enriched flue gas (oxy-combustion) operation was demonstrated during pilot-scale testing, and operating procedures for smooth transition from air to oxy-combustion then back to air-fired modes were developed. In these tests, the overall mass flow rates of combustion gases through the boiler for the oxy-combustion and air-fired conditions were similar. A stable flame, attached at the throat, was obtained under the oxy-combustion conditions. The oxy-combustion flame was more flared than the air-fired flame when both flames were under sub-stoichiometric conditions presumably due to lower burner velocities under the oxy-combustion case. Increasing the mass flow rate of the oxygen-enriched flue gas to the burner made the flame more similar to that of air-fired combustion. Visually, the oxy-combustion flame was cooler and less bright than the air-fired flame under similar sub-stoichiometric conditions, presumably because of the higher specific heat of CO₂. The oxy-firing flame could be made similar to or somewhat hotter than the air-fired flame by adjusting the oxygen and flue gas flows to the burners. No special materials considerations are necessary in constructing oxy-combustion capable, low-NO_x burners for existing boilers retrofits.

NO_x emissions characteristics - Figure A.4.4 compares the air-fired and oxy-combustion NO_x results. NO_x emissions were approximately 65% lower in oxy-combustion conditions than the baseline results. Baseline NO_x emission ranged from 0.22 to 0.40 lb/MBtu when burner stoichiometry varied from 0.57 to 1.21. At the design stoichiometry of 0.9, the NO_x level was 0.28 lb/MBtu. By contrast, during the oxy-combustion tests NO_x levels ranged from 0.065 to 0.106 lb/MBtu as the burner stoichiometry varied from about 0.8 to 1.05. Under oxy-combustion conditions, NO_x emissions are impacted by the burner stoichiometry, oxygen flow rate into the primary zone and the overall flue gas recycle rate. Flame temperature can increase with increasing O₂ concentration in the sub-stoichiometric primary combustion zone, or by reducing the flue gas flow rate. High flame temperatures, under reducing conditions, promote the reactions between recirculated NO_x and coal pyrolysis nitrogen-bearing products to form molecular nitrogen. Since the flue gas recycle flow rates into the primary and secondary zones are set by pulverizer and boiler operation requirements, NO_x can be reduced by optimizing the primary zone oxygen level and burner stoichiometry.

Figure A.4.4 - NO_x Emissions for fired and oxy-fired combustion of PRB coal at 3.4-3.7% overall excess O₂

Figure A.4.5 shows the effect of recycled gas flow rate on NO_x emissions at a nominal burner stoichiometry of about 0.85. NO_x emissions decreased to a minimum of 0.065 lb/MBtu as the recycled flue gas flow rate decreased. As explained earlier, this is a consequence of enhanced NO_x reduction reactions at higher flame temperatures in the reducing zone of the burner. Although in retrofit applications, the flue gas recycle rate has to be sufficient to maintain proper mass flux through the furnace and convection pass of the boiler, further reduction of recycled flue gas may be possible for new boilers specifically designed for lower mass flux of flue gases.

Figure A.4.5 - Impact of FGR Flow Rate on NO_x Emissions

Unburned Combustibles - Oxy-combustion generally reduces the unburned carbon and CO levels [2]. Due to the high reactivity of PRB coal, the corresponding baseline unburned carbon in the fly ash ranged only from 0.62% to 0.80% in comparison with 0.52% to 0.86% for oxy-combustion tests. At optimum conditions, CO levels ranged from 12 to 15 ppm for the baseline tests versus 70 to 91 ppm under the oxy-combustion operation.

CO₂ Content in Flue Gas - One area of concern during oxy-combustion tests was the amount of air infiltration into the unit. During the tests, the maximum CO₂ concentration that could be attained was 80% (corrected to 3% boiler exit oxygen concentration). The infiltrated air was approximately 5% of the total boiler gas flow rate (i.e., air, oxygen, flue gas, and coal). The SBS boiler is balanced draft and operates under slightly negative pressure. The main sources of air leaks are suspected to be the induced draft fan, baghouse and scrubber that were operating under negative pressures. This level of air leakage is a good representation of potential air leaks in commercial boiler retrofits from the electrostatic precipitator, air heater, etc.

Temperature Profiles and Heat Absorption - Oxy-combustion was originally thought to change the heat transfer characteristics to the point that addition of extra heat transfer surfaces would be necessary to generate steam. Two tests were performed at the optimum air-fired and oxy-combustion conditions to explore the effect of burning mode on the boiler temperature profiles, and overall heat absorption in the furnace. Flame temperature, furnace exit gas temperature, and convection pass exit temperature were measured at fixed boiler load, overall mass flow rate through the boiler, and excess oxygen level. The total heat absorption in the convective pass was estimated from the intake water flow rates to the drum in the period of 8 hours. The results are illustrated in Table A.4.2. The boiler temperature profiles were similar for both cases. Oxy-

combustion flame temperature, measured with 2 color pyrometer, was 2,480°F; about 100 °F more than air firing combustion conditions (2,370°F) presumably due to differences in the burner stoichiometry and flue gas molecular weights. The convection pass exit temperature was very similar indicating that slightly more heat was transferred from gas to water in the convection pass. This agreed well with the convection pass heat absorption, measured by intake water to steam drum. The higher convection pass heat absorption under the oxy-combustion is due to smaller flue gas volume (higher molecular weight) and longer residence time for heat transfer within the tube banks. The smaller flue gas volume also reduces the flue gas velocity and decreases the heat transfer coefficient. But in this facility, the effect of higher residence time is more than the lower heat transfer coefficient. In order to further explain these results, additional flame temperature and emissivity measurements were made.

Table A.4.2 Temperature and Heat Transfer Measurements		
	Air Firing	Oxy-combustion
Boiler Mass Flow Rate	4,167 lb/h	4,030 lb/h
Burner Stoichiometry	0.86	1.05
FEGT	1797 °F	1880 °F
Convection Pass Temperature	555 °F	547 °F
Convection Pass Heat Absorption	43.6%	46.8%

In a different test, the oxy-combustion temperature profiles under sub-stoichiometric conditions showed that the average furnace exit gas temperature was slightly lower than the air firing case. Although, the measured convection pass exit gas temperatures for oxy-combustion conditions were generally higher than air-fired operation values, additional heat transfer surface may not be required.

Flame Temperature and Emissivity - Total radiation and spectral radiance from the flame were measured from an area approximately 3 feet from the burner throat. Total radiation was measured with a thermopile detector in the range of 0.2 to 50 micrometers. The spectral radiance was measured using a computer controlled scanning monochrometer. The monochrometer is equipped with multiple photodetectors and gratings that can span a measurement range of about 0.2 to 15 micrometers.

The magnitude of the spectral emittance for the air-fired and oxy-combustion are nearly identical as illustrated in Figure A.4.6. Differences between air-fired and oxy-combustion are most pronounced at wavelengths of about 2.6 μm , 4.3 μm , and for wavelengths from 12 to 14 μm . These wavelengths correspond to wavebands specific to CO_2 . This is also the likely reason that the measured emittance exceeds a value of 1 at wavelengths of about 4.6 and 13 microns. The similarities in spectral emittance for the two firing conditions indicate that thermal radiation is primarily dependent on the coal, fly ash, and soot particles in the flame, and not on the concentration of CO_2 . The similarity of flame emissivities for oxy-combustion and air firing conditions and similarity of boiler temperature profiles are remarkable.

Figure A.4.6 Flame Emissivity Comparisons

Wet Scrubber Performance with High Sulfur Coal - In addition to combustion characterization tests with a PRB coal, a baseline test under air firing and an oxygen firing test were performed to determine the effect on oxygen firing on the effectiveness of the wet scrubber using a high-sulfur Illinois coal. These tests were performed at a nominal load of 5 MBtu/h, excess oxygen of 3.7%. In both tests, air staging with a nominal burner stoichiometry of 0.9 was used. In the oxygen firing mode, the secondary air was replaced with flue gas/oxygen mixture, which contained approximately 20% oxygen and 30% OFA oxygen concentration.

The wet scrubber operating conditions are presented in Table A.4.3 for the air-fired and oxygen/recycle combustion tests. Listed in the table are the inlet gas flow rate and temperature, absorber slurry flow rate and liquid to gas ratio (L/G), the inlet SO₂ and O₂ concentration and SO₂ removal efficiency and the slurry pH in the absorber mix tank. The liquid flux and slurry pH values for the air combustion tests are representative of commercial wet scrubber operation. The L/G ratio is somewhat greater than typical because the gas flow rate was slightly lower than normal. For the oxy-combustion tests the liquid flux was maintained the same as for air combustion so that the SO₂ removal capacity (lb/h) of the scrubber could be maintained. Even though the gas flow rate was reduced by about a factor of four for oxygen/recycle combustion, the sulfur mass loading to the wet scrubber was approximately the same since the SO₂ concentration increased by about a factor of 4. For all tests the SO₂ removal efficiency was quite high, in excess of 97%.

Table A.4.3 Wet Scrubber Test Conditions for the Air And Oxy-Combustion Tests				
Test Parameter	Air-Fired Combustion		Oxygen/Recycle Combustion	
	1A	1B	2A	2B
Inlet Flow Rate (lb/h)	4556	4805	1248	1273
Inlet Temperature (°F)	172	215	104	108
Inlet SO ₂ (ppm)	1988	2237	8084	8162
Inlet O ₂ (%)	6.44	6.32	5.68	5.56
Total Wet Scrubber Liquid Flux (gpm/ft ²)	59	59	59	59
Wet Scrubber L/G Ratio (gal/1000ft ³)	147	129	669	682
Wet Scrubber Slurry pH	5.31	5.38	5.30	5.43
SO ₂ Removal Efficiency (%)	99.3	97.6	99.6	99.6

In summary, the oxy-combustion technology has been successfully demonstrated and characterized in a 5 million Btu/h pilot facility. The overall oxy-combustion characteristics were comparable to the air-firing case in spite of the enormous change in oxidant composition from air to oxygen-enriched flue gas. The NO_x emissions from oxy-combustion were significantly lower (65%) than the air-fired case. The steam generation and heat transfer in the furnace and the convection pass changed only modestly. In commercial applications, site-specific feasibility studies will determine the boiler performance, but based on this pilot-scale study, no additional heat transfer surfaces are required. Air infiltration into the boiler under O₂-combustion conditions was reduced to the equivalent of 5% of the overall stoichiometry. Substitution of combustion air with oxygen and recycled flue gas increased the CO₂ concentration from 15% to 80% at the boiler exit. Boiler upgrades can further reduce the air infiltration and increase the flue gas CO₂ concentration. With oxy-combustion, the flue gas volume exiting the boiler is reduced by 70% relative to air-fired operation, thus facilitating additional processing or treatment that may be necessary to prepare the CO₂-rich stream for sequestration.

Appendix 5

Technical Specification and Cost Estimates

ALPC Reference No.: 30257

Revision: 0

American Air Liquide

Oxycombustion Technology for CO₂ Capture in Pulverized
Coal-Fired Boilers – DOE Project Task #6

683 MT/D Air Separation Unit

City of Hamilton, Ohio

April 12th. 2006

Air Liquide Process & Construction, Inc.
An Air Liquide Group Company
2700 Post Oak Blvd., Suite 1800
Houston, Texas, USA 77056-3026

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ATTACHMENT A: Preliminary Bloc Flow Diagram

ATTACHMENT B: Heat and Material Balance

ATTACHMENT C: Preliminary Plot Plan

ATTACHMENT D: Preliminary Load List

INTRODUCTION

This document describes the technical specification and cost estimates for the turnkey supply of one (1) air separation unit (ASU), referred to as the "Plant", for an oxy-combustion pilot project to be located in the City of Hamilton (COH), Ohio.

SECTION 1: PLANT PERFORMANCE AND DESIGN BASIS

1.1 Site Conditions

Table 1: Ambient Site Conditions

	Units	Nominal	Summer
Dry Bulb Temperature	°F	59	88
Relative Humidity	%	60	49.1
Wet Bulb Temperature	°F	51.5	72.7
Cooling Water Temperature	°F	64.1	81.7
Barometric Pressure	psia	14.42	14.42

1.2 Production

The Plant, under steady conditions, with the utilities and the air quality described in Section 1.3 through 1.8, operated at Summer ambient site conditions as stated in Section 1.1, will produce the following products:

Table 2: Air Separation Unit Production

Product	Flow (ST/D)	Pressure (psia)	Purity (minimum)
LP Gaseous Oxygen	739.3	22.47	95%
MP Gaseous Oxygen	12.9	44.37	95%

* Product flowrate is contained

1.3 Utilities

The Plant, under steady conditions, with the utilities and the air quality described in Section 1.3 through 1.8, operated at Nominal ambient site conditions as stated in Section 1.1, will consume the following utilities:

Table 3: Utility Requirements

Utility	Units	Nominal
LP Steam Flow (average)	lbs/h	1,263
LP Steam Flow (peak)	lbs/h	2,900
Make-up Water Flow	gpm	100
Blowdown water	gpm	27
Total Expected Power Consumption*	kW	7,340

* Power consumption is given at motor terminals including main air compressors, water wash and chilled water tower pumps, cryogenic liquid pump, operating cooling water system pump(s) and fan(s).

1.4 Air Quality

The ambient air quality at the site, which is being taken into consideration by Air Liquide for the design of the plant, is specified below.

Table 4: Design Air Quality

Impurities	Chemical Formula	Basis of Design (volume - vapor phase)
Nitrogen	N ₂	78.11%
Oxygen	O ₂	20.96%
Argon	Ar	0.93%
Carbon Monoxide	CO	< 0.6 ppm
Carbon Dioxide	CO ₂	< 480 ppm
Methane	CH ₄	< 8 ppm
Ethane	C ₂ H ₆	< 0.1 ppm
Acetylene	C ₂ H ₂	< 0.4 ppm
Ethylene	C ₂ H ₄	< 0.2 ppm
Propylene	C ₃ H ₆	< 0.2 ppm
Propane	C ₃ H ₈	< 0.05 ppm
Other Hydrocarbons	C ₄ +	< 0.05 ppm each
Hydrogen	H ₂	< 0.7 ppm
Hydrogen Sulfide	H ₂ S	< 0.001 ppm
Ammonia	NH ₃	< 0.01 ppm
Nitrous Oxide	N ₂ O	< 0.35 ppm
Nitrogen Oxides	NO _x	< 0.1 ppm
Ozone	O ₃	< 0.1 ppm
Sulfur Dioxide	SO ₂	< 0.1 ppm
Chloride	Cl	< 0.1 ppm
Total Strong Acid	HCl + NO ₃	< 0.05 ppm
Dust		< 0.2 mg/Nm ³

If, during the life of the plant, there are modifications in the environment conditions, which affect or could affect the design and the safe operation of the Plant, and in particular any modification in the air quality figures mentioned above, the Purchaser shall inform Air Liquide immediately of any such modifications.

Air Liquide shall inform Purchaser of the consequences (if any) of these modifications including cost impact (if any).

In case of peaks for C_nH_m in air that might exceed the limits indicated above, it may be necessary to shut down the Plant.

1.5 Make-Up Water Quality

Typical quality required for the make-up water is:

Table 5: Make-Up Water Quality

Property	Chemical Formula	Value
Supply Pressure		45 to 60 psig
Supply Temperature		< 64°F
pH value		6.5 to 7.5
Conductivity		500 µS/cm
Suspended solids		<3 mg/l
Iron	Fe	<0.05 mg/l
Manganese	Mn	<0.05 mg/l
Ammonia	NH ₄ ⁺	<2 mg/l
Nitrates + Nitrites	NO ₃ ⁻ + NO ₂ ⁻	50 mg/l
Silicates	SiO ₂	1 mg/l
Chloride ions	Cl ⁻	50 mg/l
Sulphate ions	SO ₄ ²⁻	150 mg/l

Note: The water must be free of living organisms, biological growth, algae.

1.6 Low-pressure Steam Quality

Typical quality required for the low-pressure steam:

Table 6: Low-pressure Steam Quality

Property	Units or Chemical Formula	Value
Supply Temperature	°F	370 (<i>Saturated</i>)
Supply Pressure	psig	160
pH value		7.0 to 9.5
Conductivity		<0.2 µS/cm
Silicates	SiO ₂	<0.02 mg/kg
Iron	Fe	<0.02 mg/kg
Copper	Cu	<0.003 mg/kg
Sodium	Na	<0.01 mg/kg
Organics		<0.2 mg/kg
Calcium + Magnesium	Ca + Mg	<0.05 mg/kg
Oxygen	O ₂	<0.25 mg/kg
Chloride ions	Cl ⁻	<0.1 mg/kg
Bromide ions	Br ⁻	<0.1 mg/kg
Iodide ions	I ⁻	<0.1 mg/kg
Fluoride ions	F ⁻	<0.02 mg/kg
Sulphate ions	SO ₄ ²⁻	<0.1 mg/kg
Solids		<1.0 mg/kg

1.7 Electrical Power Properties

Table 7: Electrical Power Properties

Property	Units	Value
Nominal Tension	kV	13.8 +/- 10 %
Frequency	Hz	60.0 +/- 0.5 Hz

1.8 Tie-Point Definition

The following are the defined Plant tie-points for piping. Tie-points are at the battery limit.

Table 8: Tie-Points

Tie-Point	Stream	Line Size (in.)	Pressure (psig)	Temp (°F)*
TP-1	LP Gaseous Oxygen	20	8	50
TP-2	MP Gaseous Oxygen	3	30	50
TP-3	Cooling Water Supply (flue gas cooler)	10	45	64
TP-4	Cooling Water Return (flue gas cooler)	12	30	94
TP-5	LP Saturated Steam Supply	2	160	188 (saturated)
TP-6	Condensate Return	2	145	TBD
TP-7	Make-up Water Supply	3	45	57
TP-8	Grey Water Return	2	TBD	TBD
TP-9	Drinking (Potable) Water	TBD	TBD	TBD
TP-10	Power Supply (13.8 kV)	n/a	n/a	n/a

*Approximate, at nominal conditions

1.9 Scheduled Maintenance

Over time, impurities in the air can enter the low temperature section of the ASU and begin to build up. From Air Liquide's operating experience, it is recommended that the ASU be derimed once every three (3) years to remove these impurities.

Other than for routine maintenance, there is no scheduled maintenance requirement of the ASU within the three (3) year derime period.

1.10 Plant Availability

The expected plant availability is greater than 98 % excluding maintenance periods. This availability figure assumes a normal skill level of operation and maintenance personnel, utilization of recommended spare parts and specific tools, and availability of utilities.

1.11 Load Change

The rate of oxygen production load change of 3% per minute is achieved with the proposed design with a fluctuation of oxygen purity of +/- 1.5% vol. during transient phases.

1.12 Emissions

Noise

Process vents and automatic valves are designed for a noise level limit of 85 dBA at the Plant battery limit. The noise level for major rotating equipment will exceed the noise limit. If required, attenuation walls, enclosures and/or buildings may be added to reduce noise level. These are currently not included.

Gas

Air separation is a clean process and does not generate environmentally-sensitive effluents under normal operating conditions. Typical process vents under normal operation include:

- N₂ with some contaminants of inlet air from adsorber reactivation and air pre-cooling operation.

During startup and shutdown periods, these process stream flows will be lower with the exception of the product header vents. The product header vents will blow-off off-spec gaseous oxygen. These vent rates will be lower than or equal to the rated production of the Plant and pose no concern to the environment.

Liquids

Condensate from the compression of humid air is collected and sent to the battery limit where it will be accepted and treated by Purchaser. Condensate from the steam reactivation heater is sent to the battery limit where it will be accepted and treated by Purchaser.

For a Plant shutdown lasting longer than forty-eight (48) hours, it is recommended that the cryogenic liquid contained within the distillation columns be drained to the Plant disposal header. From this piping system, the cryogenic liquid flows by gravity to the cryogenic disposal vaporizer.

Solids

No solid emissions are expected from normal operation of the Plant.

1.13 Manpower Requirements

The Plant will be designed for minimal staffing. The operator will be able to start, control and stop all major pieces of equipment in the Plant from the control room. The Plant is equipped with suitable electronic instruments, process analyzers, and control devices to ensure safe and efficient Plant operation.

SECTION 2: PROCESS DESCRIPTION

2.1 Air Compression

Atmospheric air is drawn through inlet air filters (F01-1/2) to remove particulate matter. The filtered atmospheric air is then compressed by two main air compressors (C01-1/2) which are driven by electric motors. The heat of compression from the main air compressors is removed in water cooled intercoolers.

2.2 Air Pre-Cooling

Warm air from the discharge of the main air compressor is cooled in a water-wash tower (E07). Cooling water is pumped by the water wash tower pumps (P60.1/2) into the water-wash tower. Another stream of water is sent to the top of the water chiller tower (E60). In this tower (E60), cooling water is cooled against dry waste nitrogen. The resulting chilled water coming off the bottom of E60 is pumped by the water chiller pumps (P61.1/2) to the water wash tower (E07). Inside E07, warm air from the discharge of the main air compressor is cooled through direct contact with the two water streams.

In addition to cooling the water, the water wash tower helps reduce or eliminate some gaseous impurities that may be present in the ambient air.

2.3 Air Purification

A front-end temperature swing adsorption-type (TSA) purification system is provided for the removal of water and carbon dioxide from the air feed. The system is made up of two radial flow bed vessels (R01/R02) containing activated alumina and molecular sieve adsorbents. By means of a set of automatically controlled switching valves, the compressed air passes alternately through one of the adsorbers or the other. This adsorption process removes moisture, carbon dioxide and some other impurities. The activated alumina removes the moisture and the molecular sieve removes the carbon dioxide contained in the air, as well as some of the other impurities.

At the end of every adsorption cycle, the air stream is redirected to the other bed and the vessel containing the adsorbed impurities is regenerated by a stream of dry, heated waste nitrogen (WN2). The waste nitrogen is heated by low-pressure saturated steam in a reactivation heater (E08). After it passes through the bed in regeneration, the gas is vented to the atmosphere. Once the adsorber has been heated for the desired amount of time, the vessel is cooled down and pressurized for the next adsorption phase.

2.4 Air Refrigeration

Dry compressed air exits the adsorber vessels and is split into three streams. Two streams enter the main heat exchanger (E01) and are cooled against the waste nitrogen and product streams to near the liquefaction temperature. After exiting the exchanger, one stream is sent directly to the medium-pressure (MP) column (K01) as feed. The second air stream liquefies against vaporizing GOX product. The liquid air stream is

flashed through a valve and the resulting two-phase stream is separated into liquid and vapor streams. The vapor stream enters the LP column (K02). The liquid stream is totally sent to the LP column.

The third stream goes to the booster compressor which is driven by the expander, through a water-cooled aftercooler and then introduced into the main exchanger (E01). At an intermediate location, air is extracted and sent to the expander. The expansion of this air stream provides the necessary refrigeration for the liquefaction of the air. After its expansion, the air enters the LP column as feed.

2.5 Cold Production and Distillation

Air sent to the HP column is separated into an oxygen-rich liquid at the bottom and almost pure nitrogen at the top. The separation of the air is carried out in this column, as in any other rectification column, by the interaction of vapor rising with a descending stream of reflux liquid over sections of structured packing materials. The descending liquid gradually becomes rich in oxygen and, at the same time, the vapors passing through the packing lose a corresponding small amount of oxygen at each pack until, when they reach the top pack, almost pure nitrogen is obtained. The gaseous nitrogen at the top of the HP column enters the side of the main vaporizer (E02) where it is condensed by liquid oxygen boiling on the other side of the heat transfer surface. The resulting liquid nitrogen falls by gravity from the top of the HP column where it acts as pure reflux. A liquid nitrogen stream is withdrawn from the top of the HP column and sent to the top of the LP column as reflux. Additionally, rich liquid is withdrawn from the bottom of the HP column and expanded to the LP column as feed.

The rectification in the LP column is the same as that in the HP column. Impure liquid nitrogen, liquid air, oxygen-rich liquids and the expanded air are fed into the appropriate levels of the LP column. Liquid arriving at the bottom is almost entirely oxygen (95% purity). The reboil duty for the LP column is provided by the main vaporizer, which by vaporizing liquid oxygen condenses the pure nitrogen at the top of the HP column. The main vaporizer is a bath-type vaporizer submerged in liquid oxygen drained from the bottom of the LP column.

Liquid oxygen (LOX) is extracted from the bottom of the LP column, vaporized, then warmed in the main heat exchanger (E01) and exits the coldbox as GOX product.

Low-pressure waste nitrogen gas (WN2) is extracted at the top of the LP then sent through the main heat exchanger (E01) where it cools down the incoming air. Part of the stream is then used to chill water in the nitrogen chiller tower (E60) while the remaining portion is sent to the steam reactivation heater (E08) and subsequently used to regenerate the adsorbers.

SECTION 3: SCOPE OF SUPPLY

3.1 Scope of Supply Definition

The following table defines the scope division between Air Liquide and **American Air Liquide** used for the cost estimates.

S: Supplier - Air Liquide

P: Purchaser – **American Air Liquide**

Table 9: Scope of Supply

Description	S	P	Remarks
Main Equipment and Materials			
Air Compressors and Accessories (C01-1/2)	X		2 x 50%, Electric Motors
Pre-cooling Equipment	X		
Purification skid (Including switching valves)	X		
Reactivation heater (E08)	X		LP Steam
Two air purification adsorbers with alumina and molecular sieve (R01/R02)	X		
Booster Air Compressors and accessories (C05)			n/a
Booster Air Compressor aftercooler (C05E)			n/a
Main cryogenic exchangers (E01)	X		
Cryogenic Liquid Pumps (P02-1/2)	X		
Distillation columns and cold boxes (K01/K02)	X		
Expansion turbine and compressor brake + aftercooler (ET01/ET01C)	X		
Cold box insulation	X		
All necessary field instruments and junction boxes	X		
Initial fillings for adsorbers and lube oil systems	X		
Cooling Water System (E62-1/2; P62-1/2)	X		Includes CW load for flue gas cooler
Liquid oxygen/Liquid nitrogen storage and associated back-up system			None included
Cryogenic Liquids Truck Loading system			None included
Pipes, pipe racks, and supports (excluding cooling water piping)	X		Inside battery limit. Includes 900 ft of piping for both oxygen pressure product requirements.
Cooling Water Piping	X		Inside battery limit.
Valves, safety valves	X		
Instrumentation	X		
Analyzers	X		
Instrumentation shelter	X		
DCS/ Operators stations	X		
Control strategy and DCS implementation	X		
Connection between battery limit and Purchaser control room for interface data (Interlocks / Control signals)		X	
HV sub station		X	
Motor Soft Starting Devices		X	None included. To be discussed later.
MV equipments and distribution	X		From 13.8 kV feeder provided by Purchaser
LV equipments and distribution	X		
Transformer	X		13.8 kV to 480V Only

Description	S	P	Remarks
UPS	X		For controls only
Emergency Power Generator		X	
Area Lighting		X	
Cables	X		Inside battery limit
Construction and commissioning spare parts	X		
Capital spare parts		X	None included
Civil Works (Cincinnati, OH basis)			
Soil investigation + report + Eventual decontamination		X	
Demolition of existing equipment or underground obstructions, and if any decontamination		X	
Site clearance and leveling		X	No overhead nor underground obstruction
Equipment foundations	X		
Pilings (3000 psf soil bearing assumed)		X	None included
Civil works inside battery limit	X		
Road, access to site		X	
Fencing		X	
Drainage system		X	
General grounding network	X		Inside battery limit
Electrical building	X		
Control / Analyzer Room	X		Combined w/ electrical building
Compressor building/Warehouse/Noise enclosures		X	None Included
Fire water networks		X	
Erection Works (Cincinnati, OH basis)			
Adjacent site for erection		X	
Works and access permits		X	
Necessary agreements		X	
Access to erection areas		X	
Utilities supply for erection works		X	
Utilities consumption for erection works		X	
Cleaning up of construction site & removal of temporary facilities	X		
Drinkable water & industrial water		X	
Storage of equipment	X		
Unloading of the supplied equipment	X		
Transportation	X		
Lifting and installation	X		
Mechanical erection	X		
Electrical works	X		
Instrumentation works	X		
Piping works	X		
Quality assurance & control	X		
Water for hydraulic tests		X	
Treatment/recovery of water w/ chemicals used for storage/pressure tests		X	As required
Supervision	X		
Guarding personnel		X	
All risks insurance (Builder's Risk Insurance)		X	
Commissioning and Start-up			
Necessary agreements		X	
Utilities consumption (electric power, make-up water, instrument air, LP steam)		X	
Supervision	X		
Experienced plant operators		X	
Performance test	X	X	

3.2 Exclusions

All work, equipment, and services not expressly mentioned as supplied by Air Liquide in previous chapters are excluded from the scope of this proposal, more particularly:

- Utility piping and associated structural steel and insulation outside the Plant battery limits namely the cooling water supply and return lines to/from the flue gas cooler.
- Instrumentation and electrical cables outside the Plant battery limits.
- DCS signal interconnection from the Plant to Purchaser's or other control systems.
- Any electrical substation.
- All import taxes, duties, VAT, etc.
- Any Purchaser engineering standards (proposal based on Air Liquide engineering standards).

SECTION 4: COST ESTIMATES

4.1 Cost Estimates

The cost estimates (+/- 20%) for the **Turnkey supply** of one (1) air separation units as specified in Section 1 through 3 (construction labor rates and productivity are on a Cincinnati, Ohio basis) and the associated engineering services based on AIR LIQUIDE Standards is:

\$25,500,000 United States Dollars

4.2 Currency Exchange Rates

The pricing of the proposal is based on currency exchange rates as commercially quoted on the date of the proposal. Air Liquide reserves the right to adjust the proposal prices for fluctuations in applicable exchange rates.

4.3 Typical Delivery Time

The typical turnkey delivery time for the scope of supply described above would be approximately **18 to 20 months**. This delivery time is based on current factory loading and can be further optimized at the time of the Nitrogen liquefier contract award.

4.4 Mechanical Guarantee

The mechanical guarantee period is 12 months of operation or 18 months after equipment delivery, whichever occurs first.

4.5 Taxes

Taxes, fees, bank or other charges to be levied during the implementation of the contract are excluded from the above-mentioned prices.

4.6 Terms of Payment

Our price is based on progressive payments in order to avoid any financial fee, and with cash payment conditions.

4.7 Validity

This proposal is valid for thirty (30) calendar days from date of issue.

4.8 General Commercial Conditions

To be discussed and mutually agreed upon by the executive management of both parties at a later date.

<div><div><div>AIR LIQUIDE</div><div>INGENIERIE</div></div><div>57 Av Carnot BP313 94503 Champigny cedex (FRANCE)</div></div>		Job nbr : 30257 Name : COH study								Document number 30257-013-110			
PROCESS DATA Ver 2.11 / A 1.2		HEAT AND MATERIAL BALANCE											
Stream ID Fluid Name		1 INLET AIR	2 MP AIR	3 WN2	4 WN2	5 GOX	6 GOX						
Composition, dry	molar %	100.00	100.00	100.00	100.00	100.00	100.00						
N2		78.11	78.11	99.53	99.53	0.45	1.54						
Ar		0.93	0.93	0.23	0.23	2.37	3.47						
O2		20.96	20.96	0.25	0.25	97.18	94.99						
Partial flowrate, dry	kgmole/h												
N2		75,735	75,735	60,786	13,881	2	317						
Ar		902	902	138	31	9	715						
O2		20,323	20,323	151	35	349	19,586						
Total flowrate, dry	Nm3/h	96,960	96,960	61,075	13,947	359	20,619						
Total flowrate, dry	kgmole/h	4,326	4,326	2,725	622	16	920						
Pressure	psia	14.420	78.50	14.420	15.00	44.42	22.50						
Temperature	°F	59.0	183.0	60.8	50.0	50.0	50.0						
Liquid Fraction		0.00	0.00	0.00	0.00	0.00	0.00						
Molar Weight, dry	kg/kgmole	28.96	28.96	28.05	28.05	32.17	32.21						
Relative Humidity	%	60.0%	8.8%	0.0%	0.0%	0.0%	0.0%						
Mass Flow, H2O	kg/h	811	715	0	0	0	0						
Revision	DATE	BY		VISA		CHECKED		VISA					
a	20-Mar-06	Chen, Guofu											
b													
1													
2													
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EQUIPMENT LIST

EQUIP. #	DESCRIPTION
CB#1	EXCHANGER COLD BOX
CB#2	COLUMN COLD BOX
F01.1/2	INLET AIR FILTERS
C01.1/2	MAIN AIR COMPRESSORS
E07	WATER WASH TOWER
R01/R02	ADSORPTION VESSELS
E08	REACTIONATION HEATER
E09	SPECIAL REACTIONATION HEATER
E60	WATER CHILLER TOWER
P60.1/2	WATER CHILLER TOWER PUMPS
P61.1/2	WATER WASH TOWER PUMPS
E13	CLOSED LOOP WATER EXCHANGER
ET01/ET01C	EXPANDER/BOOSTER
ET01CE	BOOSTER AFTERCOOLER
P02.1/2	LOX PUMPS
E40/E80	LOX/PURGE VAPORIZER
E62.1/2	COOLING WATER TOWERS
P62.1/2	COOLING WATER PUMPS
XFMR	TRANSFORMER 13.8 kV TO 480 V
-	AIR PURIFICATION SKID

TIE-IN POINTS

TIE-IN	STREAM
TP-1	LP GASEOUS OXYGEN
TP-2	MP GASEOUS OXYGEN
TP-3	COOLING WATER SUPPLY (FLUE GAS COOLER)
TP-4	COOLING WATER RETURN (FLUE GAS COOLER)
TP-5	LP SATURATED STEAM SUPPLY
TP-6	CONDENSATE RETURN
TP-7	MAKE-UP WATER SUPPLY
TP-8	GREY WATER RETURN
TP-9	DRINKING (POTABLE) WATER
TP-10	POWER SUPPLY (13.8 kV)

REV	DATE	REVISION DESCRIPTION	BY	CHK	APP	REV	DATE	REVISION DESCRIPTION	BY	CHK	APP
1	4/19/08	ISSUED FOR PROPOSE									
DESIGNER: Dept. of Energy (DOE) NAME/LOC.: Organometallic Technology for CO2 NAME/LOC.: Capture in Pulverized Coal-Fired Boiler NAME/LOC.: DOE Project Task #6 PROJECT NO.: 30257 CUST. No.:											
682.6 MTD ASU						PLOT PLAN					
PRELIMINARY											
SCALE: 1"=10' 1/2"						SHEET NO. OF 10					
DATE: 4/19/08						DRAWING NO. OF PAGE NO.					
						REV.					
						PP-30257-001					
						A					

Job No.: 30257
Revision No.: B
By: DC
Date: 04/11/06

Preliminary Motor/Load List

Description: *Pilot Plant for Oxycombustion Study - City of Hamilton*
Customer: Air Liquide America - B&W
Location: Hamilton, OH

Tag Number		Equipment	Voltage (Volts)	Rated Load HP	Operating Load Kw	Service	
						Continuous	Stand-By
XFMR-1		4,160 V / 480 V Transformer	4160		TBD	X	
C01-1		Main Air Compressor "1"	4160	5,500	3,540	X	
LOP-C01-1		Main Air Compressor "1" Lube Oil Pump (Auxiliary)	480	7.5			X
LOH-C01-1A		Main Air Compressor "1" Lube Oil Heater A	480		4	X (intermitent)	
LOH-C01-1B		Main Air Compressor "1" Lube Oil Heater B	480		4	X (intermitent)	
VB-C01-1		Main Air Compressor "1" Mist Eliminator	480	1		X	
C01-2		Main Air Compressor "2"	4160	5,500	3,540	X	
LOP-C01-2		Main Air Compressor "2" Lube Oil Pump (Auxiliary)	480	7.5			X
LOH-C01-2A		Main Air Compressor "2" Lube Oil Heater A	480		4	X (intermitent)	
LOH-C01-2B		Main Air Compressor "2" Lube Oil Heater B	480		4	X (intermitent)	
VB-C01-2		Main Air Compressor "2" Mist Eliminator	480	1		X	
LOP-ET01-1		ET01 Lube Oil Pump Motor (Main)	480	3		X	
LOP-ET01-2		ET01 Lube Oil Pump Motor (Auxiliary)	480	3			X
LOH-ET01		ET01 Lube Oil Heater	480		3	X (intermitent)	
VB-ET01		ET01 Mist Eliminator	480	0.5		X	
P02-1		Low Pressure LOX Pump "1"	480	0.5	0.2	X	
P02-2		Low Pressure LOX Pump "2"	480	0.5	0.2		X
P61-1		Water Wash Tower Pump "1"	480	40	25	X	
P61-2		Water Wash Tower Pump "2"	480	40	25		X
P60-1		Water Chiller Pump "1"	480	35	21	X	
P60-2		Water Chiller Pump "2"	480	35	21		X
P62-1		Cooling Water Pump "1"	480	250	156	X	
P62-2		Cooling Water Pump "2"	480	250	156		X
E62-1		Cooling Water Tower Fan "1"	480	60	36	X	
E62-2		Cooling Water Tower Fan "2"	480	60	36	X	
E-81		Derime Heater	480	n/a	45		X
E-08		Regeneration Heater (STEAM REGENERATION HEATER)	n/a	n/a	n/a	n/a	n/a
UPS-1		UPS Primary Feeder	480		TBD		
HVAC-1		PCR HVAC	480		TBD		
HVAC-2		PCR HVAC	480		TBD		
LP-02		Plant Lighting Panel	240/120		TBD		
DP-01		Plant 480 V Distribution Panel	480		TBD

SFT

SFT, Inc
Consulting
Engineers

SIGNATURE	DATE
DR. P. H. PETERSON	02-20-9
RE. D. D. D.	02-20-9
DR. J. J. J.	02-20-9
DR. R. R. R.	02-20-9

1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33	34	35	36	37	38	39	40	41	42	43	44	45	46	47	48	49	50	51	52	53	54	55	56	57	58	59	60	61	62	63	64	65	66	67	68	69	70	71	72	73	74	75	76	77	78	79	80	81	82	83	84	85	86	87	88	89	90	91	92	93	94	95	96	97	98	99	100	101	102	103	104	105	106	107	108	109	110	111	112	113	114	115	116	117	118	119	120	121	122	123	124	125	126	127	128	129	130	131	132	133	134	135	136	137	138	139	140	141	142	143	144	145	146	147	148	149	150	151	152	153	154	155	156	157	158	159	160	161	162	163	164	165	166	167	168	169	170	171	172	173	174	175	176	177	178	179	180	181	182	183	184	185	186	187	188	189	190	191	192	193	194	195	196	197	198	199	200	201	202	203	204	205	206	207	208	209	210	211	212	213	214	215	216	217	218	219	220	221	222	223	224	225	226	227	228	229	230	231	232	233	234	235	236	237	238	239	240	241	242	243	244	245	246	247	248	249	250	251	252	253	254	255	256	257	258	259	260	261	262	263	264	265	266	267	268	269	270	271	272	273	274	275	276	277	278	279	280	281	282	283	284	285	286	287	288	289	290	291	292	293	294	295	296	297	298	299	300	301	302	303	304	305	306	307	308	309	310	311	312	313	314	315	316	317	318	319	320	321	322	323	324	325	326	327	328	329	330	331	332	333	334	335	336	337	338	339	340	341	342	343	344	345	346	347	348	349	350	351	352	353	354	355	356	357	358	359	360	361	362	363	364	365	366	367	368	369	370	371	372	373	374	375	376	377	378	379	380	381	382	383	384	385	386	387	388	389	390	391	392	393	394	395	396	397	398	399	400	401	402	403	404	405	406	407	408	409	410	411	412	413	414	415	416	417	418	419	420	421	422	423	424	425	426	427	428	429	430	431	432	433	434	435	436	437	438	439	440	441	442	443	444	445	446	447	448	449	450	451	452	453	454	455	456	457	458	459	460	461	462	463	464	465	466
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SITE PLAN

[illegible]

REVISIONS			
REV	DESCRIPTION	DATE	APPROVAL
01	REVISED PER ENGINEERING UPDATE, /JA HENDRIX	06.29.06	AA LOGUE
02	REMOVED PROPRIETARY STATEMENTS FROM TITLE BLOCK	03.11.08	J. CRONIN

- NOTES:**
- FOR DRAWING INDEX, SYMBOLS AND NOMENCLATURE, SEE B&W DRAWINGS 544977E THRU 554979E
 - * - INDICATES ITEMS SUPPLIED LOOSE BY EQUIPMENT SUPPLIER. (INSTALLED BY B&W)
 - OXIDATION AIR DISCHARGE PIPING UP TO HUMIDIFICATION POINT TO BE INSULATED ONLY IN THOSE LOCATIONS THAT ARE ACCESSABLE TO PERSONNEL.
 - ALL TAG NUMBERS ARE PREFIXED BY A NUMBER 8 (UNIT NO. 8) UNLESS NOTED OTHERWISE.
 - LOCATE AT LOW POINT.
 - FOR COMPRESSOR LUBE OIL SYSTEM SEE DRAWING 554993E.
 - HARDWIRED COMMUNICATION TO/FROM THE PLC AND DCS ARE NOT SHOWN ON THIS DRAWING. THESE HARDWIRED SIGNALS ARE INPUTS AND OUTPUTS OF THE COMPRESSOR MOTORS. ALL OTHER SIGNALS ARE AVAILABLE FROM THE PLC TO THE DCS VIA ETHERNET CONNECTIONS.

B&W PROPOSAL- PRELIMINARY USE ONLY!

895-1463 03/11/08
CITY OF HAMILTON
HAMILTON, OH
UNIT #8

DWN	JA HENDRIX
CHK'D	AA LOGUE
APPR'D	
DATE	03.28.06
CONTRACT NO.	895-1463
ORG. CONTRACT NO.	
- DO NOT SCALE - USE DIMENSIONS ONLY	

The Babcock & Wilcox Company a McDermott company			
PROCESS & INSTRUMENTATION DIAGRAM OXIDATION AIR SYSTEM			
SCALE	CC-SECT	DWG. NO.	REV
NONE	8570	554992E	02

REVISIONS			
REV	DESCRIPTION	DATE	APPROVAL
01	REVISED PER ENGINEERING UPDATE	06.29.06	AA LOGUE
02	REMOVED PROPRIETARY STATEMENTS FROM TITLE BLOCK	03.11.08	J. CRONIN

- NOTES:**
1. FOR DRAWING INDEX, SYMBOLS AND NOMENCLATURE, SEE B&W DRAWINGS 554977E THRU 554979E.
 2. ALL TAG NUMBERS ARE PREFIXED BY A NUMBER 8 (UNIT NO. 8) UNLESS NOTED OTHERWISE.
 3. 1/4" RUBBER LINED.

B&W PROPOSAL- PRELIMINARY USE ONLY!

895-1463 03/11/08
CITY OF HAMILTON
HAMILTON, OH
UNIT #8

DWN	JA HENDRIX
CHK'D	AA LOGUE
APPR'D	
DATE	03.28.06
CONTRACT NO.	895-1463
ORG. CONTRACT NO.	
- DO NOT SCALE - USE DIMENSIONS ONLY	

The Babcock & Wilcox Company <small>a McDermott company</small>			
PROCESS & INSTRUMENTATION DIAGRAM VACUUM BELT FILTER SYSTEM A SHEET 3 OF 4			
SCALE	CC-SECT	DWG. NO.	REV.
NONE	8570	554989E	02

REVISIONS			
REV	DESCRIPTION	DATE	APPROVAL
01	REVISED PER ENGINEERING UPDATE, /JA HENDRIX	06.29.06	AA LOGUE
02	REMOVED PROPRIETARY STATEMENTS FROM TITLE BLOCK	03.11.08	J. CRONIN

- NOTES:**
- FOR DRAWING INDEX, SYMBOLS AND NOMENCLATURE, SEE B&W DRAWINGS 554977E THRU 554978E.
 - ALL TAG NUMBERS ARE PREFIXED BY A NUMBER 8 (UNIT NO. 8) UNLESS NOTED OTHERWISE.

B&W PROPOSAL- PRELIMINARY USE ONLY!

895-1463 03/11/08
CITY OF HAMILTON
HAMILTON, OH
UNIT #8

DWN	TJ ELLIOTT
CHK'D	AA LOGUE
APPR'D	
DATE	03.28.06
CONTRACT NO.	895-1463
ORG. CONTRACT NO.	
- DO NOT SCALE - USE DIMENSIONS ONLY	

The Babcock & Wilcox Company <small>a McDermott company</small>			
PROCESS & INSTRUMENTATION DIAGRAM VACUUM BELT FILTER A SHEET 1 OF 4			
SCALE	CC-SECT	DWG. NO.	REV
NONE	857	554987E	02

REVISIONS			
REV	DESCRIPTION	DATE	APPROVAL
01	REVISED PER ENGINEERING UPDATE, /JA HENDRIX	06.29.06	AA LOGUE
02	REMOVED PROPRIETARY STATEMENTS FROM TITLE BLOCK	03.11.08	J. CRONIN

- NOTES:**
- FOR DRAWING INDEX, SYMBOLS AND NOMENCLATURE, SEE B&W DRAWINGS 554977E THRU 554979E.
 - ALL TAG NUMBERS ARE PREFIXED BY A NUMBER 8 (UNIT NO. 8) UNLESS NOTED OTHERWISE.

B&W PROPOSAL- PRELIMINARY USE ONLY!

895-1463 03/11/08
CITY OF HAMILTON
HAMILTON, OH
UNIT #8

DWN	JA HENDRIX
CHK'D	AA LOGUE
APPR'D	
DATE	03.28.06
CONTRACT NO.	895-1463
ORIG. CONTRACT NO.	
- DO NOT SCALE - USE DIMENSIONS ONLY	

The Babcock & Wilcox Company
a McDermott company

**PROCESS & INSTRUMENTATION
DIAGRAM**
**PRIMARY & SECONDARY
HYDROCLONES**

SCALE	CC-SECT	DWG. NO.	REV
NONE	8570	554986E	02

REVISIONS			
REV	DESCRIPTION	DATE	APPROVAL
01	REVISED PER ENGINEERING UPDATE, /JA HENDRIX	06.29.06	AA LOGUE
02	REMOVED PROPRIETARY STATEMENTS FROM TITLE BLOCK	03.11.08	J. CRONIN

- NOTES:**
1. FOR DRAWING INDEX, SYMBOLS AND NOMENCLATURE, SEE B&W DRAWINGS 554977E THRU 554979E.
 2. ALL TAG NUMBERS ARE PREFIXED BY A NUMBER 8 (UNIT NO. 8) UNLESS NOTED OTHERWISE.
 3. SEE DRAWING 554994E FOR GEAR BOX COOLING WATER.

B&W PROPOSAL- PRELIMINARY USE ONLY!

895-1463 03/11/08
CITY OF HAMILTON
HAMILTON, OH
UNIT #8

DWN	JA HENDRIX	The Babcock & Wilcox Company a McDermott company	
CHK'D	AA LOGUE		
APPR'D		PROCESS & INSTRUMENTATION DIAGRAM ABSORBER TOWER RECIRCULATION PUMPS	
DATE	03.28.06		
CONTRACT NO.	895-1463		
ORG. CONTRACT NO.			
- DO NOT SCALE - USE DIMENSIONS ONLY		SCALE	CC-SECT
		NONE	8570
		DWG. NO.	554982E
		REV	02

REVISIONS			
REV	DESCRIPTION	DATE	APPROVAL
01	REVISED PER ENGINEERING UPDATE, /JA HENDRIX	06.29.06	AA LOGUE
02	REMOVED PROPRIETARY STATEMENTS FROM TITLE BLOCK	03.11.08	J. CRONIN

- NOTES:
1. FOR DRAWING INDEX, SYMBOLS AND NOMENCLATURE, SEE B&W DRAWINGS 554977E THRU 554979E.
 2. ALL TAG NUMBERS ARE PREFIXED BY A NUMBER 8 (UNIT NO. 8) UNLESS NOTED OTHERWISE.

B&W PROPOSAL- PRELIMINARY USE ONLY!

895-1463 03/11/08
CITY OF HAMILTON
HAMILTON, OH
UNIT #8

DWN	JA HENDRIX
CHK'D	AA LOGUE
APPR'D	
DATE	03.28.06
CONTRACT NO.	895-1463
ORG. CONTRACT NO.	
- DO NOT SCALE - USE DIMENSIONS ONLY	

The Babcock & Wilcox Company
a McDermott company

PROCESS & INSTRUMENTATION
DIAGRAM
REAGENT FEED SYSTEM

SCALE	CC-SECT	DWG. NO.	REV
NONE	8570	554981E	02

REVISIONS			
REV	DESCRIPTION	DATE	APPROVAL

MISCELLANEOUS

PIPING SPECIALTIES

INSTRUMENTS

SERVICE ABBREVIATIONS

AB – ABSORBER RECIRC	SA – SERVICE AIR
AQ – ABSORBER QUENCH	SS – HYDROCLONE UNDERFLOW
BA – FLUE GAS/BOILER AIR	ST – STEAM
ED – EQUIPMENT DRAIN/VENTS	SV – FILTRATE WATER/VACUUM SYSTEM
HF – HYDRAULIC FLUID	WB – ABSORBER BLEED
IA – INSTRUMENT AIR	WC – CLOSED COOLING/COOLING WATER
LH – LIMESTONE HANDLING	WK – MIST ELIMINATOR WASH WATER
LO – LUBE OIL	WN – RECLAIM WATER
LS – LIMESTONE SLURRY	WS – SERVICE WATER
OA – OXIDATION AIR	WT – WASTE WATER

MATERIAL ABBREVIATIONS

CA1D – CLASS 150, ANSI B31.1, ALLOY C276 (UNS N10276), STD WT

CC1B – CLASS 150, ANSI B31.1, ALLOY 225 (UNS S32550), SCH 10, 14" AND LARGER

CC1C – CLASS 150, ANSI B31.1, ALLOY 255 (UNS S32550), SCH 20, 2 1/2" THRU 6"

CC1I – CLASS 150, ANSI B31.1, ALLOY 255 (UNS S32550), SCH 80, 2" AND BELOW

CC4B – CLASS 150, ANSI B31.1, ALLOY 2205 (UNS S32205), SCH 10, 14" AND LARGER

CC4F – CLASS 150, ANSI B31.1, ALLOY 2205 (UNS S32205), SCH 40, 2 1/2" THRU 6"

CC4I – CLASS 150, ANSI B31.1, ALLOY 2205 (UNS S32205), SCH 80, 2" AND BELOW

CD1B – CLASS 150, ANSI B31.1, 304 STAINLESS STEEL (UNS S30400), SCH 10, 12" AND LARGER

CD1D – CLASS 150, ANSI B31.1, 304 STAINLESS STEEL (UNS S30400), STD WT, 2 1/2" – 10"

CD1I – CLASS 150, ANSI B31.1, 304 STAINLESS STEEL (UNS S30400), SCH 80, 2" AND SMALLER

CG1F – CLASS 150, ANSI B31.1, 316L STAINLESS STEEL (UNS S31603), SCH 40, 2 1/2" AND LARGER

CG1I – CLASS 150, ANSI B31.1, 316L STAINLESS STEEL (UNS S31603), SCH 80, 2" AND SMALLER

CL1D – CLASS 150, ANSI B31.1, A53B CARBON STEEL, STD WT, 2 1/2" AND LARGER

CL1G – CLASS 150, ANSI B31.1, A53B CARBON STEEL, SCH 80, 2" AND SMALLER

CM1D – CLASS 150, ANSI B31.1, A53B CARBON STEEL, STD WT, NATURAL RUBBER LINED 3/16", 2" AND LARGER

CP3D – CLASS 150, ANSI B31.1, A53B CARBON STEEL, STD WT, URATHANE LINED, 2" AND LARGER

CP6D – CLASS 150, ANSI B31.1, A53B CARBON STEEL, STD WT, NI-HARD LINED, 2" AND LARGER

CS4D – CLASS 150, ANSI B31.1, ASTM B-466 COPPER (UNS C70600), STD WT, ALL SIZES

CV1D – CLASS 150, ANSI B31.1, SOLID NI HARD, STD SCHEDULE

CV2D – CLASS 150, ANSI B31.1, SOLID CERAMIC, STD WT.

GC1B – CLASS 150, ANSI B31.3, ALLOY 255 (UNS S32550), SCH 10, 8" THRU 12"

GC4B – CLASS 150, ANSI B31.3, ALLOY 2205 (UNS S32205), SCH 10, 8" THRU 12"

QE1M – CLASS 2500, ANSI B36.1, 304L STAINLESS STEEL (UNS S30403), SCH 160, 2" AND SMALLER

RLGZ – DEAD SOFT, COLD DRAWN, SEAMLESS TUBING, SAE1010 STEEL CONFORMING TO SAE J524, SAE J525 AND SAE J526, WITH A 0.095 WALL HAVING A BURST PRESSURE OF 17,000 PSI.

VT1 – CLASS 150, ASTM D2310, FIBERGLASS, SUPPLIERS STD WT, ALL SIZES

VT2 – CLASS 150, ASTM D2310, FIBERGLASS, SUPPLIERS STD WT, ABRASION RESISTANT INNER LINER, ALL SIZES

LINE LEGEND

INSTRUMENT IDENTIFICATION LETTERS

	FIRST LETTER		SUCCEEDING LETTERS		
	MEASURED OR INITIATING VARIABLE	MODIFIER	READOUT OR PASSIVE FUNCTION	OUTPUT FUNCTION	MODIFIER
A	Analysis		Alarm		
B	Burner, Combustion		User's Choice	User's Choice	User's Choice
C	Conductivity			Controller	Closed
D	Density	Differential			
E	Voltage		Sensor (Primary Element)		
F	Flow Rate	Ratio (Fraction)			
G	User's Choice		Glass, Viewing Device		
H	Hand				High
I	Current (Electric)		Indicate		
J	Power	Scan			
K	Time, Time Schedule	Time Rate of Change		Control Station	
L	Level		Light		Low
M	Moisture (Humidity)	Momentary			Middle, Intermediate
N	User's Choice		User's Choice	User's Choice	User's Choice
O	User's Choice		Orifice, Restriction		Open
P	Pressure, Vacuum		Point (Test) Connection		
Q	Quantity	Integrate, Totalize			
R	Radiation		Record		
S	Speed, Frequency	Safety		Switch	
T	Temperature			Transmit	
U	Multivariable		Multifunction	Multifunction	Multifunction
V	Vibration, Mechanical Analysis			Valve, Damper, Louver	
W	Weight, Force		Well		
X	Unclassified	X Axis	Unclassified	Unclassified	Unclassified
Y	Event, State or Presence	Y Axis		Relay, Compute, Convert	
Z	Position, Dimension	Z Axis		Driver, Actuator, Unclassified Final Control Element	

NOTE:

FOR DRAWING INDEX, SEE B&W DRAWING 554977E.

<i>B&W PROPOSAL- PRELIMINARY USE ONLY!</i> <div style="border: 1px solid black; padding: 10px; text-align: center;"> 895-1463 03/11/08 CITY OF HAMILTON HAMILTON, OH UNIT #8 </div>	DWN CHND APPVD	JA HENDRIX AA LOGUE	The Babcock & Wilcox Company a McDermott company			
			PROCESS & INSTRUMENTATION DIAGRAM SYMBOLS & NOMENCLATURE SHEET 1 OF 2			
	DATE 03.28.06					
	CONTRACT NO. 895-1463					
	ORIG. CONTRACT NO.		SCALE CC-SECT NONE 8570		DWG. NO. 554978E	REV. 01
- DO NOT SCALE - USE DIMENSIONS ONLY

B&W PROPOSAL- PRELIMINARY USE ONLY!

895-1463 03/11/08
CITY OF HAMILTON
HAMILTON, OH
UNIT #8

DWN: J.J.YODER & J.L.M.
CHK'D: W.D.ZIEGLER
APP'D: J.CRONIN
DATE: 04/20/06
CONTRACT NO.: 895-1463
ORG. CONTRACT NO.: 113-0146
1/8" = 1'-0"
USE DIMENSIONS ONLY

The Babcock & Wilcox Company
A B&W COMPANY

PROPOSAL
CONVERT UNIT TO
OXY-FIRING

SCALE: 1/8" = 1'-0"
CC-SECT: 683-04
DWG. NO.: 81820J
REV: 01

REVISIONS		
REV	DESCRIPTION	DATE

<div>895-1463-10-53 ACID PART NAME: 81819J.dwg</div> <div>The Babcock & Wilcox Company a McDermott company</div>		<div>PROPOSAL CONVERT UNIT TO OXY-FIRING</div>	
		DATE: 04/17/06	
		CONTRACT NO.: 895-1463	
		ORG. CONTRACT NO.: 113-0146	
DWG. NO.: 81819J		SCALE: 1/8" = 1'-0"	
REV: 01		CC-SECT: 683-04	
DWG. NO.: 81819J		REV: 01	

B&W PROPOSAL- PRELIMINARY USE ONLY!

895-1463 03/11/08
CITY OF HAMILTON
HAMILTON, OH
UNIT #8

DWG. NO.	81819J
REV.	01

B&W PROPOSAL- PRELIMINARY USE ONLY: <div><div>895-146303/11/08</div><div>CITY OF HAMILTON</div><div>HAMILTON, OH</div><div>UNIT #8</div></div>	DRN	J.J.YODER & J.L.M.	The Babcock & Wilcox Company a McWorrel company			
	CHIC	W.D.ZIEGLER				
	APVD	J.CRONN				
	DATE	D4/17/06				
	CONTRACT NO.	895-1463				
	ORIG. CONTRACT NO.	113-0146				
	- DO NOT SCALE - USE DIMENSIONS ONLY		SCALE	CC-SECT	DWG. NO.	REV.
			1/8" = 1'-0"	683-04		81817J01
			30-Contour-Jay 12/15/05

REVISIONS			
REV	DESCRIPTION	DATE	APPROVAL

21	DWG. NO. 78727J	02
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B&W PROPOSAL- PRELIMINARY USE ONLY!		DWN J.J.YODER & J.L.M. CHK'D W.D.ZIEGLER APP'D J.CRONIN	The Babcock & Wilcox Company <small>a MacArthur company</small>	
<div>895-1463 03/11/08 CITY OF HAMILTON HAMILTON, OH UNIT #8</div>		DATE 05/18/06	PROPOSAL CONVERT UNIT TO OXY-FIRING	
		CONTRACT NO. 895-1463 ORG. CONTRACT NO. 113-0146 <small>* DO NOT SCALE - USE DIMENSIONS ONLY</small>		
SCALE 1/8" = 1'-0"		CC-SECT 683-04	DWG. NO. 78727J	REV. 02

D'-D'

SECTIONAL SIDE ELEVATION S-C-C

DOOR CHART				
SIZE	DOOR	DOOR	DOOR	DOOR
12'0"	12'0" X 10'0"	12'0" X 10'0"	12'0" X 10'0"	12'0" X 10'0"
12'0"	12'0" X 10'0"	12'0" X 10'0"	12'0" X 10'0"	12'0" X 10'0"
12'0"	12'0" X 10'0"	12'0" X 10'0"	12'0" X 10'0"	12'0" X 10'0"
12'0"	12'0" X 10'0"	12'0" X 10'0"	12'0" X 10'0"	12'0" X 10'0"

Field Change
FOR ADDED PIPE TAP CONN. REFER TO M.J. ZALLEN LETTER 6-4-65
FOR ROUTING M.C. PIPING REFER TO C.F. SCHULZ LETTER 5-6-65 & 5-7-65
FOR DAMPER OUTLET FLANGE WELDS REFER TO C.F. SCHULZ LETTER 7-27-65 & C.E. DUNN LETTER 7-20-65

REVISION
NO. 1
DATE
BY
CHECKED
APPROVED

- 1. THIS DRAWING IS TO BE USED FOR THE DESIGN AND CONSTRUCTION OF THE BOILER AND ALL ACCESSORIES THEREON. IT IS THE RESPONSIBILITY OF THE ENGINEER TO SEE THAT THE BOILER IS CONSTRUCTED IN ACCORDANCE WITH THE REQUIREMENTS OF THE ASME CODE AND ALL APPLICABLE LAWS AND ORDINANCES.
- 2. THE BOILER IS TO BE DESIGNED FOR A WORKING PRESSURE OF 150 PSIG AND A TEMPERATURE OF 350°F. IT IS TO BE PROVIDED WITH ALL NECESSARY SAFETY DEVICES, INCLUDING BUT NOT LIMITED TO, SAFETY VALVES, WATER LEVEL INDICATORS, AND LOW WATER CUT-OFFS.
- 3. THE BOILER IS TO BE PROVIDED WITH A COMPLETE SET OF PIPING, INCLUDING BUT NOT LIMITED TO, WATER INLET, STEAM OUTLET, AND DRAINAGE. ALL PIPING IS TO BE PROVIDED WITH THE NECESSARY FLANGES, BOLTS, AND NUTS.
- 4. THE BOILER IS TO BE PROVIDED WITH A COMPLETE SET OF ELECTRICAL CONNECTIONS, INCLUDING BUT NOT LIMITED TO, POWER, CONTROL, AND SAFETY. ALL ELECTRICAL CONNECTIONS ARE TO BE PROVIDED WITH THE NECESSARY WIRING, TERMINALS, AND GROUNDING.
- 5. THE BOILER IS TO BE PROVIDED WITH A COMPLETE SET OF ACCESSORIES, INCLUDING BUT NOT LIMITED TO, SAFETY VALVES, WATER LEVEL INDICATORS, AND LOW WATER CUT-OFFS. ALL ACCESSORIES ARE TO BE PROVIDED WITH THE NECESSARY FLANGES, BOLTS, AND NUTS.

113-0146-13	SECTIONAL SIDE VIEW
113-0146-20	20'-0" X 14'-0" WIDE S.P. BOILER
113-0146-21	PULVERIZED COAL OIL
113-0146-39	113-0146-39
113-0146-39	113-0146-39

REVISIONS		
REV	DESCRIPTION	DATE
0	ISSUED FOR ESTIMATE	4/17/06

CITY OF HAMILTON	
BABCOCK & WILCOX A McDermott Company	
PROPOSAL MECHANICAL AIR FLOW DIAGRAMS	
DATE: 03/13/06	CONTRACT NO. 435-0004
SCALE: NONE	CC-SECT: NONE
DATE: 03/13/06	REV: 0

REVISIONS			
REV	DESCRIPTION	DATE	APPROVAL

PRELIMINARY

CITY OF HAMILTON
UNIT 8

The Babcock & Wilcox Company
a McDermott company

PROCESS &
INSTRUMENTATION
DIAGRAM
FLUE SYSTEM

DWN	AM TRAUGH
CHK'D	GJ KORNAKER
APPR'D	
DATE	03.28.05
CONTRACT NO.	895-1463
ORIG. CONTRACT NO.	

SCALE	CC-SECT	DWG. NO.	REV.
NA	6571	559917E	00

- NOTES
1. FOR PROCESS & INSTRUMENTATION DIAGRAM- SYMBOLS AND NOMANCLATURE SEE B&W DRAWING 554978E.
 2. * EXISTING EQUIPMENT.

O₂ SUPPLY
TO BURNERS BA1 & BA2
AMT01E

APPENDIX – TASK 7: OXY-FUEL COMBUSTION SYSTEM ANALYSIS FOR NEW POWER PLANT APPLICATION

DOE has published the results of the Task 7 in a separate report. The reference of the report is given below:

“Pulverized coal Oxycombustion Power Plants, Volume 1: Bituminous coal to electricity”
DOE/NETL-2007/1291, October 2007.

The report can also be downloaded at the URL (July 2008): <http://www.netl.doe.gov/energy-analyses/pubs/PC%20Oxyfuel%20Combustion%20Volume%20I%20Final%20Revision%201.pdf>