**GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING
IN CIRCULATING FLUIDIZED BED BOILERS:**

**PHASE II – PILOT SCALE TESTING AND UPDATED
PERFORMANCE AND ECONOMICS FOR OXYGEN FIRED CFB
WITH CO₂ CAPTURE**

FINAL TECHNICAL PROGRESS REPORT

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PUBLIC ABSTRACT

Given that fossil fuel fired power plants are among the largest and most concentrated producers of CO₂ emissions, recovery and sequestration of CO₂ from the flue gas of such plants has been identified as one of the primary means for reducing anthropogenic CO₂ emissions.

In 2001, ALSTOM began a two-phase program to investigate the feasibility of various carbon capture technologies. This program is sponsored under a Cooperative Agreement from the US Department of Energy's National Energy Technology Laboratory.

Phase I entailed a comprehensive study evaluating the technical feasibility and economics of alternate CO₂ capture technologies applied to Greenfield US coal-fired electric generation power plants. Thirteen cases, representing various levels of technology development, were evaluated. Seven cases represent coal combustion cases in CFB type equipment. Four cases represent Integrated Gasification Combined Cycle (IGCC) systems. Two cases represent advanced Chemical Looping systems. Details of this work have been reported by Marion, *et al*, (2003).

One of the cases studied was the oxygen-fired circulating fluidized bed (CFB) concept. When fired with oxygen, the flue gas from the CFB contains over 70% CO₂. This flue gas can be processed to produce over 90% CO₂ for sequestration or use in enhanced oil recovery (EOR). The Phase I study identified the O₂-fired CFB as having a near term development potential.

Phase II - which is the subject of this report - consisted of pilot-scale testing followed by a refined performance and economic evaluation of the oxygen-fired CFB concept. As a part of this workscope, ALSTOM modified its 9.9 MM-Btu/hr Multiuse Test Facility (MTF) pilot plant to operate with O₂/CO₂ mixtures of up to 70 % O₂ by volume. Tests with coal and petroleum coke were conducted. The test objectives were to determine the impacts of oxygen firing on heat transfer, bed dynamics, potential agglomeration, and gaseous and particulate emissions. The test data was used to refine the design, performance, costs, and economic models developed in Phase-I for an O₂-fired CFB with CO₂ capture.

Key results from the MTF testing are summarized below.

- The furnace was successfully operated on bituminous coal and petcoke in O₂/CO₂ combustion mediums containing local oxygen enrichments of up to 70%. There was no evidence of particle agglomeration or defluidization in the furnace.
- Because of the high CO₂ content of the flue gas, the furnace operated above 1650°F to ensure calcination of the limestone for sulfur capture. In regions where the temperature was much cooler, there was evidence of recarbonation.
- Because of the high temperature, the sulfur emissions from the bituminous coal were higher than normal for the same limestone feed rate. For pet coke, the optimum temperature for sulfur capture is higher, so the oxygen-fired emissions were very low.
- NO_x emissions were lower with oxygen firing. This is due to the reduction in prompt NO_x, and perhaps some reduction in thermal NO_x and fuel NO_x.
- Carbon monoxide emissions were higher with oxygen firing. This is likely due to the

high CO₂ content of the flue gas, which hinders oxidation of the CO.

- Carbon heat loss in the fly ash was lower than, or comparable to, the levels with air firing. The carbon loss was lower for pet coke than for bituminous coal.
- There was no significant difference in heat transfer to the furnace waterwall test sections between air and oxygen firing. This heat transfer is dominated by solids effects, which do not depend on the gas composition.

These results are largely as expected, based upon earlier test results, and did not identify any major technical barriers to the oxygen-fired CFB concept.

Two of the thirteen cases evaluated in Phase I are: Case-1: Air Fired CFB without CO₂ Capture (Base Case) Conventional air-fired CFB without CO₂ capture; and Case-2: New Compact Oxygen-Fired CFB with CO₂ Capture, Purification, Compression and Liquefaction.

In the present study, Case-2 was updated to incorporate the results from the MTF pilot tests. The significant changes in the design were the addition of the Flash Dryer Absorber (FDA) system and some changes in the low-level heat recovery. Key results are summarized below.

- The plant efficiency of Updated Case-2 was 26.0 percent compared to 25.2 percent for Original Case-2. The heat penalty due to the FDA was more than compensated for by the improvement in low-level heat recovery. The efficiency for Case-1 was 35.5%.
- The capital cost of the Updated Case-2 was 2320 \$/kW(net) compared to 2443 \$/kW(net) for Original Case-2 - about a 5 percent decrease. The cost for Case-1 was 1304 \$/kW(net).
- Levelized cost of electricity (COE) is about 4.5 cents/kWh without CO₂ capture (Case-1). Original Case-2 had a levelized COE of about 8.1 cents/kWh with the addition of the CO₂ capture system (incremental COE of 3.6 cents/kWh) The Updated Case-2 has a levelized COE of about 7.9 cents/kWh (incremental COE of 3.4 cents/kWh). This 2.5% decrease in levelized COE for Updated Case-2 is due to a small decrease in the specific plant cost and a small increase in the net plant output.
- CO₂ mitigation costs are about 40 and 37 \$/ton of CO₂ avoided for Original Case-2 and Updated Case-2, respectively.

To prepare for a commercial demonstration of the oxygen-fired CFB concept, there remain several items to be investigated:

- the operation and performance of the moving bed heat exchanger (MBHE) to avoid recarbonation and also for cost savings compared to the standard fluid bed heat exchanger (FBHE).
- performance of the back-end FDA for sulfur capture under high CO₂ / high moisture flue gas environment using calcined limestone in the fly ash and using fresh commercial lime directly in the FDA.
- determine the effect of recarbonation on fouling in the convective pass.

- assess the impact of oxygen firing on the mercury, trace elements, and VOC emissions.
- update the technical and economic analysis of the oxygen-fired CFB based on these further results.
- develop a proposal-level design for a commercial demonstration of the oxygen-fired CFB concept.

ALSTOM has proposed to the DOE a project to address just these issues with further MTF pilot testing and subsequent update of the technical and economic analyses. This project, entitled Commercialization Development of Oxygen Fired CFB for Greenhouse Gas Control, Cooperative Agreement DE-FC26-04NT42205, has been awarded with a performance period of 1 October 2004 to 30 September 2005. In parallel with this new project, ALSTOM is actively seeking partners for the next step - a commercial demonstration.

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ACRONYMS AND ABBREVIATIONS

acfm	Actual cubic feet per minute
ANSI	American National Standards Institute
ASME	American Society of Mechanical Engineers
ASU	Air Separation Unit
BI	Boiler Island
BOP	Balance of Plant
Btu	British Thermal Unit
CFB	Circulating Fluidized Bed
cfm	Cubic Feet per Minute
CO ₂	Carbon Dioxide
COE	Cost of Electricity
DCS	Distributed Control System
DOE/NETL	Department of Energy/National Energy Technology Laboratory
EOR	Enhanced Oil Recovery
EPC	Engineered, Procured, and Constructed
FBC	Fluidized Bed Combustion
FDA	Flash Drier Absorber
gpm	Gallons per minute
GPS	Gas Processing System
HHV	Higher Heating Value
HP	High Pressure
hr	Hour
ID	Induced Draft
IGCC	Integrated Gasification Combined Cycle
in. H ₂ O	Inches of Water
in. Hga	Inches of Mercury, Absolute
IP	Intermediate Pressure
ISO	International Standards Organization
kWe	Kilowatts electric
kWh	Kilowatt-hour
lbm	Pound mass
LHV	Lower Heating Value
LP	Low Pressure
LT	Low Temperature
MBHE	Moving Bed Heat Exchanger
MCR	Maximum Continuous Rating
MTF	Multi-use Test Facility
MWe	Megawatt electric
N ₂	Nitrogen Gas

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

O&M	Operation & Maintenance
PA	Primary Air
PC	Pulverized Coal
PFD	Process Flow Diagram
PFWH	Parallel Feedwater Heater
PHX	Primary Heat Exchanger
ppm	Parts per million
psia	Pound per square inch, absolute
psig	Pound per square inch, gauge
SA	Secondary Air
TPD	Ton Per Day
TPH	Ton Per Hour

EXECUTIVE SUMMARY

Background

Because fossil fuel fired power plants are among the largest and most concentrated producers of CO₂ emissions, recovery and sequestration of CO₂ from the flue gas of such plants has been identified as one of the primary means for reducing anthropogenic CO₂ emissions. In this Phase II study, ALSTOM Power Inc. (ALSTOM) has investigated one promising near-term coal fired power plant configuration designed to capture CO₂ from effluent gas streams for sequestration.

Burning fossil fuels in mixtures of oxygen and recirculated flue gas (made principally of CO₂) essentially eliminates the presence of atmospheric nitrogen in the flue gas. The resulting flue gas is comprised primarily of CO₂, along with some moisture, nitrogen, oxygen, and trace gases like SO₂ and NO_x. Oxygen firing in utility scale Pulverized Coal (PC) fired boilers has been shown to be a more economical method for CO₂ capture than amine scrubbing (Bozzuto, et al., 2001). Additionally, oxygen firing in Circulating Fluid Bed Boilers (CFB's) can be more economical than in PC or Stoker firing, because recirculated gas flow can be reduced significantly. Oxygen-fired PC and Stoker units require large quantities of recirculated flue gas to maintain acceptable furnace temperatures. Oxygen-fired CFB units, on the other hand, can accomplish this by additional cooling of recirculated solids. The reduced recirculated gas flow with CFB plants results in significant Boiler Island cost savings resulting from reduced component sizes.

Objective

The overall objective of the Phase II workscope, which is the subject of this report, is to generate a refined technical and economic evaluation of the Oxygen fired CFB case (Case-2 from Phase I) utilizing the information learned from pilot-scale testing of this concept. The objective of the pilot-scale testing was to generate detailed technical data needed to establish advanced CFB design requirements and performance when firing coals and delayed petroleum coke in O₂/CO₂ mixtures. Firing rates in the pilot test facility ranged from 2.2 to 7.9 MM-Btu/hr. Pilot-scale testing was performed at ALSTOM's Multi-use Test Facility (MTF), located in Windsor, Connecticut.

Results Summary

The MTF operated successfully on coal and petroleum coke with O₂/CO₂ mixtures with up to 70% O₂ in the firing zone. No technical barriers to continued development were found. Specific conclusions from the MTF pilot-scale testing are summarized here.

- Furnace Operability/Agglomeration: For the most part, the furnace operated well when firing with the O₂/CO₂ mixture rather than with air. There was no evidence of particle agglomeration or defluidization in the furnace at local oxygen enrichments of up to 70%.
- Recarbonation: The furnace was kept above the calcination temperature for these tests. Solids samples from the furnace showed only 1-2 % residual CaCO₃. Fly ash samples had cooled to below the calcination temperature; they had higher levels of CaCO₃, as expected.

The MTF sealpot operated below the calcination temperature; recarbonation occurred,

upsetting the sealpot operation. This would not occur in a commercial unit, where the sealpot temperature is as hot, or hotter, than the furnace.

There was a pluggage in the cyclone due to recarbonation when firing pet coke.

- Gaseous Emissions: The gaseous emissions compared to air firing were generally as expected. The sulfur emissions were higher with the bituminous coal, since the furnace was above the optimum temperature. For pet coke, the optimum temperature is higher, so the SO₂ emissions were very low with oxygen-firing.

The carbon monoxide emissions were higher with oxygen firing, as expected based on earlier bench-scale tests. This is likely due to the high CO₂ content of the flue gas, which hinders oxidation of the CO.

Earlier tests have shown that the NO_x emissions are lower with oxygen firing; this was also the case in these pilot tests. It is not clear to what extent this is due to reductions in thermal, prompt, and/or fuel NO_x.

Nitrous oxide emissions were lower than expected from air firing at comparable conditions.

- Unburned Carbon Heat Loss: With oxygen firing, the carbon heat loss in the fly ash was lower than, or comparable to, the levels with air firing. The carbon loss was lower for pet coke than for bituminous coal.
- Heat Transfer: There was no significant difference in heat transfer to the furnace waterwall test sections between air and oxygen firing. This heat transfer is dominated by solids effects, which do not depend on the gas composition.

At the conclusion of the pilot scale testing and data analysis tasks, updated performance calculations, component designs, costs, and economics were calculated for the Case-2 design concept (i.e., New Compact O₂-Fired CFB with CO₂ Capture, Purification, Compression, and Liquefaction). Comparisons were also made to a business as usual scenario (Case-1: Air fired CFB utilizing the same steam cycle but without CO₂ capture).

A brief performance summary for the Updated Case-2 plant reveals the following information. The Updated Case-2 plant produces a net plant output of about 138 MWe. This represents a reduction of about 28 percent as compared to Case-1 (Air fired CFB utilizing the same steam cycle but without CO₂ capture). The net plant heat rate and thermal efficiency are calculated to be 13,152 Btu/kWh and 25.95 percent respectively (HHV basis) for this case. This represents an energy penalty of about 27 percent as compared to Case-1. The large output and efficiency reduction is primarily the result of large power consumption requirements for the cryogenic Air Separation Unit and the Gas Processing System. The Gas Processing System for Updated Case-2 captures about 93 percent of the CO₂ produced. Carbon dioxide emissions are subsequently reduced to about 0.17 lbm/kWh on a normalized basis. This represents about 1.82 lbm/kWh of avoided CO₂ emissions. (Case-1, without CO₂ capture, emits about 2.0 lbm/kWh of CO₂).

Investment costs are shown in the following figure. The graph on the left shows the impact of the relatively costly and power consuming Air Separation Unit and Gas Processing System with the specific cost of the power plant (\$/kW-net) increasing by almost 80 percent as compared to the air fired Case-1 without CO₂ capture. The use of \$/kWgross units for the graph on the right clearly shows the savings incurred for the

traditional power plant equipment as a result of oxygen firing. This savings totals to about 20 percent and is primarily a result of cost reductions in the Boiler Island for equipment where cost is strongly dependent on gas flow (combustor, cyclones, backpass heat exchangers, air heater, fans, ductwork, particulate removal system, etc.).

The economic study results are summarized by comparing the levelized COE, incremental COE, and CO₂ mitigation cost results for these CO₂ capture technologies as shown in the following figure respectively. The levelized COE value is increased from about 4.5 cents/kWh without CO₂ capture (Case-1) to about 8.1 cents/kWh with the addition of the CO₂ capture system for Original Case-2 and 7.9 cents/kWh for Updated Case-2. The update of Case-2 provided a relatively minimal impact on levelized COE (about -2%). The slight improvement in COE for Updated Case-2 is primarily the result of a small decrease in the specific plant cost (about 2.5% - \$/kW basis) and a small increase in the net plant output (about 2.9%). The incremental COE as compared to the air-fired CFB reference plant without CO₂ capture (Case 1) is about 3.6 cents / kWh for Original Case-2 and 3.4 cents/kWh for Updated Case-2. Similarly, the CO₂ mitigation costs are about 40 and 37 \$/ton of CO₂ avoided respectively.

Recommendations

To prepare for a commercial demonstration of the oxygen-fired CFB concept, there remain several items to be investigated:

- the operation and performance of the moving bed heat exchanger (MBHE) to avoid recarbonation and also for cost savings compared to the standard fluid bed heat exchanger (FBHE).
- performance of the back-end Flash Dryer Absorber (FDA) for sulfur capture under high CO₂ / high moisture flue gas environment using calcined limestone in the fly ash and using fresh commercial lime directly in the FDA.
- determine the effect of recarbonation on fouling in the convective pass.
- assess the impact of oxygen firing on the mercury, trace elements, and VOC emissions.
- update the technical and economic analysis of the oxygen-fired CFB based on these further results.
- develop a proposal-level design for a commercial demonstration of the oxygen-fired CFB concept.

ALSTOM has proposed to the DOE a project to address just these issues with further MTF pilot testing (at the full 9.9 MMBtu/hr firing rate) and subsequent update of the technical and economic analyses. This project has been selected and awarded by the DOE - Commercialization Development of Oxygen Fired CFB for Greenhouse Gas Control, Cooperative Agreement DE-FC26-04NT42205. The performance period for this project is from 1 October 2004 to 30 September 2005. In parallel with this new project, ALSTOM is actively seeking partners for the next step - a commercial demonstration.

1 INTRODUCTION

The greenhouse effect is created by the presence of a number of gases in the atmosphere, with CO₂ accounting for about 50 percent of the greenhouse phenomenon. Large quantities of CO₂ are produced from fossil fuel combustion. Previous studies (e. g., Bozzuto, et al., 2001) have shown that CO₂ capture from existing coal fired plants utilizing commercial amine-based flue gas scrubbing systems would reduce plant output and efficiency by up to one-third and increase cost of electricity by almost 5 cents/kWh.

An alternative method for CO₂ capture is to burn fossil fuels in a mixture of oxygen and recycled flue gas. This concept eliminates the presence of almost all atmospheric nitrogen in the flue gas, thereby resulting in a flue gas that is composed primarily of CO₂, along with small quantities of moisture, oxygen, nitrogen, and trace gases like SO₂ and NO_x. The combination of recycled flue gas/oxygen mixtures in concert with combustion in a circulating fluidized bed (CFB) boiler offers unique advantages compared to alternative methods of firing fossil fuels with oxygen. Unlike pulverized coal (PC) combustion or stoker firing, fluidized bed combustion has the advantage of controlling combustion chamber temperatures by virtue of modulating the recycle rate of cooled solids. This unique feature of a fluidized bed combustor means that much higher percentages of oxygen can be used in the recycled flue gas/oxygen mixtures than would be possible in alternate firing applications.

Though the primary motivation for using oxygen is to facilitate CO₂ capture, newly constructed CFB combustors will be able to capitalize on the use of high oxygen content firing. Specifically, and importantly, the use of higher oxygen contents will improve overall system thermal efficiency, and allow the design and construction of more compact, relatively less expensive CFB boilers.

To investigate the feasibility of various carbon capture technologies, including the oxygen-fired CFB concept, the DOE sponsored a 2-phase program. ALSTOM received a Cooperative Agreement from the US Department of Energy National Energy Technology Laboratory (DOE) in 2001 to carry out a project entitled “Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers.” This two-phased project is in effect from August 31, 2001, to October 27, 2004. (U.S. DOE NETL Cooperative Agreement No. DE-FC26-01NT41146). ALSTOM provided a 20% cost share to this project.

The technical and economic analyses and bench scale test results from Phase I (Marion, et al., 2003) indicate that the proposed oxygen-firing technology in circulating fluidized boilers could be developed and deployed economically in the near future in an enhanced oil recovery (EOR) application.

To further develop this oxygen-fired CFB technology in Phase II, ALSTOM conducted a pilot scale test program followed by an update of the technical and economic analyses from Phase I. The results of this Phase II work are presented in this report.

2 OVERALL PROJECT STRUCTURE:

The Work Breakdown Structure of each phase of the two-phase effort is shown in Figure 2-1.

Phase I:

The Phase I workscope is comprised of three tasks, as follows:

- Task 1: Conceptual Performance and Economic Analyses of Thirteen Study Cases
- Task 2: Bench-Scale Fluidized Bed Combustion (FBC) Testing
- Task 3: Phase I Topical Report.

Figure 2-1: Work Breakdown Structure

The major work within Phase I was Task 1. The key goals of Task 1 were to evaluate the impacts on the plant output, efficiency, and CO₂ emissions, resulting from the addition of various CO₂ capture systems to an array of CFB combustion based, IGCC based, and advanced Chemical Looping based power plants. Cost estimates were developed for these power plants and the impact of CO₂ capture on the levelized cost of electricity (COE) and on the mitigation cost for CO₂ (\$/ton of CO₂ avoided) were also evaluated.

Task 2, Bench Scale FBC testing, supported the Task 1 case studies. The objective of Task 2 was to derive pertinent combustion performance and bed dynamic information under highly controlled operating conditions in a 4-inch bubbling fluidized bed test

facility. Results from oxy-fuel firing of three fuels, two coals and one delayed petroleum coke, were compared to those results obtained similarly from air firing.

Phase II:

The overall objective of the Phase II workscope, which is the subject of this report, is to generate a refined technical and economic evaluation of the Oxygen fired CFB case (Case-2 from Phase I) utilizing the benefits derived from pilot-scale testing of this concept. The objective of the Task-4 pilot-scale testing was to generate detailed technical data needed to establish advanced CFB design requirements and performance when firing coals and delayed petroleum coke in O₂/CO₂ mixtures. Firing rates in the pilot test facility ranged from 2.2 to 7.9 MM-Btu/hr. Pilot-scale testing was performed at ALSTOM's Multi-use Test Facility (MTF), located in Windsor, Connecticut.

At the conclusion of Task-4, updated performance calculations, component designs, costs, and economics were calculated in Task-5 for the Case-2 design concept (*i.e.*, New Compact O₂-Fired CFB with CO₂ Capture, Purification, Compression, and Liquefaction).

2.1 Phase I Project Summary

Phase I entails a comprehensive study evaluating the technical feasibility and economics of alternate CO₂ capture technologies applied to Greenfield US coal-fired electric generation power plants. Thirteen cases were evaluated, representing various levels of technology development. The first seven cases represent coal combustion cases in CFB type equipment. The next four cases represent Integrated Gasification Combined Cycle (IGCC) systems. The last two cases represent advanced Chemical Looping systems, which were completely paid for by ALSTOM. Details of this work have been reported by Marion, *et al*, (2003).

Combustion Cases:

- Case-1: Air Fired Circulating Fluidized Bed (CFB) without CO₂ Capture (Base Case for Comparison to Cases 2-7)
- Case-2: Oxygen Fired CFB with CO₂ Capture
- Case-3: Oxygen Fired CFB with CO₂ Capture (sequestration only option)
- Case-4: Oxygen Fired Circulating Moving Bed (CMB) with CO₂ Capture (advanced boiler concept)
- Case-5: Air Fired CMB with CO₂ Capture utilizing Regenerative Carbonate Process
- Case-6: Oxygen Fired CMB with Oxygen Transport Membrane (OTM) and CO₂ Capture
- Case-7: Indirect Combustion of Coal via Chemical Looping and CO₂ Capture

IGCC Cases:

- Case-8: Built and Operating Present Day IGCC without CO₂ Capture (Base Case for Comparison with Case-9)
- Case-9: Built and Operating Present Day IGCC with shift reaction and CO₂ Capture
- Case-10: Commercially Offered Future IGCC without CO₂ Capture (Base Case for Comparison with Case-11)

- Case-11: Commercially Offered Future IGCC with shift reaction and CO₂ Capture

Advanced Chemical Looping Cases:

- Case-12: Indirect Gasification of Coal via Chemical Looping (Base Case for comparison to Case-13)
- Case-13: Indirect Gasification of Coal and CO₂ Capture via Chemical Looping

The results from the Phase I analysis led to the conclusion that further development work on the Oxygen-Fired CFB (Case-2) was justified. This recommendation was made to the DOE for the following reasons:

- The Oxygen-Fired CFB is the most near-term solution (~5-year horizon), as it uses enabling technologies, which are available commercially today, for example:
 - ♦ Oxygen production by cryogenic air separation unit
 - ♦ CO₂ capture, purification, compression and liquefaction
- Significant boiler cost savings for O₂ fired CFB as compared to O₂ fired PC or Stoker.
- Preliminary economic analyses look viable for commercial EOR applications.
 - ♦ CO₂ product credit for oil field stimulation
 - ♦ N₂ by-product (from ASU) credit for oil field pressurization
- CO₂ in the flue gas is highly concentrated (~90 percent vs.~15 percent), thus making the processing of this stream to achieve the required CO₂ purity for EOR applications less expensive.
- Oxy-fuel firing in CFB's offers added emissions benefits over air firing, as had been preliminarily shown through combustion testing in an ALSTOM bench-scale fluidized bed combustor (Marion et al., 2003):
 - ♦ Typically low NO_x emissions in combustion-staged air-fired CFB's are further reduced due primarily to elimination of thermal NO_x.
 - ♦ SO₂ emissions are reduced by up to 90 percent. Sorbent utilization should not be negatively impacted if the bed temperature is maintained the same as in air firing. Furthermore, ALSTOM has a commercial product called "Flash Drier Absorbent (FDA)," which has been successfully demonstrated in the multi-use test furnace (MTF) to reduce SO₂ emissions by as much as 99 percent.
 - ♦ Unburned carbon (UBC) loss should not be negatively impacted.
 - ♦ A simplified gas processing system (GPS) design (drying and compression only) allows a "zero gaseous emission plant" although product gas purity would not be acceptable for EOR applications.

The DOE concurred with ALSTOM's recommendation of developing the O₂ fired CFB technology for capturing CO₂ and, hence, authorized the implementation of Phase-II of the current project.

2.2 Phase II Objective and Workscope

The overall objective of the Phase II workscope is to generate a refined technical and economic evaluation of Case-2 of Phase I (i.e., Case 2: New Compact O₂-Fired CFB with CO₂ Capture, Purification, Compression, and Liquefaction), utilizing the information

derived from pilot-scale testing of the same concept. The Phase II workscope has been developed based upon the findings from Phase I and will specifically address both retrofit (moderate O₂ enrichment/ high flue gas recirculation) and Greenfield (Case 2) applications (high oxygen enrichment/low flue gas recirculation). The objective of the pilot-scale testing was to generate detailed technical data needed to establish advanced CFB design requirements and performance when firing coal and delayed petroleum coke in O₂/CO₂ mixtures. Firing rates in the pilot test ranged from 2.2 to 7.9 MM-Btu/hr. Pilot-scale testing was performed at ALSTOM's Multi-use Test Facility (MTF), located in Windsor, Connecticut. Outputs from this testing address key technical parameters including:

- Flue Gas Quality
- Bed Dynamics
- Heat Transfer to the Waterwalls
- Flue Gas Desulfurization
- NO_x Emissions Reduction
- Other Pollutants' Emissions (N₂O and CO)
- Bed and Ash Characteristics (e.g., Potential Bed Agglomeration)

Results from the test data analysis will be available for the design of systems to retrofit existing CFB units for oxygen firing and for the design of new oxygen-fired CFB boilers. Test data analysis results were also used in Task-5 of Phase II. The results of this task are discussed in Sections 5, 6, and 7 of this report where the revised design, performance calculations, costs and economics are updated for original Case 2 of Phase I (New Compact O₂-Fired CFB with CO₂ Capture, Purification, Compression, and Liquefaction).

3 CHARACTERIZATION OF FUEL AND LIMESTONE SAMPLES

Standard ASTM protocols and special in-house methodologies were used to characterize the three fuel and two limestone test samples. The purpose of these characterizations was to determine the chemical nature and reactivity characteristics of these samples, and to derive pertinent information for use in combustion calculations associated with Multi-use Test Facility (MTF) testing (Section 4). The results obtained from these characterizations are given in the next two sections.

3.1 Characterization of Test Fuels

3.1.1 Identification of Samples

Two coals and one petroleum coke were selected for this study. One of two coals is a medium volatile bituminous (mvb) coal obtained from Tri-Star Mining, Inc. This coal (from here on referred to as Tri-Star coal) is a 50/50 (weight %) blend of Big Vein and Morantown coal seams from Garrett County in Maryland. The second coal is a high volatile A bituminous (hvAb) coal from the Pittsburgh #8 coal seam. The shot petroleum coke was acquired from Conoco Phillips's VENCO plant in Moundsville, WV.

3.1.2 Analyses of Samples

Each of the three test fuels was subjected to proximate and ultimate analyses and higher heating (HHV) value determinations. The fuel analyses are given in Table 3-1. The Tri-Star mvb coal has nitrogen and sulfur contents of 2.0% and 1.7% on a dry-ash-free (DAF) basis, respectively. These values result in corresponding nitrogen and sulfur (expressed as SO₂) loadings of 1.3 and 2.3 lb/MMBtu. The Pittsburgh # 8 hvAb coal has nitrogen and sulfur contents of 1.4 and 4.0%, respectively; these values are equivalent to nitrogen and SO₂ loadings of 0.9 and 5.5 lb/MMBtu. The shot petroleum coke (petcoke) has a nitrogen content that is similar to those of the test coals (i.e., 1.6% DAF), and, as expected, a very high sulfur content (5.5% DAF). These values are equivalent to nitrogen and SO₂ loadings of 1.1 and 7.2 lb/MMBtu, respectively. These data foreshadow that if these fuels were tested under similar conditions, they would have: (1) similar NO_x emission potentials; and (2) variable SO₂ emissions potentials, namely, the pet coke, mvb coal, and hvAb coal would have the highest, intermediate, and lowest potentials, respectively.

Also, the coal ash samples were subjected to composition and fusibility temperature measurements (Table 3-2). The ash compositions of both Tri-Star and Pittsburgh #8 coals are typical of those from Eastern US bituminous coals. That is, their iron contents (as Fe₂O₃) are greater than the alkali and alkaline earth metal contents (i.e., CaO+MgO+Na₂O+K₂O). The combination of high sulfur content in the Pittsburgh #8 coal and ash composition leads to low ash fusibility temperatures, indicating, as is well known, high slagging potential for this coal. As expected, the petcoke has very low ash (1.5-%) content.

Table 3-1: Fuel Analyses

QUANTITY	Tri-Stat (mvb) Coal		Pittsburgh #8 (hvAb) Coal		Shot Petroleum Coke	
	As Received	Dry-Ash-Free	As Received	Dry-Ash-Free	As Received	Dry-Ash-Free
Prox. & Ult. Analyses, Wt. %						
Moisture	4.5	---	2.4	---	0.4	---
Volatile Matter	16.7	20.8	34.9	40.3	9.1	9.3
Fixed Carbon	63.4	79.2	51.8	59.7	89	90.7
Ash	15.4	---	10.9	---	1.5	---
Hydrogen	3.8	4.7	4.9	5.7	3.7	3.8
Carbon	70.1	87.5	70.9	81.8	87.1	88.8
Sulfur	1.4	1.7	3.5	4.0	5.4	5.5
Nitrogen	1.6	2.0	1.2	1.4	1.6	1.6
Oxygen	3.2	4.0	6.2	7.2	0.3	0.3
Chlorine, Wt. % Dry	0.005		0.051		0.003	
Fluorine, Wt. % Dry	0.013		0.049		---	
Mercury, dppmw	0.05	---	0.04		0.05	
HHV, Btu/lb	12107	15115	12804	15985	14949	18663
FC/VM (Fuel Value)	3.8		1.5		9.8	
Ash _{Loading} , lb/MMBtu	12.7		8.5		1.0	
Nitrogen _{Loading} , lb/MMBtu	1.3		0.9		1.1	
Sulfur (as SO ₂) _{Loading} , lb/MMBtu	2.3		5.5		7.2	

Table 3-2 depicts the results on the slagging and fouling potentials of the Tri-Star and Pittsburgh #8 coals, based on an algorithm that uses the formulation of Stultz and Kitto (1992). This algorithm uses the ash properties plus, where applicable, sulfur content of the coal under study. The results from this algorithm show that the ash-type of these two Eastern US bituminous coals is indeed “Bituminous-type. The algorithm also indicates that:

- The Tri-Star coal ash has medium slagging and low fouling potentials; and
- The Pittsburgh #8 coal ash has high slagging and medium fouling potentials.

These three fuels fall well within the range of various fuels which have been burned in ALSTOM-designed CFBs (e.g., Tanca, 1994; and Marchetti, et al., 2003).

Table 3-2: Slagging and Fouling Characteristics of the Two Coal Samples

QUANTITY	Tri-Star mvb Coal	Pittsburgh #8 hvAb Coal
Fusibility Temperatures (°F)		
I.T.	2445	2155
S.T.	2515	2265
H.T.	2604	2318
F.T.	2650	2372
Composition, Wt.% (dry)		
SiO ₂	55.2	42.5
Al ₂ O ₃	24.1	19.8
Fe ₂ O ₃	9	22.8
CaO	2.3	2.4
MgO	1	0.5
Na ₂ O	0.4	0.8
K ₂ O	2.3	1.5
TiO ₂	1.2	0.9
P ₂ O ₅	0.5	0.3
SO ₃	2.3	2.5
MnO	0.1	3.3
Total	98.4	97.3
Coal Sulfur Content, Wt.% (dry)	1.47	3.59
Ash Type	Eastern Bit.	Eastern Bit.
Viscosity based Slagging Index Definition: Valid for both Bituminous ashes and Lignitic Ashes		
$R_{vs} = (T(250 \text{ oxid}) - T(10,000 \text{ red})) / (97.5 * f_s)$		
f_s :	4.87	1.74
INDEX: R_{vs}	0.74	1.52
$R_{vs} < 0.5$ = low		
$0.5 < R_{vs} < 1.0$ = medium	MEDIUM	
$1.0 < R_{vs} < 2.0$ = high		HIGH
$2.0 < R_{vs}$ = Severe		
Fouling Index Definition		
Bituminous Ash: $R_F = B/A * Na_2O$	TRUE	TRUE
Lignitic Ash, Na ₂ O, when CaO+MgO+Fe ₂ O ₃ >20% (w/w)		
Lignitic Ash, Na ₂ O, when CaO+MgO+Fe ₂ O ₃ <20% (w/w)		
Fouling Index & Potential		
INDEX: Bituminous Ash: $R_F = B/A * Na_2O$	0.075	0.354
$R_F < 0.2$ = low	LOW	
$0.2 < R_F < 0.5$ = medium		MEDIUM
$0.5 < R_F < 1.0$ = high		
$R_F > 1.0$ = severe		
INDEX: (Lignitic Ash, Na ₂ O, when CaO+MgO+Fe ₂ O ₃ >20% (w/w))		
Na ₂ O < 3 = low to medium		
3 < Na ₂ O < 6 = high		
Na ₂ O > 6 = severe		
INDEX: (Lignitic Ash, Na ₂ O, when CaO+MgO+Fe ₂ O ₃ <20% (w/w))		
Na ₂ O < 1.2 = low to medium		
1.2 < Na ₂ O < 3 = high		
Na ₂ O > 3 = severe		
Ash Classification: Fe ₂ O ₃ > CaO+MgO ==> Bituminous ash; Fe ₂ O ₃ < CaO+MgO ==> Lignitic ash		

3.1.3 Limestone Analyses

The chemical analyses of the two limestones evaluated are shown in Table 3-3. Each of these sorbents have been successfully used in ALSTOM-designed CFBs. The sorbents are non-dolimitic - that is they have low magnesium content. The Aragonite has a higher inert content than the Chemstone (5.9 vs. 2%).

Table 3-3: Limestone Analyses

QUANTITY	Chemstone Limestone	Aragonite
Weight % Dry		
Total Ca	38.86	37.40
Total Mg	0.29	0.20
CO ₂	43.55	42.32
CaCO ₃	97.04	93.40
MgCO ₃	1.01	0.69
Inerts (Diff.)	1.95	5.91

3.2 TGA Reactivity Characteristics of Fuel and Limestone Samples

3.2.1 TGA Description and Testing Procedure

ALSTOM uses a Perkin-Elmer Thermo-Gravimetric Analysis (TGA) Model 7 (Figure 3-1) to obtain specific reactivity parameters of fuels and limestones, among other things. The TGA is also used to derive the “micro-proximate” analyses of coals and coal chars, based on a test protocol developed in-house. Testing for the current program was conducted as shown below.

Fuel reactivity determinations were conducted as follows. An approximately 5-mg sample, normally sized to 200x400 mesh, was placed in the TGA sample pan. The sample amount was such that it was distributed over the bottom of the pan in almost a monolayer, thus minimizing the O₂ mass transfer control phenomenon during combustion. Equal amounts of nitrogen (which serves to protect the balance) and air or a combination of O₂ and CO₂ were allowed to pass through the reaction tube containing the coal sample. The temperature control mechanism was actuated, such that the heating rate was maintained at 10 °C/min from room temperature (~ 25 °C) to the completion of combustion. Both weight loss and rate of weight loss were monitored and recorded throughout the combustion process. The information from this testing was used, as shown in Section 3.2.2, to derive the weight loss and rate of weight loss curves. The weight loss curves, expressed on a dry, ash-free fuel basis, depict the fuel’s combustion efficiency (or burn-off) characteristic. The rates of weight loss curves, which can be plotted as a function of temperature, depict “pseudo” surface reaction kinetic parameters of a given fuel. The fuel's reactivity can be compared with the reactivity of other fuels in ALSTOM’s database, including those used commercially.

Figure 3-1: Thermo-Gravimetric Analysis (TGA) Apparatus

Each of the three subject fuels was burned in air along with other reference fuels, and the subject three fuels were also burned in an O_2/CO_2 mixture, as shown in Table 3-4. Essentially, 50 cc/min of N_2 gas were admitted, in all circumstances, into the balance compartment to protect it from over-heating. Burning the fuels in air consisted of admitting in 50 cc/min of air. Burning the fuels in the O_2/CO_2 mixture consisted of admitting 10.5 and 25.3 cc/min of O_2 and CO_2 , respectively. These tests were tailored such that the total mass flow rate of the gas was the same (i.e., 0.1162 gm/min) for both test scenarios (i.e., in air and in O_2/CO_2 mixture). The rationale for a “mass conservation” design was to simulate a retrofit scenario of an existing air-fired furnace to oxy-fuel firing. It had been shown by ALSTOM (Bozzuto, et al., 2001; Marion, et al. 2003) that such a case would require only minimal modification to a plant’s boiler island.

Table 3-4: TGA Gas Mixtures

Combustion Medium	Balance Cooling N ₂		Combustion Air						CO ₂		Grand Total		Combustion Medium Composition, Vol. %		
			Total Air		O ₂		N ₂								
	(cc/min)	(gm/min)	(cc/min)	(gm/min)	(cc/min)	(gm/min)	(cc/min)	(gm/min)	(cc/min)	(gm/min)	(cc/min)	(gm/min)	N ₂	CO ₂	O ₂
Air	50	0.0573	50	0.0592	10.5	0.01373	39.5	0.045524	0	0	100	0.1162	89.5	0	10.5
O ₂ /CO ₂ Mixture	50	0.0573	0	0	10.5	0.01373	0	0	25.2	0.04524	85.7	0.1162	58.4	29.4	12.3

In the case of limestone sulfation reactivity tests, a number of measurements are normally done as a function of limestone particle size and temperature. Normally, three particle size fractions [60x100-mesh (150-250 μm), 25x35-mesh (500-710 μm), and 16x20-mesh (850-1180 μm)] and three temperatures (850, 900, and 950 $^{\circ}C$) are used in this evaluation. Testing entails the following steps. The sample is first brought to the test temperature (e.g., 850 $^{\circ}C$) in a 15% $CO_2/85\% N_2$ mixture to effect calcination. After the weight has stabilized, the reacting atmosphere is switched to a mixture of 3000 vppm SO_2 , 5% O_2 , 15% CO_2 with the remainder as N_2 . The limestone sample is then sulfated

at a constant temperature in this atmosphere for up to 180 minutes. The resulting weight gain is attributed to the reaction of CaO with SO₂ and O₂ to form CaSO₄. The weight gain is then used to calculate the percent calcium utilization and plotted as a function of reaction time, as shown in Section 3.2.2. Calcium utilization results from a given limestone are compared with those from the limestone samples in ALSTOM's database. Also, the reactivity kinetic parameters from this evaluation can be used in conjunction with ALSTOM's models for predicting limestone performance in CFBs.

3.2.2 TGA Fuel Combustion Results

Burn-off curves of the three subject fuels are shown in Figure 3-2 along with those of three reference fuels covering a wide coal-rank spectrum. The reference coals consisted of a subbituminous coal from Powder River Basin seam, a high volatile C bituminous (hvCb) coal from Illinois #6 seam, and an anthracite (an) from Pennsylvania. Results indicate that these fuels burn to completion in the following order: Subbituminous coal, Illinois #6 hvCb coal, Pittsburgh #8 hvAb coal, Petcoke, Tri-Star mvb coal, and Pennsylvania anthracite. It is noteworthy that while the medium volatile bituminous coal's combustion takes off earlier than the petcoke, due to its relatively higher volatile matter, it takes the longer than the petcoke to burn to completion.

Figure 3-3 shows data from the same TGA data above plotted as the rate of weight loss as a function of temperature. Plotting the data in this manner shows the position of the combustion intensity peaks, as evidenced by peak rates of weight loss. The relative positions of fuel reactivity are unchanged from the previous figure, but the shapes of the curves denote something about the nature of the combustibles. Generally, the position of the weight loss peak varies inversely with the volatile matter content in the fuel. This is supported by the fact that the volatile matter contents of the subbituminous, hvCb, hvAb, mvb, and anthracite are 47, 44, 40, 21, and 4% (dry-ash-free), respectively. The corresponding value for the petcoke is 9%.

Figure 3-4, and Figure 3-5 show the weight loss (i.e., burn-off), and rate of weight loss curves for the three subject fuels in air and O₂/CO₂ mixture. The burn-off curves show that combustion in O₂/CO₂ mixture takes off earlier than it does in air, the completion times of combustion are virtually the same. Consistent with the burn-off curves, the rate of weight loss curves show that the combustion intensity peaks in O₂/CO₂ mixture occur slightly ahead of those in air. Nevertheless, the completion time of combustion is virtually the same in both cases. Hence, the basic message from these plots is that oxy-fuel firing does not negatively impact the overall combustion efficiency of a fuel compared to air firing. These results are consistent with those obtained previously by ALSTOM (Bozzuto, et al., 2001; and Marion, et al., 2003).

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Figure 3-2: Burn-Off Curves of Various Fuels as a Function of Temperature

Figure 3-3: Rate of Weight Loss of Various Fuels in Air as a Function of Temperature

Figure 3-4: Weight Loss in Air and O₂/CO₂ as a Function of Temperature

Figure 3-5: Rate of Weight Loss in Air and O₂/CO₂ as a Function of Temperature

3.2.3 TGA Limestone Utilization Results

Figure 3-6 illustrates results on calcium utilization in the TGA for the 25x35-mesh size fractions of the Aragonite and Chemstone limestones. The ultimate calcium utilization results obtained from the three size fractions of each limestone sample, at three temperatures, are summarized in Table 3-5. The Chemstone limestone is much less reactive than the Aragonite, resulting in correspondingly lower calcium utilization. This is consistent with commercial experience. The utilization in the TGA increases for each sorbent with higher temperature and with smaller particle size. The situation is not so straightforward in the CFB combustor. Fuel factors are more important to determine the optimum temperature for sulfur capture. And while finer particles are generally more reactive, if they are too fine, they quickly escape from the furnace.

Figure 3-6: Sulfation Curves of 25x35-Mesh Limestone Samples at 850 °C

Table 3-5: Summary of Limestone Utilization Results

Limestone	Particle Size (Mesh/ μm)	Utilization at Sulfation Temperature		
		850 °C	900 °C	950 °C
Chemstone	(60x100/150-250 μm)	19.5	21.5	26.9
	(25x35/500-710 μm)	12.8	13.2	11.3
	(16x20/850-1180 μm)	7.7	11.4	7.4
Aragonite	(60x100/150-250 μm)	66.5	51.1	70.6
	(25x35/500-710 μm)	41.1	43.7	46.8
	(16x20/850-1180 μm)	42.5	39.7	36.3

3.2.4 Conclusions from Fuel and Limestone Characterization and TGA Combustion Results

A number of conclusions can be drawn from the fuel and limestone characterization and TGA combustion results, namely:

- All three test fuels have similar NO_x emission potentials. With respect to SO₂ emissions potentials, the order is: Tri-Star mvb Coal < Pittsburgh #8 hvAb Coal < Petcoke
- Oxy-combustion of any of these fuels should not negatively impact their combustion efficiencies (or unburned carbon losses)
- Pittsburgh #8 hvAb coal has higher slagging and fouling potentials than Tri-Star mvb coal. Both of these coals are currently successfully burned in ALSTOM-designed commercial CFB or pulverized coal boilers.
- Aragonite is substantially more reactive than Chemstone for sulfur capture. Potential utilization of each sorbent increases with smaller particle size.

4 PILOT-SCALE CFB COMBUSTION TESTING

4.1 Background and Objectives

A major task of the Phase II program was to conduct a pilot plant test in ALSTOM's 9.9 MMBtu/hr pilot plant. The objective of the pilot testing was to simulate an oxygen-fired commercial plant and demonstrate successful operation. The testing also generated data on the following aspects of oxygen-fired combustion. This information will be used for the design of commercial sized units.

- Agglomeration Potential
- Heat Transfer
- Recarbonation
- Sulfur Capture
- Other Emissions
- Combustion Efficiency

4.1.1 Agglomeration

A circulating fluidized bed combustor typically operates with a furnace temperature of between 1500 and 1650°F - depending on the type of fuel. This temperature, measured by a thermocouple, is the average temperature of the coal, ash, and gas at any location in the furnace. The actual temperature at the surface of the burning coal particles is set by a heat balance: that is, the **rate of heat release** by combustion of carbon and carbon monoxide at the particle surface and the **rate of heat transfer** between the surface and the bulk solids and gas by conduction, convection, and radiation (see for example, Field, et al, 1967, p. 186). Depending on the fuel reactivity and the particle size, the surface temperature may be lower than the bulk temperature (for quite large particles) or several hundreds of degrees hotter than the bulk temperature (hottest for the smallest particles). It is possible for the surface of some of the fuel particles to become hot enough for the fuel ash to become sticky or to melt. Since these small fuel particles are vigorously fluidized in the midst of a much larger quantity of inert ash and sorbent, there is usually not a problem with large scale agglomeration of ash, which would lead to defluidization.

With oxygen firing, the furnace will likely run above 1650°F to avoid limestone recarbonation (see Section 4.1.3 below). The higher temperature increases the fuel reaction rate, so the surface temperature might increase even further. The combustor will operate with a normal excess of oxygen - 3 to 4% O₂ in the flue gas. Far from the oxygen injection points, the oxygen content of the gas should be similar to the air-fired case. However, the incoming oxidant is highly enriched with up to 70% oxygen. Near the oxygen injection locations, the reaction rate can be greatly increased by a high oxygen content, with much higher surface temperatures.

In Phase I of this program, ALSTOM burned three fuels (Tri-Star mvb coal, Illinois #6 coal, and a delayed petcoke) in their four-inch bubbling fluidized bed combustor in air and various O₂/CO₂ mixtures (Marion, et al., 2003). As the oxygen content increased above 30%, the bed changed from an orange-yellow color to bright white. At 70% O₂ with Tri-Star coal, the bed so bright it was difficult to look at. Even at 50% oxygen, there

was agglomeration when fluidized at the normal fluidization velocity of 1.7 ft/sec. It did not agglomerate at an increased velocity of 2.8 ft/sec. The Illinois #6 hvCb coal, which is known to be a high slagging coal, was successfully burned at 2.7 ft/sec in up to 50% oxygen. The petroleum coke was burned only in air and 30% oxygen.

A CFB operates at a much higher fluidizing velocity - above 15 ft/sec - so there is less concern for defluidization. Nevertheless, it was an objective to demonstrate successful operation with a range of fuels at up to 70% oxygen at the injection locations.

4.1.2 Heat Transfer

In a coal-fired boiler, the chemical energy of the fuel is converted to thermal energy in the hot combustion products, which is in turn transferred to the water/steam cycle for power generation. This heat transfer takes place mainly in three areas of a CFB boiler system:

- In the **combustor** the hot flue gas and solids transfer heat to the combustor waterwall and to any additional surface in the gas path.
- In the **External Heat Exchanger**, circulating solids are cooled before returning to the furnace. This is typically a Fluidized Bed Heat Exchanger, but may also be a Moving Bed Heat Exchanger (Jukkola, *et al.*, 2003).
- In the **Convective Pass** after the cyclone, the flue gas and fly ash are cooled by tube banks.

Since an oxygen-fired system has a lower flue gas flow and smaller furnace than an air-fired system, there is less heat transferred in the combustor and in the convective pass. This means the External Heat Exchanger (EHE) must do a much larger proportion of the total heat transfer. Figure 4-1 shows a typical distribution of heat duties comparing air-fired with 70% oxygen firing.

Figure 4-1: Distribution of Heat Duties - Air vs. Oxygen Fired

Besides the different distribution of heat duties, the designer must account for any differences in the heat transfer coefficients due to oxygen firing.

Without the nitrogen in the flue gas, the CO₂ and H₂O contents are higher. Table 4-1 shows typical flue gas compositions. More CO₂ and H₂O increase the non-luminous radiation heat transfer from the gas. The calculated impact of the increased radiation is

that the heat transfer coefficient in the convective pass will be about 10% higher with oxygen firing.

In the furnace and EHE, hot solids are the dominant heat transfer medium. The impact of changes in the gas radiative properties is expected to be much less.

Table 4-1: Typical Flue Gas Composition - Air vs. Oxygen Fired

	Air	30% O₂	70% O₂
N ₂ (%)	74.78	0.81	0.74
CO ₂ (%)	14.49	82.78	74.91
H ₂ O (%)	7.40	13.05	20.97
O ₂ (%)	3.31	3.31	3.31
SO ₂ , ppm	199	469	764

4.1.3 Recarbonation

One of the major benefits of fluidized bed combustion is the ability to capture SO₂ from the flue gas by the addition of limestone to the combustor. The sulfur capture occurs in two steps:

1. The calcium carbonate in the limestone is calcined to calcium oxide.

2. The calcium oxide reacts with SO₂ to form calcium sulfate.

Calcination occurs when the limestone is heated up to a temperature which depends on the CO₂ content of the surrounding gas - see Figure 4-2.

Figure 4-2: Equilibrium Temperature for Calcination

With air firing, the CO₂ content of the flue gas is under 20%. Limestone will calcine at about 1400°F which is well below the typical CFB operating temperature of 1500 to 1650°F.

With oxygen firing, however, the CO₂ content is above 70% (see Table 4-1). This requires a temperature above 1600°F for calcination to occur. There are two consequences of this:

1. The combustor needs to operate at a high temperature to ensure calcination. This can generally be designed for - anthracites and petroleum cokes are typically combusted at above 1600°F in CFB combustors. For some fuels, there may be concerns for ash fusion and sulfur capture in the furnace may suffer at high temperature (see Section 4.1.4).
2. Where the ash cools to below the calcination temperature while exposed to the high CO₂ content, recarbonation (the reverse of calcination) may occur.

Recarbonation is a concern in those locations where the temperature drops below the calcination temperature: fluidized bed or moving bed heat exchanger and the convective pass.

Recarbonation in the External Heat Exchanger

A conventional CFB boiler uses a fluidized bed heat exchanger fluidized with air. Solids are cooled in the FBHE to usually below 1200°F, then returned to the furnace. There is some flue gas entering the FBHE (entrained with the solids entering) and there may be

some combustion of unburned carbon in the ash in the FBHE, so there is some CO₂ in the FBHE gas environment. If there is 1% CO₂ in the gas, the calcination temperature is about 1200°F (from extending the equilibrium curve in Figure 4-2 to lower temperatures). Thus there may be some recarbonation in the FBHE, but this only consumes a small amount of the fluidizing gas and any calcium carbonate formed in the FBHE is re-calcined when it returns to the furnace.

The situation is a bit trickier in an oxygen-fired combustor. If the FBHE is fluidized with air, the air returns to the furnace with the cooled solids. This represents a rather large air leakage into the system, which is undesirable. The FBHE could be redesigned so that the air vents separately. The heat in this air must be recovered, which requires a separate gas cooler.

If the FBHE is fluidized with recycled flue gas there will be recarbonation. In fact, in most cases there would be enough calcium oxide in the ash to remove nearly all of the CO₂, which is over 70% of the flue gas (see Table 4-1). Thus a large initial amount of flue gas must be recirculated and boosted to high pressure to leave enough steam to fluidize the bed.

The recirculated flue gas could be blended with oxygen, but still the CO₂ will be removed down to below 1% in the gas, leaving a high oxygen enrichment diluted with steam.

The option selected for our Phase I design is to replace the FBHE with a Moving Bed Heat Exchanger (MBHE) which requires no fluidizing gas. This MBHE is being developed under a separate contract with DOE/NETL ("Circulating Moving Bed (CMB) Combustion Proof of Concept - Phase 2," DE-FC26-01NT41223).

The option selected for these pilot tests was to modify the FBHE to vent the fluidizing air without the need for heat recovery.

Recarbonation in the Convective Pass

In a conventional CFB, the flue gas leaving the cyclone is cooled in the convective pass followed by an air preheater. In the convective pass, fly ash typically deposits on the tube banks. If necessary, steam sootblowers periodically clean the tubes. As the gas cools to below about 1400°F (see Figure 4-2), there is the potential CO₂ in the flue gas to recombine with calcium oxide in the deposits to form calcium carbonate. This can increase the hardness of the deposits, making them harder to remove.

With oxygen firing, the CO₂ content is higher, so the recarbonation occurs in a broader temperature range and at a higher rate.

There was no convective pass test section in the pilot plant during these tests to investigate this effect.

Recarbonation in the MTF Pilot Plant

The MTF pilot plant has a water-cooled cyclone and inlet duct. Therefore in the pilot plant, the temperature may drop below the calcination temperature in the cyclone, dipleg and sealpot.

Laboratory Demonstration of Recarbonation

An ash sample collected from the FBHE after testing the Tri-Star coal in O₂/CO₂ mixtures was tested in the thermo-gravimetric analysis (TGA) apparatus to evaluate the

recarbonation process described above. The test procedure entailed heating the as-received ash sample in the TGA in 100% CO₂ gas to 1800 °F, and allowing the process to stabilize thermally at this temperature for about fifteen minutes. Subsequently, the temperature was lowered to a given value, and held constant for about sixty minutes. Temperature and weight loss/gain were monitored/recorded continuously throughout the process.

Figure 4-3 and Figure 4-4 show the results of two of these tests. In both cases, the ash increases in weight as the sample heats up. This is due to recarbonation of the calcium oxide in the ash. At about 1640°F, the weight drops sharply as the newly formed calcium carbonate re-calcines to calcium oxide. This is just about the calcination temperature under 1 atmosphere CO₂ as shown in Figure 4-2. In the first test (Figure 4-3) the calcined ash was then cooled to 1500°F, which is well below the calcination temperature. There was a weight gain of about 2 % in one hour at this temperature (equivalent to about 6% calcium utilization). In the second test (Figure 4-4) the ash was cooled to 1675°F, which is just above the predicted calcination temperature. In this case, there was no weight gain - again as expected from Figure 4-2.

Figure 4-3: TGA Recarbonation Test at 1500°F

Figure 4-4: TGA Recarbonation Test at 1675°F

4.1.4 Sulfur Capture

There is generally an optimum temperature for sulfur capture in a fluidized bed; Figure 4-5 is an idealized representation. The general shape of the relationship applies to most fuels, but the optimum temperature varies. For bituminous coals, the optimum is around 1550°F. Lower rank fuels, such as lignite, usually have a lower optimum temperature; anthracite and especially petroleum coke often show improved sulfur capture up to 1650°F and beyond.

Figure 4-5: Sulfur Capture vs. Combustor Temperature

For oxygen firing, the furnace should be above 1650°F to allow calcination (refer to Section 4.1.3). Except for pet coke and anthracite, the ability to capture sulfur will suffer at these high temperatures.

The surface area and porosity of calcined lime depends on the calcination conditions. Laboratory studies at Westinghouse (Ulerich, *et al*, 1977) showed that limestone calcined under higher CO₂ atmosphere produced a sorbent with a greater capacity for sulfur capture. The sorbent produced in an oxygen-fired CFB with over 70% CO₂ may have a higher capacity for sulfur capture than if calcined in 15% CO₂ at the same temperature.

4.1.5 Other Emissions

There are three main sources of nitrogen oxide (NO_x) emissions from the combustion of coal.

- **Thermal NO_x** is formed by dissociation of elemental nitrogen (N₂) in the air. This is only a significant source at combustion temperatures of about 2500°F or higher. It is the largest source of NO_x in pulverized coal combustion systems. In fluidized bed combustion the temperatures are below 1800°F, but there might be a small amount of thermal NO_x formed at the surface of the coal particles where the temperature can be hotter (see Section 4.1.1).
- **Prompt NO_x** is also formed from elemental nitrogen in the air. In this case it is a chemical dissociation by reaction with hydrocarbon fragments. This occurs in both pulverized coal and fluidized bed conditions.
- **Fuel NO_x** forms from the oxidation of nitrogen compounds in the fuel.

NO_x formed by any of these mechanisms can be converted to elemental N₂ by reducing agents such as CO, ammonia, hydrogen, and char. This is enhanced in the furnace by staging the combustion conditions.

Oxygen firing should nearly eliminate thermal and prompt NO_x. The effect will be strongest in pulverized coal systems where the thermal NO_x dominates. Previous testing has indeed shown strong reductions in NO_x - up to a factor of 3 reduction (e.g., Bozzuto, et al., 2001 and Thambimuthu, et al., 1998).

Bench scale tests have also shown lesser reductions in NO_x with oxygen firing at fluidized bed temperatures (Marion, et al. 2003). This is due to the elimination of prompt NO_x and perhaps some thermal NO_x. In addition, the formation of fuel NO_x and the destruction of NO_x may be affected by the different flue gas composition. For example, the much lower N₂ content of the flue gas may enhance the conversion of NO_x to elemental nitrogen.

CO emissions were also found by Marion, et al., (2003) to be higher with oxygen firing than with air firing. It is believed that the high CO₂ concentrations prevailing in the furnace (>70% for oxygen firing vs. ~15% for air firing) may hinder the oxidation of CO to CO₂ (i.e., the reaction $\text{CO} + \text{O}_2 \rightarrow \text{CO}_2$ is suppressed).

Nitrous oxide (N₂O) is a greenhouse gas which is currently not regulated. N₂O emissions have not been reported from these previous studies. It will be monitored during these MTF pilot tests.

4.1.6 Combustion Efficiency

In the Phase I bench-scale testing, Marion, et al. (2003) found the combustion efficiency with oxygen firing to be equal to or better than with air firing.

4.2 MTF Pilot Plant

ALSTOM Power Inc.'s "Multi-Use Combustion Test Facility" (MTF) was developed by its US Power Plant Laboratories to support the Power Generation Businesses strategic development needs. This facility (Figure 4-6) provides the flexibility to perform pilot-scale testing with conventional pulverized-coal firing, fluidized bed combustion, and gasification firing conditions, thus addressing the requirements for several test facilities. The test facility is located in ALSTOM Power Inc.'s Combustion Research Complex at its US Power Plant Laboratories facilities in Windsor, Connecticut.

Figure 4-6: Multi-use Test Facility (MTF)

The MTF also allows testing with both circulating and bubbling fluidized bed conditions, as well as various other conditions being considered for advanced processes. Capabilities for testing under FBC modes provide detailed data on heat transfer, hydrodynamics, combustion, sulfur capture and process control.

Investigations can be conducted with test fuels including coal, oil, and gas as well as

various alternative fuels such as petroleum coke and biomass. Complete solid fuel and additive handling systems, a flue gas scrubbing system and a Fabric Filter Test Facility are also incorporated into the MTF.

4.2.1 General Facility Description

This section gives a description on the MTF in its basic CFB configuration. The modifications made for the oxygen-fired testing are described in Section 4.2.2.

The MTF can be operated under atmospheric conditions at firing capacities up to 9.9 million Btu per hour (3 MWth). The combustor has an overall height of more than 60 feet. The inside diameter is 40 inches in the upper furnace; in the bottom nine feet it tapers to 26" diameter. The area of the fluidizing grid is 42% of the upper furnace area ($26^2/40^2$). The furnace is equipped with extensive instrumentation and control systems and is housed in an enclosed building with supporting ancillary equipment.

Combustion air is supplied through a Spencer forced draft fan. The combustion air is split into underbed and overfire air streams. The underbed air passes through an electric heater, where it can be preheated up to 1000°F. The underbed air then enters a plenum, before passing through the air distributor.

Overfire air is injected into the furnace at one or more locations. A large number of ports are available for evaluating the effect of overfire air location. The overfire air connects to the combustor ports with high temperature flexible tubing, which makes the relocation of overfire air locations a rapid and easy process.

The combustor is made of several modular sections. The upper combustor sections are lined with 10-inch thick refractory. This refractory liner consists of a composite of three layers: 4" of refractory brick on the interior surface followed by 4" of low density insulating refractory and 2" of mineral wool board against the facility housing.

At two elevations along the combustor, there are 1 foot wide by 68" tall openings to accommodate water-wall test sections for heat transfer measurements. When the water-wall panels are not used, the openings are fitted with flat water-cooled panels with a thin refractory covering.

Additional heat transfer surface can also be installed in the upper furnace if desired - e.g., horizontal tube bundles or vertical wing walls.

The hot combustion gases and solids exit the top of the combustor and enter a refractory-lined, water-cooled cyclone, where the circulating solids are separated from the hot gases. The separated solids drop through a dipleg into a sealpot. The dipleg can be water-cooled, steam cooled, or uncooled. When the dipleg is cooled, the solids recirculation rate through the combustor can be calculated from a heat balance across the dipleg. The hot solids in the sealpot either return directly to the combustor through an insulated stainless steel solids return pipe, or a portion of the solids may be diverted to one of two water-cooled heat exchangers before returning back to the combustor. The main heat exchanger is a fluid bed heat exchanger, similar to that in a commercial design. The second heat exchanger has at different times been configured as a fluid bed heat exchanger, a moving bed heat exchanger, and a falling solids heat exchanger. The heat exchangers are used to cool the recirculating solids and thus control the combustor bed temperature. They provide the test facility with a great deal of flexibility in operating the

combustor over a wide range of process conditions.

Circulating ash can be drained from the FBHE into 55-gallon drums as needed to help control furnace inventory. This ash can be added back into the furnace if necessary to increase inventory. Otherwise it may be saved as startup material for future tests.

The hot flue gas leaving the cyclone flows through a water-cooled heat exchanger in the back room. The cooled gases then flow through a fabric filter then to a wet caustic scrubber for final SO₂ and particulate control. The baghouse can be bypassed if desired - e.g., during warm-up. The induced draft fan and the stack follow the wet scrubber.

The baghouse has been modified into a Flash Dry Absorber (FDA) test system. The FDA system is a dry process based on the reaction between SO₂ and Ca(OH)₂ in humid conditions. Additional equipment for the FDA test system include a FDA mixer/hydrator, additive feed system, FDA reaction duct, modification of the flue gas ducting, and additional gas analyzers, instrumentation, and controls systems.

The fly ash collected in the baghouse is discharged through a screw into 55-gallon drums which are weighed then saved or disposed of as required.

The combustor is warmed up with a natural gas igniter, which is sized for a maximum heat input of 9.9 million Btu per hour. The igniter is located 4.5 feet above the air distributor, with the flame directed downward toward the bed at a 55° angle. Crushed coal and sized limestone are supplied to the combustor through the fuel feed system. Coal and limestone are metered from the storage silos by gravimetric feeders and are then lifted up to the feed inlet chute by a drag chain conveyor. The fuel and sorbent drop through a rotary valve either directly into the furnace at one of two elevations or into the return pipe that carries the hot recycle solids back to the fluidized bed.

A drain port is located on the opposite side of the bed for removing large rocks and for maintaining bed level. The hot ash removed in the bed drain system passes through a water-cooled screw into 55-gallon drums which are weighed then saved or disposed of.

The Multi-Use Combustion Test Facility uses an ABB Advant 460 distributed control system for the process control and data acquisition needs of this facility, and for the other major combustion facilities in the Combustion System Development Complex. The MTF is very well instrumented, with over 500 temperature, pressure, and flow measurements throughout the facility. Figure 4-7 is a simplified Process and Instrumentation Diagram (P&ID) for the main furnace system - as set up for the oxygen-fired tests.

A LabView data acquisition system is used to collect these measurements and process calculations from the Advant system. The LabView program provides on-line trend analysis, data archival, and data analysis.

Ports are located at 16 different elevations along the height and around the circumference of the combustor. They provide a great deal of flexibility for detailed in-furnace measurements, overfire air location, and observation ports. Test probes are used to measure process conditions both radially and axially within the furnace. Typical test measurements across the combustor profile can include solids loading and composition, local solid fluxes, gas composition and temperature, and local heat flux. The gas analysis system allows measurement of important species, including O₂, CO, CO₂, SO₂, NO_x, N₂O, and THC. Gas samples can also be collected for more detailed species analysis in a gas chromatograph.

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

Figure 4-7: Simplified P&ID for the MTF

Logged Data

Hundreds of data points are monitored and logged by the Advant and LabView systems.

- Over 150 Type-K thermocouples are installed on the MTF to measure air, flue gas, water, combustor and refractory temperatures.
- Over 40 pressure cells are installed on the furnace and gas ducts.
- Water flows are measured with turbine flow meters.
- Coal and limestone belt feeder rates are logged. These feeders are calibrated before each test and periodically checked during each test by collecting material off the belt for ½ to 2 minutes.
- Additional pressure differential cells are used (along with pressure and temperature) to calculate air, natural gas, and steam flow.
- An in-situ Rosemount O₂ analyzer measures the wet oxygen content of the flue gas in the back room before the heat exchanger.
- At the same point, a gas sample is extracted, filtered, drawn through a heated sample line to the control room, and dried. Analyzers measure O₂, CO₂, CO, NO, NO₂, N₂O, NO_x, and SO₂. The analyzers usually operate continuously with purges every hour or so as the filter pressure drop increases. The analyzers are calibrated twice per day.
- The gas analyzers in the control room can be switched over to analyze in-furnace gas samples which are extracted and filtered using a water-cooled gas sampling probe.
- Waterwall panels can be installed in the furnace to obtain heat transfer data.
- There are also water-cooled heat transfer probes for measuring local total and radiation heat flux throughout the furnace.

Solids Samples

Samples are taken at several locations.

- Coal and limestone samples are taken normally off the feed belts periodically and mixed together for a composite each 8 or 12-hour shift.
- Bed drain material from the water-cooled screw outlet is regularly sampled.
- FBHE drain material is usually taken for analysis of the circulating material. There are also ports in the heat exchanger box for directly withdrawing samples.
- Samples of fly ash are also collected from the baghouse drain.
- There are several water-cooled solids probes for collecting samples from the furnace. These probes can also measure the local solids flux. There were no in-furnace solids samples taken during these tests.
- Crossover solids at the cyclone inlet are sampled at the calculated average isokinetic conditions with the water-cooled solids probe to determine the rate of solids circulation and the size distribution of the solids entering the cyclone. This sampling

can be done at different locations horizontally and vertically across the cyclone inlet; as expected the solids loading is higher at the top and lower at the bottom of the duct. The profile is roughly linear so a single sample at the midpoint is often used to determine the solids loading.

- Isokinetic dust load can be measured in the downflow water-cooled duct after the cyclone according to EPA Method 5. Sampling should be done at a grid of points covering the entire duct, although physical access constraints makes it difficult to cover the entire duct. No Method 5 sampling was done during these tests.
- A High Volume technique is used to collect a larger fly ash sample at the calculated average isokinetic conditions of the duct at 4 points along one axis. The original purpose of this method was to get a reasonably unbiased sample for size and composition analysis. It turns out that the measured dust load is often quite accurate as well.

Selected solids samples are analyzed as required. Samples not analyzed are retained for future use as needed.

4.2.2 Facility Modification for Oxygen Firing

The O₂/CO₂ supply and control infrastructure and other modifications made to the MTF furnace and ancillary equipment are discussed below.

Oxygen and Carbon Dioxide Supply Infrastructure

In the commercial design for oxygen-fired boilers, pure oxygen is delivered to the plant and is tempered with recirculated dried flue gas in order to achieve the desired oxygen content of the net oxidant. For the pilot plant testing, mixtures of oxygen and pure carbon dioxide were used, each supplied by purchased liquefied gases. This approach is more cost effective for the short term testing and allows additional flexibility of control not afforded by recirculating flue gas.

An O₂/CO₂ supply and control infrastructure, designed and supplied by Praxair, Inc., was integrated into the MTF facility to enable the combustion of fuels in various O₂/CO₂ mixtures (Figure 4-8).

The Spencer forced draft fan was not used at all during these tests. Additional compressors were used to supply the combustion air for startup as needed and for the air fired tests.

In the MTF's normal configuration, the primary combustion air is fed to a plenum below the furnace which distributes the fluidizing air to a number of nozzles. For the oxygen firing, the plenum and nozzles were replaced with a header and three spargers to distribute the fluidizing oxidant.

Figure 4-8: O₂ and CO₂ Supply Tanks and Control Skids

Discussion of Global and Local O₂ Enrichments

Oxygen and carbon dioxide from tanks are blended to simulate the mixture of pure oxygen and recirculated flue gas used for the oxidant in the combustor. In addition, some pure CO₂ bypasses the mixing skid to be used for various purposes where high oxygen mixtures were not desired, *e. g.*,

- seal pot fluidization
- coal assist "air"
- ignitor cooling "air"

There is also air introduced into the system as leakage as well as air used to fluidize the sealpot. Figure 4-9 indicates the various flows.

As a result, there are several definitions of the oxygen content of the oxidant.

Global O₂ - The concentration of O₂ in the overall O₂/CO₂ mixture from the tanks, which includes the bypass CO₂. There is no actual gas mixture at this concentration. This mixture represents the overall ratio of O₂ and CO₂ entering the system and is used for normalizing the emissions, among other things.

Local O₂ - The actual combustion mixture of O₂/CO₂ as it comes from the mixing skid. This has a higher O₂ content than the global mixture. This is the oxidant mixture which enters the furnace and at the bottom of the furnace first sees the coal in the fluid bed. This is the relevant mixture for materials of construction of the fluidizing nozzles and for

concerns about high oxygen concentrations resulting in elevated combustion temperatures at the coal surface, leading to agglomeration (see Section 4.1.1).

Overall O_2 - The oxygen concentration of the overall oxidant, including the air leakage. This value is not much used in the analysis of the results.

Figure 4-9: Schematic of Oxygen and Carbon Dioxide Flows to the MTF

Because of the small scale of the pilot plant, the bypass CO_2 requirements and the amount of air leakage are relatively large, so the differences between the three mixture definitions are much larger than they would be in a full-scale commercial unit.

Reduced Furnace Diameter

When firing with an enriched oxygen content, the volume of flue gas produced is reduced for a given firing rate. This results in a lower velocity through a given furnace size, or a reduced furnace size to maintain a fixed velocity.

To simulate the proper fluidizing velocities in the pilot plant, either the firing rate has to increase or the furnace diameter decrease. Since the pilot plant's environmental permit is limited to 9.9 MMBtu/hr firing rate, the furnace diameter was reduced from 40 inches to 21 inches. Figure 4-10 shows the calculated firing rate vs. the global oxygen content for constant conditions of 15 ft/sec furnace velocity and 3.5% O_2 in the flue gas. With the furnace diameter reduced to 21 inches, up to 70% oxygen enrichment can be used while meeting the permitted firing rate.

Figure 4-10: MTF Firing Rate vs. Oxygen Enrichment

Figure 4-11 shows the top of the furnace with the roof removed. The original refractory and brick (light color) is 10 inches thick with 40 inch ID. The additional refractory lining (dark color) is 9.5 inches thick bringing the ID to 21 inches.

The refractory lining added an additional 15 tons to the weight of the furnace. To support this extra weight, three steel posts were attached to the lower part of the furnace. The posts were supported by constant load springs on the floor on the facility.

Figure 4-11: MTF Permanent and Temporary Refractory Lining

Cyclone

As a result of the reduced flue gas flow, the velocities into, within, and leaving the cyclone also decrease. Calculations showed that the velocity within the cyclone barrel should still be sufficient to give an acceptable cyclone collection efficiency, so the barrel diameter was not reduced.

For the first test week, a steel baffle was installed in the cyclone inlet duct to increase the velocity entering the cyclone barrel. This has been done in the past with the normal flue gas flows to increase the collection efficiency. The cyclone seemed to operate well during the first test week, but the baffle was slightly warped at the end of the test.

For the second test week, refractory brick was installed in the duct to reduce the area to $\frac{1}{4}$ the original size. This increased the inlet velocity to roughly the design value. In addition, a steel liner was placed in the cyclone outlet duct to similarly increase the cyclone exit gas velocity.

Additional Heat Transfer Surface

In an air-fired combustor, much of the fuel's energy is carried away by the hot flue gas, which includes a large amount of nitrogen gas. Most of this sensible energy in the gas is recovered in the convective pass and air preheater.

With oxygen firing at low enrichment, this nitrogen diluent is replaced by CO_2 . At high enrichment, the flue gas has less dilution; therefore additional cooling is needed to control the combustion temperature to the desired level. Table 4-2 shows typical calculations for the MTF with the diameter reduced to 21".

Table 4-2: MTF Cooling Requirements vs. O₂ Enrichment

Oxidant	Firing Rate, MMBtu/hr	Cooling Requirement, MMBtu/hr
Air	2.61	1.41
21% O ₂	2.61	0.95
30% O ₂	3.89	2.18
40% O ₂	5.26	3.51
50% O ₂	6.57	4.78
70% O ₂	9.06	7.19

The cooling requirement is met with surface in the furnace and cooling of solids in the primary recirculation loop with the FBHE and the cooled dipleg.

Because of concerns that there would not be enough existing cooling capacity to operate above 50% O₂, an additional cooling coil was installed in the upper furnace. The cooling coil was located around the perimeter of the furnace in the upper 10 feet of furnace height (shown on Figure 4-7).

Heat Transfer Test Sections

With the additional refractory lining installed in the furnace, the heat transfer panels were covered up. In order to get some waterwall heat transfer data, two single tube test sections were installed in the furnace.

To make these sections, lengths of 2-inch and 1¼-inch stainless steel pipe was split lengthwise and placed together to form an annular region for cooling water to flow (see Figure 4-12). The small annular design allows for a high velocity in the annular region with a low water flowrate. The high velocity means a low resistance to heat transfer on the inside of the tube; the measured heat transfer thus more accurately represents the furnace side, which is of interest. The low flowrate means that heat pickup results in a higher temperature rise in the water, which increases the accuracy of the measurement.

Figure 4-12 also shows one assembled tube and a sketch of how they are installed in the furnace. The locations of the tubes are indicated on Figure 4-7. The lower tube is 36" long, the upper is 59" long.

Figure 4-12: Split Tube Heat Transfer Test Section

Venting of the FBHE

In order to avoid the recarbonation reaction discussed above in Section 4.1.3, the MTF's fluidized bed heat exchanger (FBHE) was modified as shown in Figure 4-13. This design was such that the fluidizing air was vented off into the atmosphere outside the building

complex. In this manner, the cooled solids could be recirculated into the furnace with only a small entrainment of air.

Figure 4-13: Modified Fluidized Bed Heat Exchanger Showing Air Vent

Sealpot

Because of the water-cooled cyclone, the pilot plant sealpot runs cooler than its commercial counterpart. With oxygen firing, this means significant recarbonation can (and did) occur in the sealpot. The sealpot was replumbed to allow fluidization with, CO₂, air, or a mixture of the two.

The air, which fluidizes the sealpot, enters the combustor, representing a significant air "leak." To minimize the amount of air used, the inside of the sealpot was partly blocked off to reduce the plan area and therefore the amount of gas needed to fluidize the sealpot.

Instrumentation and Control System

The materials and instrumentation for the oxygen supply and distribution were specified for service in high oxygen environment. Most of the existing instrumentation on the facility was suitable for the oxygen testing, since in the furnace, the combustion reduces the oxygen content to the typical range of 3 - 4%.

A furnace of this size burning gas and coal - even with air - does have some risks. The

control system has been programmed to handle such situations as loss of fuel, and temperature or pressure excursions.

With oxygen firing, there is the additional concern of avoiding high O₂ concentrations where they aren't wanted. If the fuel trips or plugs while firing a high oxygen mixture, the furnace and backend systems would see the high O₂ levels. If the CO₂ supply should stop, pure oxygen would enter the furnace - the results would not be pretty.

The control system logic was modified to detect and respond to these situations.

4.2.3 Relevant Differences Between the Pilot Plant and a Commercial Unit.

The MTF pilot plant is a good model of a commercial CFB boiler - it comprises most of the components of the commercial system and is large enough to simulate the process without gross distortions due to scale. There are, however, several differences which must be kept in mind when evaluating the test results.

The most obvious is the difference in scale, especially the smaller **cross sectional area**. This has several impacts. One is that everything is much more uniform across the cross section. The MTF has an inner diameter of 40", which was reduced to 21" for this O₂ fired testing. A large commercial unit may have dimensions of 25' x 80'. Even smaller commercial units have shown strong maldistributions, for example, a plume of nearly zero oxygen may extend up the entire furnace height above the fuel feed locations, while at the opposite wall the oxygen content may be 10%. We have seen maldistributions in the MTF, but they are much less severe and do not extend the whole height of the unit. As a consequence, pilot plants in general have lower CO and SO₂ emissions, which benefit from improved mixing. On the other hand, NO_x emissions tend to be higher in a pilot plant, since the better mixing reduces the beneficial effect of staged combustion. It is also not possible, for example, to accurately simulate the impact of fuel feed locations.

Another consequence of the reduced scale is the much greater **surface-to-volume ratio** in the pilot plant. If the pilot furnace were of waterwall construction, the heat removal would be much too large to allow coal combustion conditions. The pilot plant is refractory lined with the possibility for some heat transfer sections along the height of the furnace. The heat removal profile along the height is therefore different, which affects the temperature profile as well. The large internal refluxing of solids along the height of the furnace does tend to smooth out the temperature profile, but it is not perfectly uniform in either the pilot or commercial units.

A typical commercial unit has a **furnace height** of over 100', while the pilot is about 62 feet tall. This affects the gas residence time in the furnace. This impact is somewhat lessened by the fact that the furnace operates with a **superficial velocity** slightly lower than current commercial designs - 15 ft/sec vs. 18 ft/sec or higher.

There are other features of the small-scale pilot that may matter. For example, as a practical matter, the sealpot is proportionally large compared to the scale of the furnace, so the fluidizing airflow to the seal pot is relatively large.

The roof of the **furnace, the cyclone, and the cyclone inlet duct are water-cooled** (with one inch of hard face refractory for wear protection rather than insulation). Due to the heat pickup from these cooled surfaces, the cyclone outlet temperature is cooler than the

top of the furnace (see Figure 4-21 through Figure 4-23). In a commercial unit, the cyclone outlet temperature is often hotter than the top of the furnace, as CO and O₂ which never mixed in the furnace combine and burn. This temperature difference can affect the reactions occurring in the cyclone, such as carbon and CO burnout, sulfur capture, and NO_x control by ammonia injection.

The **dipleg** is made to operate water cooled, steam cooled, or uncooled. Combined with the low temperature at the cyclone, cooling of the dipleg reduces the temperature of the solids in the sealpot even further. This was especially relevant to this oxygen fired test program, since the sealpot operated below the calcination temperature (see Section 4.1.3 and Figure 4-2). The sealpot would not operate with CO₂ fluidizing only - the calcium oxide in the ash reacted with the CO₂, leaving no fluidizing gas. For these tests, the sealpot was fluidized with air or air/CO₂, which introduced additional air leakage into the system.

Another difference relevant to the oxygen-fired tests was the use of **pure CO₂** for tempering the oxygen, rather than recirculated flue gas.

Table 4-3 shows an example of the differences in flue gas composition leaving the combustor between firing with pure O₂/CO₂ mixtures and with recirculated flue gas (FGR).

Table 4-3: Flue Gas Recirculation vs. Pure CO₂

	Air Firing	30% Oxygen		70% Oxygen	
		FGR	Pure CO ₂	FGR	Pure CO ₂
N2 (%)	74.78	0.81	0.22	0.74	0.50
CO2 (%)	14.49	82.78	88.2	74.91	77.07
H2O (%)	7.40	13.05	8.24	20.97	19.06
O2 (%)	3.31	3.31	3.31	3.31	3.31
SO2, ppm	199	469	302	764	698
SO2, dry	215	540	329	967	862
Based on Bituminous coal with 2.3%S fired to a constant excess oxygen 90% S Capture in Boiler, 80% in backend for 98% total capture Flue Gas dried to 7% H2O Before Flue Gas Recirculation					

When firing with recirculated flue gas, sulfur and moisture are returned back to the combustor resulting in higher concentrations in the flue gas (recirculation loop). The consequence of using pure CO₂ is a higher CO₂ content in the flue gas with other components somewhat lowered. Compared to the major difference in gas composition going from air-fired to oxygen-fired, the changes due to using pure CO₂, were not significant for these tests.

4.3 Fuel and Limestone

4.3.1 Fuel Feeder Belt Sample Analyses

Tri-Star mvb coal and CONOCO Phillips' petcoke were sampled from the MTF belt feeder throughout the testing. The proximate and ultimate analyses and higher heating values from several composite samples were determined, and are reported in Table 4-4. These analyses are consistent with those obtained previously on the mvb coal and petcoke (See Table 3-1).

Table 4-4: Fuel Feeder Belt Sample Analyses

Fuel	Tri-Star Medium Volatile Bituminous (mvb) Coal: April 2004 Test Campaign					Tri-Star Medium Volatile Bituminous (mvb) Coal: June 2004 Test Campaign			CONOCO Phillips' Petroleum Coke: June 2004 Test Campaign
Sample Identification #	4-0291-C	4-0292-C	4-0293-C	4-0294-C	Average Values	4-0332-C	4-0333-C	Average Values	4-0334-C
Date	4/02-03	4/03-04	4/04-05	4/05-06		6/05-06	6/06		6/06-07
Prox. & Ult. Analyses, Wt.% As Received ^[1]									
Moisture	4.4	6.1	5.5	5.6	5.4	4.1	9.0	6.6	1.0
Volatile Matter	16.6	16.7	16.7	16.9	16.7	17.0	16.4	16.7	9.1
Fixed Carbon	63.4	62.1	62.2	62.9	62.7	63.0	60.0	61.5	89.5
Ash	15.6	15.1	15.6	14.6	15.2	16.0	14.5	15.2	0.5
					0.0				
Hydrogen	3.8	3.7	3.7	3.7	3.7	3.7	3.5	3.6	3.6
Carbon	69.8	68.7	68.8	69.5	69.2	69.1	66.1	67.6	87.1
Sulfur	1.4	1.4	1.4	1.4	1.4	1.6	1.4	1.5	5.6
Nitrogen	1.6	1.6	1.7	1.7	1.7	1.6	1.5	1.5	1.6
Oxygen	3.4	3.4	3.3	3.5	24.0	4.0	4.0	4.0	0.8
Volatile Matter, Wt.% Dry-Ash-Free	20.8	21.2	21.2	21.2	21.1	21.2	21.5	21.3	9.2
Carbon, Wt.% Dry-Ash-Free	87.3	87.2	87.2	87.1	87.2	86.4	86.4	86.4	88.3
HHV, Btu/lb	12151	11865	11893	12029	11985	12377	11594	11986	15071
FC/VM (Fuel Value)	3.8	3.7	3.7	3.7	3.7	3.7	3.7	3.7	9.9
Ash _{Loading} , lb/MMBtu	12.8	12.7	13.1	12.1	12.7	12.9	12.5	12.7	0.3
Nitrogen _{Loading} , lb/MMBtu	1.3	1.3	1.4	1.4	1.4	1.3	1.3	1.3	1.1
Sulfur _{Loading} , as SO ₂ , lb/MMBtu	2.3	2.4	2.4	2.3	2.3	2.5	2.3	2.4	7.4

^[1] Unless otherwise specified

Samples of each fuel were screened for particle size distribution. Screen results along with Rosin-Rammler parameters (Field, et al., 1967, p. 245) are summarized in Table 4-5 and Figure 4-14.

Table 4-5: Fuel Feeder Belt Sample Particle Size Distributions

Tri-Star mvb Coal		CONOCO Phillips Petcoke	
Size, X, μm	Wt% < X	Size, X, μm	Wt% < X
4760	84.30	4760	96.80
2380	67.20	2380	51.39
1190	49.24	1680	24.50
590	33.92	1190	16.90
297	19.05	840	13.41
149	9.97	590	10.70
		297	6.13
Rosin-Rammler Parameters		Rosin-Rammler Parameters	
n (slope)	2.17	n (slope)	1.40
x' (Fineness factor), μm	2042	x' (Fineness factor), μm	2959
x _{mw} (Mean Weight Particle Size), μm	2279	x _{mw} (Mean Weight Particle Size), μm	2698

Figure 4-14: Fuel Feeder Belt Sample Particle Size Distributions

4.3.2 Limestone Feeder Belt Sample Analyses

Chemstone limestone was used in the first test week for SO₂ control; Aragonite was used during the second test week. The analyses of selected samples, Table 4-6, are consistent with those obtained earlier from the two limestone samples (see Table 3-3).

Table 4-6: Limestone Feeder Belt Sample Analyses

Limestone	Chemstone: April 2004 Campaign					Aragonite: June 2004 Campaign
Identification #	4-0295-LS	4-0296-LS	4-0297-LS	4-0298-LS	Average Values	4-0331-LS
Beginning Time	4/02-03	4/03-04	4/04-05	4/06		6/06
Weight % Dry						
Total Ca	38.54	38.86	38.81	38.77	38.74	37.23
Total Mg	0.30	0.29	0.30	0.30	0.30	0.20
CO ₂	43.58	43.55	43.55	43.61	43.57	42.49
CaCO ₃	96.23	97.04	96.93	96.81	96.75	92.98
MgCO ₃	1.04	1.01	1.03	1.04	1.03	0.70
Inerts (Diff.)	2.73	1.95	2.05	2.16	2.22	6.32

Screen analyses for selected Chemstone and Aragonite samples from the MTF's fuel and limestone feeder belts, along with Rosin-Rammler parameters, are summarized in Table 4-7 and Figure 4-15.

Table 4-7: Limestone Feeder Belt Sample Particle Size Distributions

Chemstone		Aragonite	
Size, X, Micron	Wt% < X	Size, X, Micron	Wt% < X
1190	99.96	1190	98.73
590	60.59	590	87.13
297	27.55	297	56.16
149	9.14	210	26.70
74	3.08	149	7.28
		74	1.09
Rosin-Rammler Parameters		Rosin-Rammler Parameters	
n (slope)	1.92	n (slope)	2.17
x' (Fineness factor), μm	497	x' (Fineness factor), μm	453
x _{mw} (Mean Weight Particle Size), μm	441	x _{mw} (Mean Weight Particle Size), μm	402

Figure 4-15: Limestone Feeder Belt Sample Particle Size Distributions

4.4 Test Description

4.4.1 Test Matrix and Objectives

The pilot plant testing was done in two campaigns of roughly one week each.

Test Week 1

The objectives for the first test week (Figure 4-16) were

- Run with the Tri-Star coal air-fired
- Run with the Tri-Star coal on 21% O₂ in CO₂
- Run with the Tri-Star coal with 30% O₂ in CO₂ and higher O₂ enrichments, as possible.
- Run with the pet coke with 30% O₂ in CO₂ and higher O₂ enrichments, as possible.

Figure 4-16: MTF Test Matrix - Week 1

The first three objectives were achieved in the first test week. The pet coke firing was left for the second test week.

Test Week 2

For the second test, the additional cooling coil was installed in the upper furnace to allow higher firing rates with higher oxygen enrichments.

The objectives for the second test week (Figure 4-17) were

- Run with the base coal air-fired and at 21% O₂ in CO₂ for heat transfer comparison. The furnace would be overfired (higher than 15 ft/sec gas velocity) because of the extra cooling surface.
- Run with the base coal with 40% O₂ in CO₂ and higher O₂ enrichments, as possible.
- Run with the pet coke with 40% O₂ in CO₂ and higher O₂ enrichments, as possible.
- Run with Pitt 8 with 40% O₂ in CO₂ and higher O₂ enrichments, as possible.

Figure 4-17: MTF Test Matrix - Week 2

Because of the long startup of Test Week 2 (see Section 4.4.2), the air and 21% O₂ tests were deferred to later in the week if time allowed. In fact, the test ended after one day of pet coke firing, so neither the air / 21% O₂ comparison nor the Pittsburgh #8 firing were done.

4.4.2 Test Points

From the two weeks of testing, several test points have been defined when the furnace was at certain specified conditions - see Table 4-8. The duration of these test points ranges from 15 minutes to over six hours. They are not necessarily considered to be **steady state** points.

Table 4-8: Selected Test Points

Test Point	Start	End	Duration	% Global O ₂	% Local O ₂	% Primary Flow	QFired MMBtu/hr	Ca:S	% O ₂ dry
A1	4/2/04 18:00	4/3/04 00:10	06:10	21	21	48	3.96	2.0	3.9
B1	4/3/04 20:00	4/4/04 02:00	06:00	21	26	60	4.07	2.0	3.2
B2	4/4/04 05:20	4/4/04 08:00	02:40	20	26	51	3.81	2.1	3.4
B3	4/6/04 12:00	4/6/04 15:30	03:30	27	40	59	2.24	2.1	7.9
B4	4/7/04 05:45	4/7/04 06:55	01:10	31	40	60	4.78	2.0	2.8
B5	4/7/04 12:45	4/7/04 14:00	01:15	30	40	61	3.94	2.0	4.3
C1	6/4/04 19:30	6/4/04 23:00	03:30	36	49	74	4.24	1.6	9.3
C2	6/5/04 03:45	6/5/04 04:50	01:05	37	50	77	5.69	2.0	4.2
C3	6/5/04 07:30	6/5/04 11:30	04:00	43	60	74	6.57	2.0	3.5
C4	6/5/04 12:35	6/5/04 13:30	00:55	49	67	73	7.87	2.0	3.4
C5	6/6/04 06:00	6/6/04 08:50	02:50	50	70	72	7.57	2.1	4.9
C6	6/6/04 09:30	6/6/04 09:45	00:15	50	70	63	7.56	2.1	3.9
C7	6/6/04 10:20	6/6/04 11:10	00:50	50	70	54	7.57	2.1	4.4
D1	6/6/04 14:00	6/6/04 16:30	02:30	43	61	71	6.28	2.0	2.9
D2	6/6/04 17:00	6/6/04 21:50	04:50	49	70	70	7.08	1.9	3.1
D3	6/7/04 01:30	6/7/04 03:50	02:20	49	70	63	7.92	1.7	3.3
D4	6/7/04 04:10	6/7/04 05:35	01:25	49	70	44	7.92	1.7	2.9
D5	6/7/04 06:30	6/7/04 12:00	05:30	49	70	62	7.62	1.8	2.7

The test points are labeled from A1 to D5 according to

A	Week 1	air-fired	Tri-Star bituminous	Chemstone Limestone
B	Week 1	low O ₂ enrichment	Tri-Star bituminous	Chemstone Limestone
C	Week 2	high O ₂ enrichment	Tri-Star bituminous	Aragonite
D	Week 2	high O ₂ enrichment	pet coke	Aragonite

A brief discussion of each test point follows.

Test Week 1 - The first test week started on Wednesday, March 31, 2004 with air firing. After achieving full coal firing and continuing heatup, a large leak was discovered in the gas duct between the baghouse and the scrubber. After repairs, full coal firing was achieved again on Friday morning, April 2. The first test conditions were reached that evening.

Test Point A1 - A six-hour run with air firing after achieving full temperature and combustor solids inventory. The primary air flow (ratio of air through the fluidizing grid to the total air flow) was about 50%.

Test Point B1 - First run with oxygen firing @ 21% O₂ global. Primary flow is 60% of

total.

Test Point B2 - Decreased primary flow to 50% of total. NO_x increased while SO₂ and CO emissions decreased, which are all typical of a decrease in primary flow (increased "air" staging).

Test Point B3 - First test at nominal 30% global O₂, achieved 27% global / 40% local O₂. Some trouble trying to maintain the furnace temperature at full load with the FBHE, so this test is at about 1/2 load with 7.9% O₂ in the stack

Test Point B4 - Full load at 30% global O₂ using the steam-cooled dipleg to achieve control of the furnace temperature.

Test Point B5 - Re-establish B4 conditions after a trip of the O₂/CO₂ skids. The furnace inventory is higher than in B4, which explains the increase in sulfur capture (89 vs. 78%).

Test Week 2 - The second test week started on Tuesday, June 1, 2004 with the additional cooling coil installed in the upper furnace to accommodate the increased firing rate. This startup was plagued by a number of problems, some general site issues (*e.g.*, a major leak in the cooling water supply lines and several power interruptions) and some related to the MTF (*e.g.*, the scrubber recirculation line plugged and the coal feeders going out of calibration due to one of the power interruptions). Aside from these, the heatup was very slow due to the in-furnace coils. Charcoal briquettes were used to bring the lower furnace temperatures up enough to begin coal firing. Full coal firing was finally achieved on Friday evening, June 4.

Test Point C1 - Starting up on higher O₂: 36% global and 49% local. The furnace inventory was very low resulting in a wide spread in furnace temperatures. The stack O₂ was also very high - 9.3%.

Test Point C2 - A second, one-hour test at 50% local O₂. The furnace inventory was still low with a spread in temperatures.

Test Point C3 - Increased O₂ enrichment to 40% global / 60% local for a four hour test.

Test Point C4 - Increased O₂ enrichment to 50% global / 67% local for a one-hour test before a trip due to low house air pressure.

Test Point C5 - Re-established C4 conditions for nearly three additional hours.

Test Point C6 - A very brief test with primary flow reduced from 72% to 63%.

Test Point C7 - Primary flow reduced further to 54%. As expected, going from point C5 to points C6 and C7 decreased the NO_x and increased the CO and SO₂ emissions.

Test Point D1 - The transition to pet coke firing was very smooth. This first point was at 43% global O₂ (61% local) with 71% primary flow.

Test Point D2 - Increased the O₂ mixture to 49% global (70% local).

Test Point D3 - Decreased the primary flow to 63%.

Test Point D4 - Decreased the primary flow further to 44%.

Test Point D5 - Returned to D3 conditions with 62% primary flow for 5-1/2 hours. At that point, the bottom of the cyclone hopper plugged, causing a loss of circulation and a

spike in the furnace temperature, forcing a shutdown. End of Test Week 2.

4.4.3 Firing Rates

The calculated firing rates for constant velocity (15 ft/sec) and constant excess oxygen (3.5% in the flue gas) were shown in Figure 4-10 and Table 4-2. These calculations are based on the Tri-Star coal analysis; the results are nearly the same for the pet coke. As Figure 4-10 shows, below 70% global O₂, the firing rate and velocity can be increased without exceeding the MTF's environmental permit. Since current commercial designs have an upper furnace velocity higher than 15 ft/sec, this was done in most tests.

The actual firing rates - calculated from the fuel feed rate - are listed in Table 4-8 and shown in Figure 4-18. The calculated velocities in the upper furnace and at the grid are plotted in Figure 4-19.

The low-O₂ tests A1, B1, and B2 had high firing rates and upper furnace velocities above 20 ft/sec. The part load test B3 had a low firing rate and velocity. The other tests had calculated velocities of between 15 and 18 ft/sec.

The velocity at the bottom of the furnace varied according to the primary air split of the test condition.

Figure 4-18: Actual Firing Rates

Figure 4-19: Calculated Velocities

4.5 Furnace Temperature and Pressure Profiles

Temperature profiles along the furnace height and pressure profiles along the primary recirculation loop are useful indications of the furnace conditions. Figure 4-20 gives a key to the temperature and pressure locations in the MTF.

Temperatures are shown from the bottom of the furnace (Point A) to near the top (Point D). There was no thermocouple at the top of the furnace column (Point E) since the additional cooling coil was installed here during Test Week 2. The profile continues with 2 temperatures in the crossover / cyclone inlet duct (Point F) and the cyclone gas outlet temperature (Point G).

The pressure profile is shown from the bottom of the furnace (Point A) to the top (Point E). Most of the profiles show the expected high solids density at the bottom decreasing towards the top. (Note that the pressure cell at Point C is often reading low compared with the rest of the profile.) There is a further pressure drop as the gas and solids turn and accelerate into the crossover duct (Point F) and in the cyclone (Point H). A solids column in the dipleg (Point I), the sealpot (Point J), and the return leg (Point K) provides the pressure seal in the return path to the bottom of the furnace. Without this pressure seal, gas would bypass from the furnace up this return leg to the cyclone. The pressure cell at Point K was not installed during Test Week 1.

Figure 4-20: Key for Temperature and Pressure Profiles

The furnace temperature profiles for each of the defined test points are given in Figure 4-21 through Figure 4-23. In Test Week 1 - without the cooling coil - (Figure 4-21), the temperature rises uniformly along the furnace height; combustion takes place throughout the furnace. The temperature starts to drop in the water-cooled crossover duct, even though combustion usually continues through the cyclone.

The elevations in the figures are referenced from the bottom of the lower combustor section, which is approximately 5 inches above the lower O₂/CO₂ spargers.

In Test Week 2, the cooling coil was installed in the upper furnace (above Point D). This caused the temperature to cool off in the upper furnace, even at as low an elevation as Point C.

Note that in Test Points C1 and C2, the temperature was not uniform in the lower combustor. This was due to low solids inventory (see also Figure 4-25).

Figure 4-21: MTF Temperature Profiles - Test Week 1

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

Figure 4-22: MTF Temperature Profiles - Test Week 2

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

Figure 4-23: MTF Temperature Profiles - Test Week 2 (continued)

The furnace pressure profiles for each of the defined test points are given in Figure 4-24 through Figure 4-26. Test Points C1 and C2 have very low furnace inventories, consistent with their non-uniform temperature profiles.

Note also that the pressure at 25 foot elevation (Point C in Figure 4-20) often reads low compared to the rest of the pressure profile, perhaps due to a bad sensor.

Figure 4-24: MTF Pressure Profiles - Test Week 1

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

Figure 4-25: MTF Pressure Profiles - Test Week 2

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

Figure 4-26: MTF Pressure Profiles - Test Week 2 (continued)

Average data from these temperature and pressure profiles are summarized in Table 4-9. The furnace temperatures are averaged at four elevations and the pressured drop is split into two sections. This summary data is plotted in Figure 4-27 and Figure 4-28.

Table 4-9: Summary of Temperature and Pressure Profiles

Test	Average Temperature, °F				Average Pressure Drop, in. w.g.		
	Bottom	Mid 1	Mid 2	Upper	Total	Lower	Upper
A1	1644	1664	1680	1693	34.2	14.3	19.8
B1	1663	1681	1707	1724	28.8	17.5	11.3
B2	1617	1650	1703	1731	24.3	15.0	9.3
B3	1641	1664	1690	1706	24.8	22.7	2.1
B4	1615	1646	1678	1698	22.4	13.7	8.7
B5	1656	1691	1724	1749	37.5	33.3	4.1
C1	1628	1704	1816	1695	3.2	2.0	1.2
C2	1592	1664	1682	1609	10.4	5.2	5.2
C3	1675	1709	1705	1645	22.1	12.8	9.3
C4	1655	1704	1714	1650	24.5	13.8	10.6
C5	1671	1708	1706	1626	28.6	18.9	9.7
C6	1670	1704	1708	1637	34.4	25.0	9.4
C7	1642	1683	1701	1630	39.9	33.1	6.7
D1	1637	1662	1655	1575	30.2	22.7	7.6
D2	1671	1702	1691	1609	41.9	32.9	9.1
D3	1651	1690	1694	1629	28.4	18.8	9.7
D4	1647	1692	1714	1653	42.5	33.9	8.6
D5	1640	1679	1685	1621	34.6	24.4	10.2
Key - refer to Figure 4-20 Bottom - average of bottom 6 temperatures at 3 elevations Mid 1 - average of three temperatures at location B Mid 2 - average of three temperatures at location C and next level up Upper - average of three temperatures at location D and next level down Total Pressure Drop - Point A to Point E Lower Pressure Drop - Point A to Point B Upper Pressure Drop - Point B to Point E

Figure 4-27: Summary of Temperature Profiles

Figure 4-28: Summary of Furnace Pressure Drop

The roof of the furnace, the cyclone inlet, and the cyclone body are water-cooled with one or two inches of refractory lining for wear protection. As shown in Figure 4-21, the gas temperature drops as it passes from the furnace through the cyclone. The temperature drop was obviously even greater during the second test week with the cooling coil in the upper furnace (Figure 4-22 and Figure 4-23).

Table 4-10 lists these heat duties and the resulting temperature of the solids leaving the sealpot.

Table 4-10: Heat Removal in Upper Furnace and Cyclone

Test Point	QFired ¹	QTop ²	QCoil ³	Sealpot temperature ⁴
	MMBtu/hr			°F
A1	3.96	1.03	-	1644
B1	4.07	1.02	-	1630
B2	3.81	1.00	-	1544
B3	2.24	0.77	-	1260
B4	4.78	2.16	-	1496
B5	3.94	1.43	-	1365
C1	4.24	0.75	1.39	1138
C2	5.69	0.90	2.05	1385
C3	6.57	0.94	2.61	1460
C4	7.87	0.96	2.83	1480
C5	7.57	0.94	2.93	1482
C6	7.56	0.95	2.76	1490
C7	7.57	0.93	2.38	1478
D1	6.28	0.86	2.32	1412
D2	7.08	0.89	2.56	1440
D3	7.92	0.96	2.74	1478
D4	7.92	0.97	2.74	1488
D5	7.62	0.93	2.69	1469
¹ Heat input with fuel ² Heat removed in water-cooled/refractory-lined roof, cyclone inlet, and cyclone ³ Heat removed by upper furnace coil ⁴ Temperature at sealpot outlet				

4.6 Solids Samples

A list of all the ash samples taken during the two test weeks is given in Table 4-11.

Table 4-11: List of Solids Samples Taken

Test	DateTime	Sample	Test	DateTime	Sample	Test	DateTime	Sample
A1	4/2/04 18:30	hv	B5	4/7/04 13:55	xo	D2	6/6/04 20:00	fb
A1	4/2/04 22:05	hv				D2	6/6/04 20:20	xo
A1	4/3/04 00:07	xo	pre	6/1/04 18:30	fill	D2	6/6/04 20:45	hv
B1	4/3/04 20:27	xo	C2	6/5/04 03:52	xo		6/6/04 22:41	xo
B1	4/3/04 20:30	hv	C2	6/5/04 04:45	hv		6/6/04 22:53	hv
B1	4/3/04 22:25	hv	C3	6/5/04 09:30	bd		6/6/04 23:05	xo
B1	4/3/04 22:22	xo	C3	6/5/04 10:00	hv	D3	6/7/04 03:20	bh
	4/4/04 04:50	hv	C3	6/5/04 10:15	xo	D3	6/7/04 03:23	hv
	4/4/04 04:58	xo	C5	6/6/04 08:00	xo	D3	6/7/04 03:26	fb
B2	4/4/04 07:08	xo	C5	6/6/04 08:10	hv	D3	6/7/04 03:39	xo
B2	4/4/04 07:10	hv	C6	6/6/04 09:45	hv	D4	6/7/04 05:00	fb
	4/6/04 07:05	xo	C7	6/6/04 10:15	fb	D4	6/7/04 05:00	hv
	4/6/04 07:10	hv	C7	6/6/04 10:41	xo	D4	6/7/04 05:00	bh
B3	4/6/04 15:00	xo	C7	6/6/04 10:55	hv	D5	6/7/04 07:30	bh
B3	4/6/04 15:05	hv	D1	6/6/04 14:30	hv	D5	6/7/04 10:45	xo
B4	4/7/04 05:50	hv	D1	6/6/04 14:30	xo	D5	6/7/04 10:55	fb
B4	4/7/04 06:05	xo	D1	6/6/04 14:30	fb	D5	6/7/04 11:00	hv
	4/7/04 08:45	bd	D1	6/6/04 15:50	hv	post	6/7/04 14:00	bh
B5	4/7/04 14:05	bd	D2	6/6/04 19:50	hv	post	6/9/04 12:00	ch
B5	4/7/04 13:45	hv	D2	6/6/04 20:00	fb	post	6/10/04 12:00	coil

The sample types given in Table 4-11 have the following key:

hv - hi volume pseudo-isokinetic sample taken at the cyclone outlet

xo - pseudo-isokinetic sample taken at the crossover duct (cyclone inlet)

bd - sample of bed drain material

bh - sample of baghouse fly ash

fb - sample drained from fluid bed heat exchanger

fill - sample of material used to charge the fluid bed heat exchanger before Week 2 testing

ch - sample from cyclone hopper at end of test after cyclone pluggage

coil - sample of deposit on upper cooling coil after it was removed from the furnace

The results of chemical analyses of selected solids samples are given in Table 4-12 with a key to the analyses in Table 4-13.

Table 4-12: Analyses of Solids Samples

Test Point	Date Time	Sample	Lab #	% Ca	% Mg	% Carbonate as CO ₂	% Total Carbon	% Total Sulfur	% Sulfite as S	% Active Lime as CaO	% CO ₂ as CaCO ₃	% CaSO ₃	% CaSO ₄	Remaining Ca as CaO	% Mg as MgO	% Unburned C	% Inerts (diff)	Ca:S mole ratio	% Recarb	% of CaO Active
A1	4/2/04 18:30	hv	4-0267-A	10.14	0.29	0.34	22.84	3.19		4.90	0.77		13.54	8.18	0.48	22.75	54.28	2.54	3.0	59.9
A1	4/2/04 22:05	hv	4-0268-A	13.35	0.33	0.48	18.61	4.19	0.38	6.46	1.09	1.41	16.19	10.74	0.55	18.48	51.54	2.55	3.3	60.2
B1	4/3/04 20:30	hv	4-0271-A	12.32	0.33	0.36	18.04	3.35		5.74	0.82		14.22	10.92	0.55	17.94	55.55	2.94	2.7	52.6
B1	4/3/04 22:25	hv	4-0272-A	10.68	0.30	0.38	21.32	3.03		4.84	0.86		12.87	9.16	0.50	21.22	55.40	2.82	3.2	52.8
	4/4/04 04:50	hv	4-0274-A	5.47	0.32	0.60	31.93	1.84	0.60	2.23	1.35	2.23	5.29	3.68	0.53	31.77	55.16	2.38	9.9	60.6
	4/6/04 07:10	hv	4-0276-A	8.22	0.34	0.25	15.06	2.71		3.11	0.57		11.51	6.44	0.56	14.99	65.93	2.43	2.8	48.3
B3	4/6/04 15:05	hv	4-0278-A				13.09	2.86												
B4	4/7/04 05:50	hv	4-0279-A	4.33	0.33	0.97	33.28	2.00	0.00	1.16	2.21	0.00	8.49	1.32	0.55	33.02	54.42	1.73	20.4	87.6
B5	4/7/04 13:45	hv	4-0289-A	7.83	0.35	0.96	23.94	2.40		2.99	2.18		10.19	5.53	0.57	23.68	57.84	2.61	11.1	54.0
A1	4/3/04 00:07	xo	4-0269-A	23.25	0.33	0.19	0.17	4.63		14.28	0.43		19.66	24.19	0.55	0.12	55.05	4.02	0.7	59.0
B1	4/3/04 20:27	xo	4-0270-A	22.29	0.35	0.12	0.18	4.59		11.19	0.27		19.49	23.01	0.58	0.15	56.50	3.88	0.5	48.6
B1	4/3/04 22:22	xo	4-0273-A	20.53	0.30	0.27	0.62	4.35		11.81	0.61		18.47	20.77	0.50	0.55	59.10	3.78	1.2	56.9
	4/6/04 07:05	xo	4-0275-A	25.72	0.41	0.17	0.55	4.62		15.21	0.39		19.62	27.69	0.68	0.50	51.12	4.45	0.6	54.9
B3	4/6/04 15:00	xo	4-0277-A	17.21	0.36	3.10	11.29	4.52		7.14	7.05		19.19	12.22	0.59	10.44	50.50	3.05	16.4	58.4
B4	4/7/04 06:05	xo	4-0280-A	25.45	0.39	4.39	2.45	5.59		12.18	9.98		23.73	20.24	0.65	1.25	44.14	3.64	15.7	60.2
B5	4/7/04 13:55	xo	4-0290-A	27.66	0.41	0.23	0.34	6.87		16.12	0.52		29.17	26.39	0.67	0.28	42.96	3.22	0.8	61.1
C3	6/5/04 09:30	bd	4-0314-A	27.01	0.33	0.98	2.22	8.30			2.23		35.24	22.03	0.55	1.95	38.00	2.60	3.3	
C3	6/5/04 10:14	xo	4-0316-A	28.14	0.36	0.39	1.09	8.52			0.89		36.18	23.98	0.60	0.98	37.38	2.64	1.3	
C5	6/6/04 08:00	xo	4-0317-A	28.24	0.35	0.34	1.13	8.47			0.77		35.96	24.27	0.58	1.04	37.38	2.67	1.1	
C7	6/6/04 10:41	xo	4-0320-A	28.03	0.36	0.30	1.30	8.26			0.68		35.07	24.39	0.60	1.22	38.04	2.71	1.0	
D3	6/7/04 03:39	xo	4-0325-A	30.7	0.29	0.66	2.11	11.48			1.50		48.74	22.04	0.48	1.93	25.31	2.14	2.0	
D5	6/7/04 10:45	xo	4-0327-A	33.75	0.27	0.76	1.66	12.3			1.73		52.22	24.74	0.45	1.45	19.40	2.19	2.0	
D5	6/7/04 10:55	fb	4-0328-A	32.09	0.28	0.64	1.48	11.37			1.46		48.28	24.20	0.46	1.31	24.30	2.26	1.8	
	6/7/04 14:00	bh	4-0330-A	19.57	0.31	1.44	14.90	7.57			3.27		32.14	12.31	0.51	14.51	37.26	2.07	6.7	
C3	6/5/04 10:00	hv	4-0315-A	7.78	0.28	0.56	25.82	2.83			1.27		12.02	5.22	0.46	25.67	55.36	2.20	6.5	
C5	6/6/04 08:10	hv	4-0318-A	9.51	0.27	1.34	21.05	3.40			3.05		14.44	5.65	0.45	20.68	55.73	2.24	12.8	
C6	6/6/04 09:45	hv	4-0319-A	6.31	0.29	1.22	29.80	2.22			2.77		9.43	3.39	0.48	29.47	54.46	2.27	17.6	
C7	6/6/04 10:55	hv	4-0321-A	6.78	0.26	1.17	30.65	2.19			2.66		9.30	4.17	0.43	30.33	53.11	2.48	15.7	
D1	6/6/04 15:50	hv	4-0322-A	13.09	0.21	1.89	36	7.2			4.30		30.57	3.31	0.35	35.48	25.98	1.45	13.1	
D2	6/6/04 20:45	hv	4-0323-A	17.93	0.20	1.74	30.58	9.09			3.96		38.60	6.97	0.33	30.11	20.04	1.58	8.8	
D3	6/7/04 03:23	hv	4-0324-A	18.93	0.19	1.52	30.87	9.91			3.46		42.08	7.22	0.32	30.46	16.48	1.53	7.3	
D4	6/7/04 05:00	hv	4-0326-A	17.96	0.17	2.58	36.18	9.45			5.87		40.12	5.31	0.28	35.48	12.94	1.52	13.1	
D5	6/7/04 11:00	hv	4-0329-A				28.94	9.26												
	6/9/04 12:00	ch	4-0568-A	32.93	0.28	2.26	2.89	10.80			5.14		45.86	24.31	0.46	2.27	21.96	2.44	6.2	
	6/10/04 12:00	coil	4-0569-A	25.84	0.30	10.56	3.33	8.91			24.02		37.83	7.12	0.50	0.45	30.09	2.32	37.2	

Table 4-13: Key for Solids Analyses

Test Point		
Date Time	When Sample Taken	
Sample	Type of sample	
Lab #	ALSTOM Lab ID #	
% Ca	Measured Calcium	
% Mg	Measured Magnesium	
% Carbonate as CO ₂	Measured CO ₂ released from Carbonate	
% Total Carbon	Measured Total Carbon	
% Total Sulfur	Measured Total Sulfur	
% Sulfite as S	Measured Sulfite	
% Active Lime as CaO	Measurement of "Lime Reactivity"	
% CO ₂ as CaCO ₃	Calculated Assuming all CO ₂ is as CaCO ₃	
% CaSO ₃	Calculated Assuming all Sulfite is as CaSO ₃	
% CaSO ₄	Calculated Assuming Remaining Sulfur is as CaSO ₄	
Remaining Ca as CaO	Calculated Assuming Remaining Calcium is as CaO	
% Mg as MgO	Calculated Assuming all Magnesium is as MgO	
% Unburned C	Calculated From Total Carbon minus CO ₂	
% Inerts (diff)	Calculated by Difference	
Ca:S mole ratio	Calculated From Total Calcium and Total Sulfur	
% Recarb	Calculated CaCO ₃ as % of Total Calcium (mole basis)	
% of CaO Active	Measured Active Lime as % of Calculated CaO	

4.7 Test Results and Analysis

4.7.1 Operability and Agglomeration

Although there were some ups and downs in the testing due to site and facility issues, there were few problems with the general operation of the pilot plant attributable to the oxygen firing.

There was no evidence of agglomeration or defluidization in the furnace at local oxygen enrichments of up to 70%.

4.7.2 Recarbonation

The solids samples from the pilot tests were analyzed for CO₂, which is assumed to have been present as CaCO₃. The amount of calcium carbonate as a percentage of the total

calcium in each sample is shown as % Recarbonation in Table 4-12. This is shown in Figure 4-29 for the fly ash samples and the circulating material (sampled at the crossover duct or the FBHE drain). The circulating material stays in the furnace generally above the calcination temperature; it has a low level of carbonate in the ash. The two high carbonate samples on April 6 and 7 were not taken at long-term high temperature conditions. Perhaps the earlier upset conditions at lower temperature resulted in residual carbonate in the circulating material.

The fly ash was sampled at a point where the flue gas had cooled to below the calcination temperature. Many of these samples show a high level of recarbonation. It should be noted that recarbonation of the fly ash is possible in air-fired boilers as well (see Section 4.1.3). Recarbonation levels of up to 10% have been seen.

Figure 4-29: Recarbonation of Solids Samples

Aside from the analysis of the recarbonation in the ash, there was some operational evidence. The sealpot in the pilot plant ran below the calcination temperature because of the cooled cyclone and dipleg (Table 4-10). When the sealpot was fluidized only with CO₂, it did not operate properly. The pressure drop in the dipleg built up, and at one point the furnace temperature increased to 1800°F as the flow of cooled circulating solids was interrupted. When the sealpot was fluidized with air only or with a mixture of air and CO₂, it operated well.

At the end of the pet coke testing, the solids in the cyclone hopper set up and stopped flowing due to recarbonation. This has been noticed in commercial pet coke units in areas where the circulating material is not kept moving steadily. There is enough combustion of unburned carbon to raise the CO₂ content to the point where recarbonation occurs.

A sample of the material from the cyclone hopper was analyzed (Table 4-12). This sample had 5.14 % CaCO₃ which is 6.2% of the total calcium as carbonate. This is higher than the 2% recarbonation for the earlier pet coke circulating material.

Thin deposits were found on the upper furnace cooling coil after the test (Figure 4-30). These deposits were exposed for an extended time to the high carbon dioxide gas at well

below the calcination temperature. The analysis of a sample of the deposit is also given in (Table 4-12). Not surprisingly, there is a high 37.2% recarbonation of the calcium in this deposit. The deposit did break off easily from the tube by hand.

Figure 4-30: Upper Coil with Deposit

4.7.3 Gaseous Emissions

Correcting for Excess Oxygen and Air Leakage

The gaseous pollutants SO_2 , CO , NO_x , and N_2O are measured as volume (or molar) concentration in a dried flue gas. Obviously the concentration depends upon any change in the volume of the flue gas due to excess combustion air or air in-leakage. It is common to normalize the measured concentration to a fixed level of excess air - that is, a fixed level of oxygen in the flue gas. It is typical in the U.S. to express the concentration as parts per million (ppm) at 3% oxygen in the flue gas on a dry basis. The conversion factor is based on the fact that the excess air contains 21% oxygen:

$$\text{ppm @ 3\% O}_2 = \text{ppm measured} * (21 - 3) / (21 - \% \text{O}_2 \text{ measured})$$

For example, if we measure 100 ppm CO at 5% O_2 in the flue gas, the value normalized to 3% O_2 is $100 * (18) / (21 - 5) = 112.5$ ppm. Other common bases are 6% O_2 and 15% O_2 - the latter used for gas turbines which operate with high excess air.

It is sometimes useful to relate the emission level to the energy content of the fuel - e.g., the pounds of pollutant emitted per million Btu of fuel heating value fired (lb/MMBtu). The conversion to this unit is a two step process:

1. normalize the concentration value to zero percent oxygen - stoichiometric combustion with no excess air,
2. convert to lb/MMBtu using the calculated volume of stoichiometric flue gas generated per MMBtu fired.

The stoichiometric flue gas per MMBtu can be calculated from the fuel analysis or standard values may be used.

The situation with oxygen firing is more complicated. Excess oxygen, which is in the flue gas, may have come from excess oxidant (pure oxygen or a mixture of O₂ and CO₂) or it may have come from air in-leakage. Since the two sources have different oxygen contents, there are two different normalizations needed. It is necessary to know how much of each source there is.

This was done in these tests by also measuring the nitrogen content of the flue gas using a gas chromatograph. The fuel burned in O₂/CO₂ with no excess oxidant and no air leakage will have a small expected nitrogen content from the fuel nitrogen. Any additional N₂ in the flue gas is assumed to come from air leakage. Knowing the air leakage and its oxygen content (21%) allows us to determine how much additional oxygen is in the flue gas from excess O₂/CO₂ oxidant.

Emissions Results

A summary of the gaseous emissions from the pilot testing is given in Table 4-14.

Table 4-14: Gaseous Emissions

	O ₂ %	N ₂ %	SO ₂ ppm	CO ppm	NO _x ppm	N ₂ O ppm	SO ₂ lb/MM ²	CO lb/MM	NO _x lb/MM	N ₂ O lb/MM	SC ³ %
A1	3.9	80	33	101	220	--	0.066	0.089	0.32	--	98
B1	3.2	8	310	120	103	--	0.58	0.10	0.14	--	82
B2	3.4	5	340	170	50	72	0.68	0.15	0.07	0.10	79
B3	7.9	15	92	143	72	49	0.17	0.12	0.096	0.062	95
B4	2.8	13	550	200	55	82	0.71	0.11	0.051	0.073	78
B5	4.3	13	265	137	63	58	0.36	0.081	0.061	0.054	89
C1	9.3	16.7	760	295 ¹	133	--	1.07	0.18	0.14	--	66
C2	4.2	16.5	460	750 ¹	60	--	0.52	0.37	0.051	--	84
C3	3.5	13	455	450 ¹	89	--	0.42	0.18	0.059	--	87
C4	3.4	11	560	440	91	93	0.44	0.15	0.051	0.050	86
C5	4.9	19	415	394	144	111	0.36	0.15	0.089	0.066	89
C6	3.9	17	960	446	93	78	0.79	0.16	0.055	0.044	75
C7	4.4	17.3	1090	600	83	88	0.91	0.22	0.05	0.050	71
D1	2.9	17.4	400	210	107	61	0.38	0.088	0.074	0.040	94.7
D2	3.1	17.4	59	119	178	38	0.049	0.044	0.107	0.022	99.3
D3	3.3	17.4	342	122	131	16	0.30	0.047	0.082	0.010	95.9
D4	2.9	17.4	513	138	114	5	0.44	0.052	0.071	0.003	93.8
D5	2.7	12	100	130	150	15	0.083	0.047	0.09	0.009	98.8
¹ Data was not logged. Values are taken from the stripchart.											
² MM = million Btu fired											
³ SC = sulfur capture											

Sulfur Emissions

The level of sulfur capture and the resultant emissions depend on many factors, including

- **fuel** - sulfur content, ash content, and fuel rank/reactivity.
- **sorbent** - feed rate (Ca:S ratio), reactivity, and size
- **furnace design and operating conditions** - especially temperature, solids inventory, and extent of "air" staging.

The SO₂ emissions seen in the pilot plant or in a commercial unit do respond quickly to changes in any of these parameters. The composition of the furnace solids inventory is also important; this changes much more slowly so it can take many hours to reach a new steady state point after a change.

None of the results from these short pilot tests can be assumed to quantitatively apply to long-term commercial operation. There are several useful observations from the results.

The high sulfur capture (98%) in the air-fired test A1 is not realistic for such a high temperature with bituminous coal and the rather unreactive Chemstone sorbent. This is based on prior experience with this sorbent in the MTF. It is not, however, unusual to see very good or very poor sulfur capture at the beginning of a test campaign depending on the startup material used. If an inert material is used, it takes time to accumulate a bed of active calcium. In this test, material from an earlier campaign with Pittsburgh #8 fuel was used. Because of the high sulfur and relatively low ash content of that fuel, the calcium oxide content of the startup material was high. This may partly explain the high initial sulfur capture.

The sulfur capture in the low-O₂ enrichment tests (B1 - B5) is more in line with what is expected. At a normal fluidized bed temperature of 1550°F, we expect close to 90% sulfur capture with the Chemstone limestone in the MTF. In test points B1, B2, and B4, the sulfur capture was about 80%.

In the part load test point B3, the sulfur capture was 95%. The low gas velocity (high gas residence time) and the high excess oxygen improve the capture.

Test point B5 achieved 89% sulfur capture. This improvement is probably due to the increased inventory in the furnace during this test (see Figure 4-28).

The second test week began with the same Tri-Star bituminous coal but with the more reactive Aragonite sorbent. The first test points C1 and C2 are neglected because of the low solids inventories. Test points C3, C4, and C5 have similar inventories to test points B1, B2, and B4, but they have better sulfur capture (86-89% vs. 78 - 82%). The improvement can be due to

- more reactive sorbent,
- lower temperature in the upper furnace because of the cooling coil,
- higher primary "air" fraction.

Test points C6 and C7 were at reduced primary flow. This typically causes an increase in

SO₂ emissions, as the lower furnace becomes a more reducing environment. This trend is shown in Figure 4-31. This is partly a transient effect as the reducing conditions cause some CaSO₄ to decompose, releasing some previously captured sulfur. The steady-state impact of "air" split is less pronounced.

The sulfur capture with the pet coke was very good - 94 to 99%. This is expected as the optimum temperature for sulfur capture is high for pet coke. We have seen the sulfur capture improve even as the temperature increases above 1700°F.

Figure 4-31: Sulfur Capture vs. Primary Flow for High Enrichment Tests

Several conclusions can be drawn:

- The sulfur capture with bituminous coal is lower than normal because of the high temperatures.
- The difference between the unreactive Chemstone and the reactive Aragonite is not as great as expected at lower temperatures. It may be that the Chemstone is improved by the high CO₂ calcination environment (Section 4.1.4). The Aragonite is inherently very reactive and may not have been improved much by the high CO₂ environment.
- There is very good sulfur capture with petroleum coke even at the high temperatures.

CO Emissions

Carbon monoxide emissions depend strongly on fuel type and on furnace temperature. A medium volatile coal like the Tri-Star would typically be expected to have a CO emission rate of less than 0.1 lb/MMBtu. Petroleum coke generally has lower CO emissions than coal.

These O₂-fired tests had higher CO emissions - see Figure 4-32 and Figure 4-33. This is

likely due to the high CO₂ content of the flue gas and its retarding effect on the oxidation rate of the CO.

Figure 4-32: CO Emissions vs. Primary Flow

Figure 4-33: CO Emissions vs. Mid1 Temperature

Figure 4-33 shows that the CO emissions from bituminous coal were lower in the first test week than the second test week for comparable Mid1 temperatures. The second test week had cooler temperatures in the upper furnace due to the cooling coil. The lower temperature results in higher CO.

It is often the case in commercial units that the CO emissions decrease across the cyclone, due to final mixing of the flue gas. Because the water-cooled MTF cyclone is much cooler than its commercial counterpart, the degree of afterburning is less.

NO_x Emissions

Typical NO_x emissions from air-fired tests in the MTF pilot plant are in the range of 0.7 to 1.5 lb/MMBtu fired. The air-fired test point A1 in this campaign had much a much higher NO_x emission of 0.32 lb/MMBtu. This may be partly due to the smaller furnace diameter, which allows more rapid mixing of the fuel and oxygen in the furnace (refer to Section 4.2.3).

Even with the reduced diameter, the NO_x emissions with O₂-firing were between 0.05 and 0.14 lb/MMBtu. Figure 4-34 and Figure 4-35 show perhaps slight trends for lower NO_x with lower temperature and lower primary flow.

In a larger pilot plant - and certainly in a commercial unit - the mixing will be poorer and the emissions lower.

Figure 4-34: NO_x Emissions vs. Primary Flow

Figure 4-35: NO_x Emissions vs. Average Bottom Temperature

N₂O Emissions

Nitrous oxide (N₂O) is a greenhouse gas, which is currently not regulated. Although N₂O is released in much smaller quantities than CO₂, it is a more potent greenhouse gas. N₂O has roughly 300 times the global warming potential (GWP) of an equal mass of CO₂.

N₂O emissions are strongly dependent on temperature. Pulverized coal furnaces usually have well less than 10 ppm N₂O (normalized to 3% O₂). Fluid bed combustors, which operate much cooler, typically have from 50 to 100 ppm (@3% O₂).

The N₂O analyzer was not operating during the air-fired test A1, but in previous MTF tests, ALSTOM has seen from 70 to 100 ppm @3% O₂ when the upper furnace temperature is about 1650°F. This range is equal to about 0.1 to 0.13 lb/MMBtu. The emissions were much lower in these oxygen-fired tests - see Figure 4-36. The temperature used in Figure 4-36 is the temperature "Mid 1" from Table 4-9.

Figure 4-36: N₂O Emissions vs. Mid1 Temperature

4.7.4 Combustion Efficiencies/Unburned Carbon (UBC) Emissions

The ash, which results from the combustion process usually, contains some unburned carbon. This unburned carbon represents a heat loss, expressed as Carbon Heat Loss (CHL) which is the heating value of the unburned carbon as a percentage of the heating value of the parent fuel. Typical values for commercial CFB's are in the range of 1 to 2% of the heat input lost as unburned carbon. High and low reactivity fuels can deviate significantly from this range. Besides fuel reactivity, furnace temperature, excess air level, and cyclone capture efficiency have a large impact on CHL.

The two main ash streams from the combustor are the bed drain and the fly ash, each of which contain some unburned carbon. The CHL in the fly ash is usually much larger than that in the bed drain. As seen from Table 4-12, the carbon content of the one bed drain sample was 1.95 % compared to 15 - 30% for the fly ash samples.

The CHL in the fly ash is calculated by

$$\text{CHL} = \text{UBC} * \text{flow} * 14500 / Q$$

where,

UBC = % unburned carbon in the fly ash

flow = fly ash flow rate, lb/hr

14500 = the heating value of the unburned carbon, Btu/lb

Q = fuel firing rate, Btu/hr

The fly ash flow rate was estimated by the High Volume sample readings (see Section 4.2.1). These values with the calculated heat loss are given in Table 4-15. The CHL from the fly ash is plotted against the global O₂ in Figure 4-37.

Table 4-15: Carbon Heat Loss in the Fly ash

Test Point	Date Time	Unburned C, %	fly ash flow, lb/hr	carbon flow, lb/hr	heat loss, MMBtu/hr	Q Fired, MMBtu/hr	CHL in fly ash, %
A1	4/2/04 18:30	22.75	42	9.6	0.14	3.96	3.5
A1	4/2/04 22:05	18.48	47	8.7	0.13	3.96	3.2
B1	4/3/04 20:30	17.94	80	14.4	0.21	4.07	5.1
B1	4/3/04 22:25	21.22	62	13.2	0.19	4.07	4.7
	4/4/04 04:50	31.77	25	7.9	0.12	3.8	3.0
	4/6/04 07:10	14.99	23	3.4	0.05	3.4	1.5
B3	4/6/04 15:05	--	6	--	--	--	--
B4	4/7/04 05:50	33.02	27	8.9	0.13	4.78	2.7
B5	4/7/04 13:45	23.68	43	10.2	0.15	3.94	3.7
C3	6/5/04 10:00	25.67	40	10.3	0.15	6.57	2.3
C5	6/6/04 08:10	20.68	61	12.6	0.18	7.57	2.4
C6	6/6/04 09:45	29.47	27	8.0	0.12	7.56	1.5
C7	6/6/04 10:55	30.33	40	12.1	0.18	7.57	2.3
D1	6/6/04 15:50	35.48	16	5.7	0.08	6.28	1.3
D2	6/6/04 20:45	30.11	11	3.3	0.05	7.08	0.7
D3	6/7/04 03:23	30.46	18	5.5	0.08	7.92	1.0
D4	6/7/04 05:00	35.48	13	4.6	0.07	7.92	0.8
D5	6/7/04 11:00	--	4	--	--	--	--

Figure 4-37: Carbon Heat Loss in the Fly ash

Several conclusions can be drawn:

- The heat loss with bituminous coal was comparable for air and oxygen firing in the low enrichment tests (21% and 30% global O₂).
- The heat loss with bituminous coal was lower during the high enrichment tests. This might be partly due to the higher local oxygen content at the bottom of the furnace. It is likely that the improved cyclone design (higher inlet and outlet velocities - see Section 4.2.2) was effective in reducing the carbon loss in the fly ash.
- The pet coke had a lower CHL than the bituminous coal in the high enrichment tests. This is consistent with previous experience.
- In general, the carbon heat loss in the fly ash is higher in the MTF than in comparable commercial operation. Thus the CHL from a commercial O₂-fired CFB boiler will likely be similar to that in air-fired boilers.

4.7.5 Heat Transfer

Heat transfer to the water/steam cycle takes place in three main places in the CFB boiler system:

- in the **furnace**, the waterwall and any additional surface,
- in the **External Heat Exchanger** (FBHE or Moving bed Heat Exchanger) in the primary recirculation loop,
- in the **convective pass** where the flue gas and fly ash are cooled.

Section 4.1.2 discusses the expected differences in heat transfer between air and oxygen firing.

The convective pass heat transfer is expected to be higher with oxygen firing, but the

pilot plant had no convective test section during these tests. The MTF's FBHE was fluidized with air during these tests; no significant difference in heat transfer is expected.

Heat transfer data in the furnace was generated from the single-tube test sections and in the furnace coil in Test Week 2. There was no air firing in Test Week 2, so there is no comparison with the oxygen-fired data for the furnace coil.

The single-tube test sections were installed for the entire test. The heat transfer data can be compared between the air-fired tests and the 21% O₂ tests during Test Week 1. These tests were run at the same firing rate and velocity (Figure 4-18 and Figure 4-19).

Heat transfer to the waterwall is a strong function of the local furnace temperature and the suspension density of solids. Therefore, heat transfer between air and oxygen firing should be compared where the local temperature and the suspension densities were the same. Table 4-16 has two such comparisons for each of the two test sections.

Table 4-16: Waterwall Heat Transfer Coefficients (Uo) - Air vs. O₂ Firing

	dP ₁	T ₁	Uo ₁	dP ₁	T ₁	Uo ₁
Air	5	1520	31-33	9	1600	35-37
21%	5	1520	32-34	7-8	1600	35-37
	dP ₂	T ₂	Uo ₂	dP ₂	T ₂	Uo ₂
Air	2.1	1560	24	2.5	1670	32-33
21%	2.2	1570	26-27	2.5	1670	30-32
dP: inches w.g. ; T: °F ; Uo: Btu/hr/ft ² /°F						

The lower test section (1) has a higher local suspension density - as indicated by the pressure drop in that section of the furnace.

There is no clear difference in the heat transfer to the test sections between air and oxygen firing.

4.8 Conclusions

The conclusions from the MTF pilot-scale testing are summarized here.

- Furnace Operability/Agglomeration: For the most part, the furnace operated well when firing with the O₂/CO₂ mixture rather than with air. There was no evidence of particle agglomeration or defluidization in the furnace at local oxygen enrichments of up to 70%.
- Recarbonation: The furnace was kept above the calcination temperature for these tests. Solids samples from the furnace showed only 1-2 % residual CaCO₃. Fly ash samples had cooled to below the calcination temperature; they had higher levels of CaCO₃, as expected.

The MTF sealpot operated below the calcination temperature; recarbonation occurred, upsetting the sealpot operation. This would not occur in a commercial unit, where the

sealpot temperature is as hot as, or hotter than, the furnace.

There was a pluggage in the cyclone due to recarbonation with the pet coke.

- Gaseous Emissions: The gaseous emissions compared to air firing were generally as expected. The sulfur emissions were higher with the bituminous coal, since the furnace was above the optimum temperature. For pet coke, the optimum temperature is higher, so the oxygen-fired emissions were very low.

The carbon monoxide emissions were higher with oxygen firing, as expected based on earlier bench-scale tests. This is likely due to the high CO₂ content of the flue gas, which hinders oxidation of the CO.

Earlier tests have shown that the NO_x emissions are lower with oxygen firing; this was also the case in these pilot tests. It is not clear to what extent this is due to reductions in thermal, prompt, and/or fuel NO_x.

Nitrous oxide emissions were lower than those from air firing at comparable conditions.

- Unburned Carbon Heat Loss: With oxygen firing, the carbon heat loss in the fly ash was lower than, or comparable to, the levels with air firing. The carbon loss was lower for pet coke than for bituminous coal.
- Heat Transfer: There was no significant difference in heat transfer to the furnace waterwall test sections between air and oxygen firing. This heat transfer is dominated by solids effects, which do not depend on the gas composition.

5 CASE STUDIES: PROCESS DESCRIPTIONS, EQUIPMENT, AND PLANT PERFORMANCE

This section of the report describes the processes used for each of the cases analyzed. The equipment used for these processes is also described. Additionally, the performance of each case is also presented in terms of the associated energy and material balances as well as various plant performance summary tables and graphs.

Brief descriptions of each case are presented below with more detailed descriptions provided separately later in Sections 5.1, 5.2, and 5.3.

Case-1: Air Fired CFB without CO₂ Capture (Base Case for Comparison). Conventional air-fired CFB without CO₂ capture using 1,800 psig / 1,000 °F / 1,000 °F, 3.0 in. Hga steam cycle.

Implication: Provides reference point for performance & economic analyses to the CO₂ capture cases Original Case-2 and Updated Case-2.

Note: Case-1 is presented in this report with the original results from the Phase I Topical Report (Marion, et al., 2003) in Section 5.1.

Original Case-2 and Updated Case-2: New Compact Oxygen-Fired CFB with CO₂ Capture, Purification, Compression and Liquefaction.

Same steam cycle as Case-1 and nearly identical thermal input but significantly smaller boiler island equipment than Case-1. Oxygen is provided from a Cryogenic Air Separation Unit (ASU) Plant. The CFB Boiler Island provides a concentrated CO₂ flue gas product stream to the Gas Processing System where it is further purified, compressed and liquefied.

Implication: Near term concept. Cost savings for the Boiler Island. Cost savings for the Gas Processing System equipment as compared to amine scrubbing systems. Improved plant thermal efficiency as compared to amine based CO₂ capture systems.

Note: Case-2 is presented in two forms in this report:

1. With the original results from the Phase I Topical Report (Marion, et al., 2003) (refer to Section 5.2 and Appendix II). This case is referred to as **Original Case-2** throughout this report
2. An updated version utilizing the pilot scale test results from Section 4 as a basis to update the design, performance, costs, and economics (Sections 5.3, 6, and 7 respectively). This case is referred to as **Updated Case-2** throughout this report

5.1 Case-1: Air Fired CFB without CO₂ Capture (Marion, et al., 2003)

This case, which was shown in the Phase I Topical Report (Marion, et al., 2003), represents the Base Case and was included in the study to provide a basis for comparison with the other CFB based options which included CO₂ removal (i.e., Original Case 2 and Updated Case-2). The Base Case (Case-1) for this study is defined as the selected unit firing coal at full load, utilizing air as the oxidant, without capturing CO₂ from the flue gas. This case utilizes a subcritical steam plant with reheat (1,800 psia / 1,000 °F / 1,000 °F). This represents the “business as usual” operating scenario and is used as the basis of comparison for the CO₂ removal option investigated in this study.

A brief performance summary for this plant reveals the following information. The Case-1 plant produces a net plant output of 193,034 kW. The net plant heat rate and thermal efficiency are calculated to be 9,611 Btu/kWh and 35.5 percent, respectively (HHV basis) for this case. Carbon dioxide emissions are about 2.00 lbm/kWh on a normalized basis.

Refer to the Phase I Topical Report (Marion, et al., 2003), for a detailed summary of plant performance and design, as this information is not repeated here. Refer to Section 6 and Appendix I for plant costs for Case-1. Economic results for this air fired case are provided in Section 7.

5.2 Original Case-2: Oxygen Fired CFB with CO₂ Capture

This case is referred to as Original Case-2 throughout this report. This case, from the Phase I Topical Report (Marion, et al., 2003), represents a power plant comprised of an oxygen-fired Circulating Fluidized Bed (CFB) boiler, a Moving Bed Heat Exchanger (MBHE), a cryogenic type Air Separation Unit (ASU), and a subcritical steam plant with reheat (1,800 psia / 1,000 °F / 1,000 °F). This previously analyzed case is cited in this report for comparison purposes only. It can be compared to Updated Case-2 shown in Section 5.3. The plant is designed to produce a flue gas having a high concentration of CO₂. This stream is then further processed in a Gas Processing System to produce a CO₂ product suitable for usage or sequestration.

A brief performance summary for this plant reveals the following information. The Original Case-2 plant produces a net plant output of about 134.5 MW. The net plant heat rate and thermal efficiency are calculated to be 13,546 Btu/kWh and 25.2 percent respectively (HHV basis) for this case. Carbon dioxide emissions are about 0.18 lbm/kWh on a normalized basis.

Refer to Appendix II for a detailed summary of plant performance and design for Original Case-2. Plant costs are provided in Section 6 and Appendix I. Economic results for this oxygen fired case are provided in Section 7.

5.3 Updated Case-2: Oxygen Fired CFB with CO₂ Capture

This case is referred to as Updated Case-2 throughout this report. This section provides a review of the changes made to update the oxygen fired CFB design and plant performance cited previously in Section 5.2. This update of Original Case-2 represents a power plant comprised of the following major equipment groups:

- An oxygen-fired Circulating Fluidized Bed (CFB) boiler including a Moving Bed type of external Heat Exchanger (MBHE)
- A cryogenic type Air Separation Unit (ASU)
- A subcritical steam plant with reheat (1,800 psia / 1,000 °F / 1,000 °F / 3.0 in. Hga)
- A Gas Processing System (GPS) designed to purify, compress, and liquefy the CO₂ product for sequestration.

The basic CO₂ capture concept behind Updated Case-2 is to replace combustion air with oxygen thereby creating a high CO₂ content flue gas stream that can be further processed into a high purity CO₂ end product for various uses or sequestration. The replacement of combustion air with high purity oxygen can, as an additional benefit, significantly reduce

the gas flow throughout the Boiler Island equipment. In this case a cryogenic Air Separation Unit (ASU) supplies the oxidant (99 percent pure oxygen) for the combustion of coal rather than direct utilization of ambient air as was done in Case-1. Since the size and cost of much of the equipment contained within the Boiler Island is strongly gas flow dependent (combustor, cyclones, backpass heat exchangers, air heater, fans, ductwork, baghouse, etc.) significant cost savings are anticipated for the Boiler Island of this concept as compared to the comparable air-fired CFB (Case-1). The plant design configuration reflects current information and design preferences, the availability of a current generation steam turbine, and the design latitude offered by a Greenfield site.

A brief performance summary for the Updated Case-2 plant reveals the following information. The Updated Case-2 plant produces a net plant output of about 138 MW. The net plant heat rate and thermal efficiency are calculated to be 13,152 Btu/kWh and 25.95 percent respectively (HHV basis) for this case. Carbon dioxide emissions are about 0.17 lbm/kWh on a normalized basis.

The update is based on information obtained from the pilot scale testing and data analysis discussed previously in Section 4. Specific changes implemented for Updated Case-2 relative to Original Case-2 are described in Section 5.3.1 shown next.

5.3.1 Specific Assumption Changes implemented for Updated Case-2

The following four subsections (Oxygen Fraction in Oxidant Stream to Furnace, Furnace Heat Transfer Rate, Low Level Heat Recovery System, and Recarbonation Issue) describe areas where key assumptions were made for the analysis of the oxygen fired Original Case-2 study (Marion, et al., 2003). The key assumptions can be categorized as either assumed process variable values or as assumed process equipment arrangements. These subsections discuss the updated values for these assumed process variables or the revised system arrangements used. Additionally, the rationale for the use of these updated process values or system arrangements is also discussed.

Oxygen Content in Oxidant Stream to Furnace

The oxygen fired Original Case-2 calculations assumed the local oxygen content in the oxidant stream for combustion (Stream 17 shown in Figure 5-2: refer to Section 5.3.2) to be about 70 percent by volume oxygen with the remainder as recirculated flue gas (primarily CO₂). The high oxygen content is important because it allows reduced gas flows throughout the boiler island (as compared to traditional air firing) and therefore is responsible for significant size and cost reductions for much of the boiler island equipment. A potential problem with high oxygen content for combustion in CFB boilers is the potential for bed agglomeration, which could cause a forced shut down of the unit. Theoretically, bed agglomeration should not occur in a CFB even with 100 percent oxygen content because the bed temperature is controlled by the continuous recirculation of a large quantity of cooled solids from the external heat exchanger.

During MTF testing (refer to Section 4) it was found that the high level of oxygen (70 percent by volume, locally) assumed for the oxygen fired Original Case-2 analysis was satisfactory, since no bed agglomeration was observed throughout the test campaign. Therefore this value was maintained for the oxygen fired Updated Case-2 as reported herein.

Furnace Heat Transfer Rate

The oxygen fired Original Case-2 calculations assumed, as a first approximation, that the furnace heat transfer rate would be identical to that of air firing based on the following rationale. The analysis of the furnace flue gas with oxygen firing has much higher CO₂ and H₂O contents which would tend to increase the non-luminous radiant component of the furnace heat transfer rate (gas to walls). However, it is also known that within the furnace of a CFB the heat transfer rate is clearly dominated by solids heat transfer phenomenon (conduction, convection and radiation). Therefore, for the oxygen fired Original Case-2 calculations, furnace heat transfer rates were assumed to be identical to that of air firing.

This assumption was confirmed later by heat transfer data analysis from the Phase II MTF testing (refer to Section 4.7.5) and therefore air firing furnace heat transfer rates were also used for Updated Case-2 as reported herein.

Low Level Heat Recovery System

The oxygen fired Original Case-2 calculations utilized a parallel feedwater heater (PFWH) heat exchanger to cool the flue gas stream leaving the oxygen heater and particulate removal system (Stream 7 shown in Figure 5-2: refer to Section 5.3.2). This flue gas stream was cooled by transferring heat to a low temperature feedwater stream in parallel to extraction feedwater heaters #1 and #2 (refer to Figure 5-7: refer to Section 5.3.5). About 25 percent of the total low-pressure feedwater flow leaving the condensate pump bypassed extraction feedwater heaters #1 and #2 and recovered heat in the PFWH in the original arrangement.

In Updated Case-2, the total low-pressure feedwater stream leaving the condensate pump bypasses extraction feedwater heaters #1 and #2. The additional heat added to the feedwater (in addition to that heat provided by flue gas from the PFWH) is provided by the recovery of low level heat previously rejected to cooling water by the Air Separation Unit (ASU) air compressor aftercooler. This results in an increase in steam turbine generator output of about 2.2 MWe and an increase in condenser heat rejection of about 67.3×10^6 Btu/hr or about 6.8 percent as compared to the Original Case-2 analysis.

The overall heat rejected by the cooling tower, which in addition to the condenser heat rejection also includes heat rejection from the ASU and Gas Processing System (GPS), is reduced by about 2.8 percent as compared to the Original Case-2 system.

Recarbonation Issue

The Original Case-2 calculations did not adequately consider potential recarbonation ($\text{CaO} + \text{CO}_2 \rightarrow \text{CaCO}_3 + \text{Heat}$) occurring in the seal pot area when it was fluidized with recirculated flue gas (primarily CO₂). Figure 5-1 shows the calcination temperature of calcium carbonate as a function of temperature and CO₂ partial pressure.

Figure 5-1: Calcination Temperature of Calcium Carbonate

The area to the left of the curve represents conditions where recarbonation will occur. With oxygen firing where CO₂ partial pressures are in the 0.6 – 0.7 atmosphere range, 1,600 °F is a temperature where recarbonation could occur. During the oxygen fired testing in the MTF (refer to Section 4) the seal pot was originally fluidized with CO₂. During these initial tests it was determined that recarbonation was occurring, which caused poor fluidization, leading to plugging and finally solids flow stoppage in this area. To correct the problem in the MTF, air was substituted as the fluidizing medium for the remainder of the tests.

For Updated Case-2, reported herein, several possible solutions to eliminate recarbonation for commercial sized oxygen fired units were considered, as listed below.

1. Fluidization of the seal pot etc. with air (similar to what was done for the MTF testing) including separate air supply, heat recovery, particulate removal, and venting systems.
2. Fluidization of the seal pot etc. with recycled flue gas and operation of the bed at temperatures well beyond those conducive to recarbonation (i.e., > 1675 F). With the higher bed temperature an SO₂ polishing system would be required for high levels of SO₂ removal. This would be required since sulfur capture in the bed at reasonable Ca/S ratios would be significantly lower than 90 percent, which was calculated to be obtained in the oxygen fired Original Case-2.
3. Fluidization of the seal pot etc. with recycled flue gas with the complete elimination of limestone usage in the bed and utilization of a tail end Flash Drier Absorption (FDA) system utilizing hydrated lime as a sorbent for SO₂ capture from the flue gas.

The disadvantages of Option 1 (fluidizing with air) include additional investment costs associated with the air supply, heat recovery, particulate removal, and venting systems. Additionally, it is expected that some of the air supplied for fluidization would occupy the void spaces of the solids recycle stream and be carried into the furnace. This would lower the purity of the CO₂ produced in the furnace and also adversely impact the performance and cost of the Gas Processing System (GPS).

Options 2 and 3 are quite similar. Option 2, operating the bed at higher temperature (>1675 °F) while fluidizing with recycled flue gas, would cause lower in bed sulfur capture with reasonable Ca/S ratios and therefore would require the addition of an SO₂ polishing system such as a flash dryer absorber (FDA). The higher bed temperature would cause a slight reduction in the cost of heating surface (superheater, reheater, evaporator, and economizer) however it would also reduce the margin for bed agglomeration.

An operating cost comparison between Options 2 and 3 was shown to slightly favor Option 3 because of a slight reduction in both sorbent cost and in solid waste disposal costs. Additional investment cost savings for Option 3 as compared to Option 2 would include elimination of the limestone handling and storage and feed system for the CFB. Taking into consideration of all the above advantages and disadvantages led us to believe Option 3 would be the best way to alleviate the recarbonation issue in a commercial unit and therefore this option was selected for Updated Case-2.

The design basis for Updated Case-2 is the same as was used for Case-1 and the Original Case-2 as shown in Section 2 of the Phase I Topical Report (Marion, et al., 2003) and is not repeated here.

5.3.2 Updated Case 2: Boiler Island Process Description and Equipment

The basic design concept for Updated Case-2 is to minimize the amount of gas recirculation such that the size and cost of many of the Boiler Island components, and other components, can be reduced significantly. By using oxygen firing in a CFB system with minimal gas recirculation, the flue gas flow to coal flow ratio is reduced. This ratio was reduced from about 10.8 lbm gas/lbm coal with air firing (Case-1) to about 4.2 lbm gas/lbm coal for Updated Case-2 with oxygen firing. This gas to coal ratio equates to about 70 percent oxygen by volume in the oxidant feed streams to the combustor. Since the size of many of the Boiler Island system components is strongly dependent on gas flow, significant size and cost savings are anticipated for these components.

This section describes the Boiler Island processes for Updated Case-2 and includes a simplified process flow diagram (PFD), material and energy balance and equipment description. The equipment description includes only the major components of the Boiler Island which have changed as a result of the update of this case.

Process Description and Process Flow Diagrams

Figure 5-2 shows a simplified process flow diagram for the Boiler Island of the Updated Case-2 oxygen-fired CFB concept. This process description briefly describes the function of the major equipment and systems included within the Boiler Island. Selected mass flow rates (lbm/hr) and temperatures (° F) are shown on this figure. Complete data for all state points are shown in the material and energy balance shown in Table 5-1. In this concept coal or another high carbon content fuel (Stream 1) is reacted with a preheated mixture of substantially pure oxygen and recirculated flue gas (Stream 18) in the Combustor section of the Circulating Fluidized Bed (CFB) system. The oxygen supply (Streams 16, 17, 18) is provided from a cryogenic Air Separation Unit (ASU).

The products of combustion leaving the combustor, flue gas comprised of primarily CO₂ and H₂O vapor and unreacted hot solids, flow through a cyclone, or another type of

particulate removal device, where most of the hot solids are removed. The hot solids are recirculated to the combustor through two parallel paths as described below. The temperature in the combustor is controlled to the appropriate level by properly splitting the flow of hot recirculated solids leaving the cyclone. They are split between an uncooled stream which flows directly back to the Combustor and a stream flowing through the External Heat Exchanger where the solids are cooled before returning to the Combustor. Exchanging heat with the power cycle working fluid in the External Heat Exchanger cools the hot solids.

Draining hot solids through water-cooled fluidized bed ash coolers (Stream 21) controls solids inventory in the system while recovering heat from the hot ash. The cooling water used for the ash coolers is provided from the feedwater stream leaving the final extraction feedwater heater of the steam cycle.

Figure 5-2: Updated Case-2: Simplified Boiler Island Gas Side Process Flow Diagram

The flue gas leaving the cyclone (Stream 3) is cooled in heat exchanger sections (Economizer) located in the convection pass (back pass) of the system, also by exchanging heat with the power cycle working fluid. The flue gas leaving the convection pass heat exchanger sections (Stream 5) is further cooled in an Oxygen Heater. The oxygen stream leaving the Air Separation Unit (Stream 16) is mixed with a small stream of recirculated flue gas (Stream 15) and the mixture is preheated in the Oxygen Heater. The quantity of recirculated flue gas used is only the amount necessary to provide proper fluidization for the bed and other equipment in the CFB system requiring a fluidizing medium.

The flue gas leaving the Oxygen Heater (Stream 6) is cleaned of fine particulate matter and SO₂ in the Particulate Removal and Flash Dryer Absorber (FDA) system. The flue gas is further cooled in a Parallel Feedwater Heater (PFWH) by transferring heat to feedwater stream in parallel with extraction feedwater heaters. Finally, a Gas Cooler is used to cool the gas before the flue gas enters the Induced Draft (ID) Fan (Stream 10).

The gas cooler is used to cool the flue gas to the lowest temperature possible before recycling to minimize the power requirements for the draft system (induced draft fan, fluidizing air blower, and gas recirculation fan) and the product gas compression system which is part of the Gas Processing System. Some H₂O vapor is condensed in the Gas Cooler. The flue gas leaving the ID Fan (Stream 11), comprised of mostly CO₂ and H₂O vapor, is split with most of the flue gas going to the product stream (Stream 12) for further processing and the remainder recirculated to the CFB system.

By using oxygen instead of air for combustion, and by minimizing the amount of recirculated flue gas, the size and cost of the Combustor, Cyclone, and Convection Pass Heat Exchanger sections and other equipment can be reduced as compared to many other concepts for CO₂ capture with CFB systems.

Material and Energy Balance

Table 5-1 shows the Boiler Island material and energy balance for Updated Case-2. The stream numbers shown at the top of each column of the table refer to stream numbers shown in the Updated Case-2 simplified PFD for the Boiler Island (Figure 5-2). The performance shown was calculated at MCR conditions for this unit and at ambient conditions as defined in the design basis.

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Table 5-1: Updated Case-2: Boiler Island Gas Side Material and Energy Balance

Constituent	(Units)	1	2	3	4	5	6	7	8	9	10	11	12
O2	(Lbm/hr)	5147		17632	3223	20855	20855	20855	20855		20855	20855	13572
N2	"	2378		14337	10677	25014	25014	25014	25014		25014	25014	16279
H2O	"	6499		65170	180	65351	65351	84351	84351	65159	19192	19192	12489
CO2	"			557768		557768	557768	557768	557768		557768	557768	362979
SO2	"			8025		8025	8025	1170	1170		1170	1170	762
H2	"	5815											
Carbon	"	101076		1011		1011	1011						
Sulfur	"	3812											
Ca	"	0											
CaO	"												
CaSO3	"												
CaSO4	"												
CaCO3	"			0									
Ash	"	38166	0	19083		19083	19083						
Total Gas	(Lbm/hr)	Coal	Limestone	Flue Gas	Infiltration Air	Flue Gas	Flue Gas	Flue Gas	Flue Gas	Condensate	Flue Gas	Flue Gas	Flue Gas
Total Solids	"	162894	0	20094		20094	20094				611854	611854	406080
Total Flow	"	162894	0	683026	14081	697106	697106	689158	689158	65159	623999	623999	406080
Temperature	(Deg F)	80	80	1600	80	587	305	175	135	100	100	112	112
Pressure	(Psia)	14.7	14.7	14.7	14.7	14.4	14.2	13.8	13.7	14.7	13.6	14.7	14.7
$h_{\text{sensible-gas}}$	(Btu/lbm)			438.718		128.656	53.828	22.743	13.000		4.229	6.828	6.828
$h_{\text{sensible-solids}}$				407.729		111.891	45.357						
Energy										19.960			
Chemical	(10 ⁶ Btu/hr)	1803.888		14.245		14.245	14.245						
Sensible	(10 ⁶ Btu/hr)	0.0	0.000	299.033		89.350	37.354	15.674	8.959	1.301	2.587	4.178	2.773
Latent	(10 ⁶ Btu/hr)	0.0	0.000	68.429	0.189	68.618	68.618	88.568	88.568	0.000	20.151	20.151	13.114
Total Energy ⁽¹⁾	(10 ⁶ Btu/hr)	1803.888	0.000	381.707	0.189	172.213	120.216	104.242	97.527	1.301	22.739	24.329	15.887

Constituent	(Units)	13	14	15	16	17	18	19	20	21	22	23	24
O2	(Lbm/hr)	7283	6035	6035	319084	325119	325119	1248	1248				
N2	"	8736	7238	7238	3223	10461	10461	1497	1497				
H2O	"	6702	5553	5553		5553	5553	1149	1149			19963	963
CO2	"	194789	161399	161399		161399	161399	33389	33389				
SO2	"	409	339	339		339	339	70	70				
H2	"												
Carbon	"									1011	1011		1011
Sulfur	"												
Ca	"												
CaO	"											7801	1800
CaSO3	"												12855
CaSO4	"												
CaCO3	"												
Ash	"									19083	19083		19083
Total Gas	(Lbm/hr)	Recirc Gas	Recirc Gas	Recirc Gas	Oxygen	Oxy + Recirc	Oxy + Recirc	Grease Gas	Grease Gas	Hot Ash Drain	Cool Ash Drain	Hydrated Lime	CaSO4 + CaO
Total Solids	"	217919	180564	180564	322307	502871	502871	37354	37354	20094	20094	7801	34749
Total Flow	"	217919	180564	180564	322307	502871	502871	37354	37354	20094	20094	27764	35713
Temperature	(Deg F)	112	112	140	65	91	537	112	195	1600	520	80	175
Pressure	(Psia)	14.7	14.7	17.4	17.4	17.4	17.19	14.7	23.7	14.7	14.7	14.7	14.7
$h_{\text{sensible-gas}}$	(Btu/lbm)	6.828	6.828	12.941	-3.361	2.492	105.891	6.828					
$h_{\text{sensible-solids}}$									25.019	407.729	95.391	0.000	23.874
Energy										14.245	14.245		14.245
Chemical	(10 ⁶ Btu/hr)												
Sensible	(10 ⁶ Btu/hr)	1.488	1.233	2.337	-1.083	1.253	53.250	0.255	0.935	8.193	1.917	0.000	0.853
Latent	(10 ⁶ Btu/hr)	7.037	5.831	5.831	0.000	5.831	5.831	1.206	1.206	0.000	0.000	0.000	1.011
Total Energy ⁽¹⁾	(10 ⁶ Btu/hr)	8.525	7.064	8.168	-1.083	7.084	59.081	1.461	2.141	22.437	16.161	0.000	16.108

Notes:—
(1) Energy Basis; Chemical based on Higher Heating Value (HHV); Sensible energy above 80F; Latent based on 1050 Btu/Lbm of water vapor

The MCR condition is defined as high-pressure turbine inlet conditions of 1,400,555 lbm/hr, 1,815 psia, and 1,000 °F and intermediate-pressure turbine inlet conditions of 1,305,632 lbm/hr 469 psia 1,000 °F. These conditions were also used for the Base Case (Case-1) and the Original Case-2. The boiler was fired with enough oxygen to leave about 3 percent by volume of oxygen in Stream 3, the same as for the Base Case. This oxygen requirement results in a stoichiometry of about 1.05 for Updated Case-2. The resulting boiler efficiency calculated for Updated Case-2 was 93.07 percent (HHV basis) with an oxygen heater exit gas temperature of 305 °F and the PFWH exit gas temperature of 135 °F.

Figure 5-3 shows a general comparison of the boiler heat absorption distribution between the air firing of Case-1 and the oxygen firing of Updated Case-2. Excluding the parallel feedwater heater (PFWH) heat absorption of Updated Case-2, the total heat absorption is the same in both Case-1 and oxygen fired Updated Case-2. The primary differences occur in the Convection Pass, Combustor and External Heat Exchanger. The Convection Pass heat absorption for Updated Case-2 is about 41 percent of the Case-1 value due to the reduced flue gas flow. Similarly, the Combustor heat absorption for Updated Case-2 is about 33 percent of the Case-1 value due to its significant reduction in size. Finally, the External Heat Exchanger heat absorption for Updated Case-2 is about 3.6 times greater than the Case-1 value. Because of the very large heat duty of the external heat exchanger of Updated Case-2 (about 68 percent of the total duty as compared to about 20 percent for Case-1) a moving bed design was selected based on cost considerations. This moving bed type of external heat exchanger is currently under development by ALSTOM (Jukkola, et al., 2003). The Case-1 external heat exchanger was a traditional bubbling fluidized bed design.

Figure 5-3: CFB Boiler Heat Absorption Comparison (Air and O₂ Firing)

The moving bed external heat exchanger design that was implemented in this case provides several advantages over a bubbling fluidized bed. One significant advantage of the moving bed is that a higher temperature differential is obtained between the bed material and the steam cycle working fluid. This reduces the surface area and weight requirements for the heat exchanger pressure parts. The higher temperature differential occurs because the moving bed can be designed as a counterflow heat exchanger. The bubbling fluid bed on the other hand is more of a “stirred” heat exchanger where the bed material is at a “stirred temperature”. The “stirred temperature” is much lower than the inlet solids temperature in the moving bed. Additionally, the moving bed does not require any fluidizing air, fluidizing blower, fluidizing nozzles, and fluidizing air piping thus providing a much simpler system. With these advantages, the moving bed allows for a much more compact and less expensive design than a bubbling bed design would in this application.

Boiler Island Equipment

The Boiler Island equipment description for Updated Case-2 is nearly identical to what was described in Section 2 of the Phase I Topical Report (Marion, et al., 2003) and is not repeated here except for any significant changes for Updated Case-2 as compared to Original Case-2.

The only significant difference in boiler island equipment for Updated Case-2 was in the system utilized for SO₂ removal. In the Original Case-2 it was assumed that traditional limestone injection into the furnace would remove about 90% of the SO₂ produced. Subsequent testing and data analysis indicated a problem with recarbonation (refer to Figure 5-1 and subsection 4.1.3, which describes the recarbonation issue). Therefore the Updated Case-2 does not use limestone injection into the furnace for SO₂ capture but instead uses a tail end Flash Dryer Absorber (FDA) system with hydrated lime injection for SO₂ capture. Figure 5-4 shows a schematic diagram of the system.

Figure 5-4: Flash Dryer Absorber System Schematic Diagram

Flue gas leaving the oxygen heater, with a high SO₂ content enters a spray dryer section prior to entering the fabric filter. Here, a mixture of recirculated ash, fresh lime and water are injected into the flue gas stream and most of the SO₂ reacts with the lime to form CaSO₃ x 1/2H₂O. The particulate is collected in the fabric filter. A portion of the collected particulate is removed as the waste product stream with the remainder of the particulate being recirculated. Water is added to control the humidity of the flue gas stream leaving the fabric filter to a proper level. Fresh lime is also added. FDA systems are commercial products for both air-fired CFB units and pulverized coal fired units.

Because in this oxygen-fired application there is a high CO₂ content in the flue gas, the performance of the FDA system is somewhat of an unknown at this time and therefore various assumptions were made. As a follow-on to the Phase-II work described in this report, a separate but related project (Commercialization Development of Oxygen Fired CFB for Greenhouse Gas Control, DOE/NETL Cooperative Agreement DE-FC26-D4NT42205) will address the FDA performance issue. This work will include further MTF testing which will include the FDA concept used for SO₂ removal described above. As a part of this effort, an updated commercial design will be developed incorporating the lessons learned during the testing work scope.

Because the assumption changes utilized for the Updated Case-2 (refer to Section 5.3.1) were relatively minor with respect to the boiler design as compared to Original Case-2, the general arrangement drawings shown for Original Case-2 and Updated Case-2 are nearly identical. The only significant change required to these drawings for Updated Case-2 would be the removal of the limestone injection system including the limestone silo and the addition of the FDA system integrated with the baghouse particulate removal system. Figure 5-5 and Figure 5-6 show general arrangement drawings of the Original Case-2 CFB boiler. The plan area for both the Original Case-2 and Updated Case-2 Boiler Island is about 52 percent of that for Case-1. Similarly, the building volume is about 56 percent of that for Case-1 and the total boiler weight is about 65% of the Case-1 weight.

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Figure 5-5: Updated Case-2: Boiler Island General Arrangement Drawing – Side Elevation

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Figure 5-6: Updated Case-2: Boiler Island General Arrangement Drawing - Plan View

5.3.3 Updated Case-2: Gas Processing System Process Description and Equipment

The Gas Processing System process description and equipment description for this Updated Case-2 is nearly identical to what was described in Section 2 of the Phase I Topical Report (Marion, et al., 2003) for Case-2 and is not repeated here (Refer to Appendix II for this information). The only change is that the Updated Case-2 processes about 3 percent less gas than Original Case-2. This is primarily due to the fact that Original Case-2 used limestone injection into the Combustor for SO₂ removal. When the limestone is calcined in the combustor it liberates CO₂. Since Updated Case-2 does not use injection of limestone in the furnace, about 3 percent less flue gas is generated in this case.

5.3.4 Updated Case-2: Air Separation Unit Process Description and Equipment

The Air Separation Unit process description and equipment description for Updated Case-2 is nearly identical to what was used in Section 2 of the Phase I Topical Report (Marion, et al., 2003) for Original Case-2 and is not repeated here (Refer to Appendix II for this information). The only change is that for Updated Case-2, some of the low-level heat previously rejected to cooling water from the air compressor aftercooler was recovered by the steam cycle with low temperature feedwater.

5.3.5 Updated Case-2: Balance of Plant Equipment and Performance

The balance of plant equipment and performance description for Updated Case-2 is nearly identical to what was used in Section 2 of the Phase I Topical Report (Marion, et al., 2003) for Original Case-2 and is not repeated here (Refer to Appendix II for this information). Only areas where there are differences are shown.

Steam Cycle Performance

The steam cycle was changed somewhat for Updated Case-2 with additional low level heat recovery. The steam cycle, shown schematically in Figure 5-7, represents a primary common thread among the cases. It is nearly identical for Case-1, Original Case-2 and Updated Case-2 differing only in the arrangement of the low level heat recovery systems. The steam turbine is a nominal 210 MWe single reheat machine with steam conditions of 1,800 psig 1,000 °F / 1,000 °F and a condenser pressure of 3.0 in. Hga. The main steam flows (1,400,555 lbm/hr) and reheat steam flows (1,305,632 lbm/hr) are identical for all cases. Six extraction feedwater heaters are used for Case-1 and Original Case-2. Updated Case-2 does not use extraction feedwater heaters #1 and #2 and therefore only four extraction feedwater heaters are used in this case. The final feedwater temperature is 470 °F for all cases.

Note:
Extraction Feedwater heaters #1 and #2 are not used for Updated Case-2.

Steam Cycle Energy Balance			
Energy Outputs		Energy Inputs	
	(10 ⁶ Btu/hr)		(10 ⁶ Btu/hr)
Steam Turbine Power Output	736	Boiler Heat Input	1777
Process Steam Heat Loss	0	BFP & CP Input	12
Condenser Loss	1054	Total Energy Input	1790
Total Energy Output	1790	In - Out	0
Turbine Heat Rate		8529.1 (Btu/kwhr)	

Figure 5-7: Updated Case-2 Steam Cycle Schematic and Performance

The steam cycle starts at the condenser hot well, which is a receptacle for the condensed steam from the exhaust of the steam turbine. The condensate flows to the suction of the condensate pumps (CP), which increase the pressure of the fluid by a nominal 250-psi to transport it through the piping system and enable it to enter the open contact heater, or deaerator. The condensate passes through a gland steam condenser (SPE) first, followed in series by two low-pressure extraction feedwater heaters (Case-1 and Original Case-2 only). The heaters successively increase the condensate temperature to a nominal 221°F by condensing and partially sub-cooling steam extracted from the LP steam turbine section. Each heater receives a separate extraction steam stream at successively higher pressure and temperature. The condensed steam (now referred to as heater drains) is progressively passed to the next lower pressure heater, with the drains from the lowest heater draining to the condenser.

The Updated Case-2 condensate and feedwater system differs from Case 1 in that there are two additional heat exchangers in parallel feedwater streams with the two low-pressure extraction feedwater heaters as shown in Figure 5-7. The two heat exchangers (shown schematically as a single component labeled “PFWH & ASU Clr” in the lower left corner of Figure 5-7) include a parallel low-pressure Feedwater Heater (PFWH), used to cool the flue gas leaving the baghouse, and a second heat exchanger (ASU Clr) used to recover some of the heat rejected by the ASU air compressor aftercoolers. These two heat exchangers increase the generator output by 3,055 kW or about 1.5 percent as

compared to Case-1. The condenser heat rejection is increased by about 94.0×10^6 Btu/hr or 9.8 percent.

The Updated Case-2 condensate and feedwater system differs from Original Case 2 in that there is a second heat exchanger (ASU Clr) in addition to the parallel low-pressure Feedwater Heater (PFWH) in parallel with extraction feedwater heaters #1 and #2. This additional heat exchanger is used to recover some of the heat rejected by the ASU air compressor aftercoolers. This heat exchanger recovers some of the heat previously rejected from the air compressor aftercooler. The air compressor aftercooler previously rejected all its heat to the cooling tower. The reduced extractions increase the generator output by 2,190 kW or about 1.0 percent as compared to Original Case-2. The condenser heat rejection is increased by about 6.8 percent.

The overall heat rejected by the cooling tower, which in addition to the condenser heat rejection also includes heat rejection from the ASU and Gas Processing System (GPS), is reduced by about 2.8 percent as compared to the Original Case-2 system.

The condensate streams entering the deaerator are heated and stripped of non-condensable gases by contact with the extracted steam flow entering the unit. The extracted steam is condensed and mixes with the heated condensate, which flows by gravity to a deaerator storage tank. The boiler feedwater pumps (BFP) take suction from the storage tank and increase the fluid pressure to a nominal 2,200 psig. Both the condensate pump and boiler feed pump are electric motor driven pumps. The boosted condensate leaving the BFP flows through three more high-pressure feedwater heaters, increasing in temperature to 470°F at the entrance to the boiler economizer section. Each heater receives a separate extraction steam stream at successively higher pressure and temperature. The condensed steam leaving the feedwater heaters (called drains) is progressively passed to the next lower pressure heater, with the drains from the lowest heater (heater #4) draining to the deaerator.

Within the CFB boiler the feedwater leaving the feedwater system is evaporated and finally superheated. The high-pressure superheated steam leaving the finishing superheater (1,400,555 lbm/hr of steam at 1,815 psia and 1,000 °F) is expanded through the high-pressure turbine. Reheat steam (1,305,632 lbm/hr) is heated and returned to the intermediate pressure turbine at 469 psia and 1,000 °F. These conditions (temperatures, pressures) represent common steam cycle operating conditions for existing utility scale CFB power generation systems in use today. The reheated steam expands through the intermediate and low-pressure turbines before exhausting to the condenser. The condenser pressure used for all cases in this study was 3.0 inches of mercury absolute (in Hga).

The steam turbine performance analysis results for Updated Case-2 show the generator produces 212,097 kW output and the steam turbine heat rate is about 8,259 Btu/kWh. The generator output, turbine heat rate and condenser losses are slightly higher for Updated Case-2 than for Case-1. This is a result of the PFWH and ASU Clr, which eliminate extraction flows to the low-pressure extraction feedwater heaters and increases LP turbine power output. The generator output, turbine heat rate and condenser losses are also slightly higher for Updated Case-2 than for Original Case-2. This is a result of the addition of the ASU Clr (not used for Original Case-2), which eliminates extraction flows

to the low-pressure extraction feedwater heaters and increases LP turbine power output.

Steam Cycle Equipment

The steam cycle equipment for Updated Case-2 is nearly identical to what was used in Section 2 of the Phase I Topical Report (Marion, et al., 2003) for Original Case-2 and the equipment description is not repeated here (Refer to Appendix II).

Other Balance of Plant Equipment

The other balance of plant equipment for Updated Case-2 is nearly identical to what was used in Section 2 of the Phase I Topical Report (Marion, et al., 2003) for Original Case-2 and is not repeated here (Refer to Appendix II).

5.4 Overall Plant Performance and CO₂ Emissions Comparison

This section provides a summary and comparison of several important plant performance outputs from this study. Comparison of Updated Case-2 to the Original Case-2 and Case-1 are provided.

Table 5-2 shows a fairly detailed comparison of plant performance and CO₂ emissions for both the CO₂ recovery concepts (Original Case 2 and Updated Case-2) and the Base Case (Case-1) that employs no CO₂ recovery system.

Additionally, selected results from this table are illustrated and compared in Figure 5-8 - Figure 5-16. The comparisons shown in the figures are Boiler Efficiency, Coal Heat Input, Steam Cycle Efficiency, Gas Processing System Auxiliary Power, Total Plant Auxiliary Power, Net Plant Output, Plant Thermal Efficiency, and Plant CO₂ Emissions.

Table 5-2: Plant Performance and CO₂ Emissions Summary and Comparison

		Phase I CFB Air Fired (Case 1)	Phase I CFB Cryogenic O ₂ Fired (Original Case 2)	Phase II CFB Cryogenic O ₂ Fired (Updated Case 2)
Auxiliary Power Listing				
	(Units)			
Induced Draft Fan	(kW)	2285	511	500
Primary Air Fan	(kW)	2427	n/a	n/a
Secondary Air Fan	(kW)	1142	n/a	n/a
Fluidizing Air Blower	(kW)	920	209	210
Transport Air Fan	(kW)	n/a	n/a	n/a
Gas Recirculation Fan	(kW)	n/a	341	340
Coal Handling, Preparation, and Feed	(kW)	300	292	292
Limestone Handling and Feed	(kW)	200	195	0
Limestone Blower	(kW)	150	146	0
Ash Handling	(kW)	200	195	60
Particulate Removal System Auxiliary Power (baghouse)	(kW)	400	151	153
Boiler Feed Pump	(kW)	3715	3715	3715
Condensate Pump	(kW)	79	79	79
Circulating Water Pump	(kW)	1400	1877	1825
Cooling Tower Fans	(kW)	1400	1877	1825
Steam Turbine Auxiliaries	(kW)	200	206	220
Misc. Auxiliary Power (Controls, Lighting, HVAC etc.)	(kW)	719	719	719
Transformer Loss	(kW)	470	472	477
Subtotal	(kW)	16007	10983	10413
	(frac. of Gen. Output)	0.077	0.052	0.049
Auxiliary Power Summary				
Traditional Power Plant Auxiliary Power	(kW)	16007	10983	10413
Air Separation Unit or Fuel Compressor	(kW)	n/a	37505	37242
CO ₂ Removal System Auxiliary Power	(kW)	n/a	26905	26039
Total Auxiliary Power	(kW)	16007	75393	73695
	(frac. of Gen. Output)	0.077	0.359	0.347
Output and Efficiency				
Main Steam Flow	(lbm/hr)	1400555	1400555	1400555
Steam Turbine Heat Rate	(Btu/kwhr)	8147	8256	8529
Steam Turbine Generator Output	(kW)	209041	209907	212097
Net Plant Output	(kW)	193034	134514	138402
	(frac. of Case-1 Net Output)	1.00	0.70	0.72
Boiler Efficiency (HHV) ¹	(fraction)	0.8946	0.9412	0.9307
Coal Heat Input (HHV)	(10 ⁶ Btu/hr)	1855	1806	1804
Natural Gas Heat Input (HHV) ²	(10 ⁶ Btu/hr)	n/a	16.5	16.4
Total Fuel Heat Input (HHV)	(10 ⁶ Btu/hr)	1855	1822	1820
¹ Boiler Heat Output / (Q _{coal} -HHV + Q _{credits})				
² Required for GPS & ASU Desiccant Regeneration in Case 2				
Net Plant Heat Rate (HHV)	(Btu/kwhr)	9611	13546	13152
Net Plant Thermal Efficiency (HHV)	(fraction)	0.3551	0.2520	0.2595
Normalized Thermal Efficiency (HHV; Relative to Base Case)	(fraction)	1.00	0.71	0.73
Energy Penalty	(fraction)	0.00	0.29	0.27
CO₂ Emissions				
CO ₂ Produced	(lbm/hr)	385427	376995	364865
CO ₂ Captured	(lbm/hr)	0	352377	341040
Fraction of CO ₂ Captured	(fraction)	0.000	0.935	0.935
CO ₂ Emitted	(lbm/hr)	385427	24618	23825
Specific CO ₂ Emissions	(lbm/kwhr)	2.00	0.18	0.17
Normalized Specific CO ₂ Emissions (Relative to Base Case)	(fraction)	1.00	0.09	0.09
Avoided CO ₂ Emissions (as compared to Base Case)	(lbm/kwhr)	0.00	1.81	1.82

Boiler Efficiency:

Figure 5-8 compares boiler efficiencies among the three cases. The boiler efficiency values shown include credit for the heat recovered by the PFWH's used in Original Case-2 and Updated Case-2. Case-1, the air-fired Base Case, is lower than the oxygen fired cases primarily due to a higher dry gas loss. The higher dry gas loss is the result of higher gas flow and temperature exiting the Boiler Island. In the oxygen-fired cases the exiting flue gas flow rates from the Boiler Island are much lower than for Case-1 (about 23% of the Case-1 flow) resulting in the reduced dry gas losses for these cases. These cases also have their flue gas streams cooled to lower temperatures (via the PFWH) than Case-1 also contributing to the reduction in dry gas loss. The flue gas flow rates are much lower for these cases since nearly all of the nitrogen is removed in the ASU system. Boiler efficiency improvement is about 4 percent.

Figure 5-8: Boiler Efficiency Comparison

The boiler efficiency for Updated Case-2 is slightly lower than Original Case-2 primarily because of the use of the FDA sulfur removal system. In the FDA system additional water is added to the flue gas stream for humidity control. This water is evaporated and therefore reduces the heat recovery in the PFWH which follows this system.

Coal Heat Input:

Figure 5-9 compares coal heat input to the boilers for the three cases. Case-1, the air-fired Base Case, has a slightly higher coal heat input (by about 2.8 percent) than the two oxygen fired cases primarily due to the fact that the boiler efficiency is slightly lower for this case as explained above.

The boiler heat output is nearly the same for all cases since all the cases use a steam cycle that is nearly identical. The only differences in the steam cycles are in the low level heat recovery systems as explained previously.

Figure 5-9: Coal Heat Input Comparison

Steam Cycle Efficiency:

Figure 5-10 compares steam cycle efficiency for the cases. Case-1, the air-fired Base Case, has a slightly higher steam cycle efficiency than the oxygen fired cases. This is primarily due to the fact that in Case-1 there is no low level heat recovery by feedwater streams in parallel with the extraction feedwater heaters. The low level heat recovery system used in Original Case-2 and Updated Case-2 uses feedwater heating in parallel with the traditional low-pressure extraction feedwater heaters (Heaters #1 and #2). Updated Case-2 has a lower efficiency than Original Case-2 because additional low-level heat recovery (from the ASU aftercoolers) was utilized in this case.

Figure 5-10: Steam Cycle Efficiency Comparison

Gas Processing System Auxiliary Power:

The CO₂ capture cases require CO₂ compression, purification and liquefaction within the Gas Processing Systems (GPS) in an effort to meet the product gas specification. The GPS power requirements were calculated to be about 153 kWh/ton of CO₂ captured for both these cases as shown in Figure 5-11.

Figure 5-11: Gas Processing System Normalized Auxiliary Power Comparison

Total Plant Auxiliary Power:

Figure 5-12 compares total plant auxiliary power among the cases. There are three main categories that comprise the total plant auxiliary power. These are:

1. The Gas Processing System (discussed above)
2. The Air Separation Unit (ASU)
3. The traditional power plant auxiliaries associated with the draft system, cooling water system, material handling, etc.

Figure 5-12: Auxiliary Power Comparison between Air-Fired and Oxy-fuel Fired CFB Plants

Case-1, the air-fired Base Case without CO₂ recovery, requires much less auxiliary power than the other cases, since it does not require an ASU or a Gas Processing System to compress the CO₂. The auxiliary power for this case is only that which is attributable to the traditional power plant equipment. This includes equipment for solids handling (coal, limestone, and ash), air and gas handling, water pumping for the steam cycle and cooling water systems, as well as other miscellaneous systems within the traditional power plant. This case requires slightly less than 8 percent of the generator output for auxiliary power.

A detailed listing of plant auxiliary power was shown in Table 5-2.

Original Case-2 and Updated Case-2 both include the ASU and GPS which consume about 18% and 13% of the gross output, respectively, while the traditional auxiliary power consumption is reduced to about 5% of the generator output (see Table 5-2).

Net Plant Power Output:

Figure 5-13 compares the resulting net power output (MWe) among these three cases. The increase in net power output for Updated Case-2 over the Original Case-2 of about 3.9 MWe is due to additional low-level heat recovery from the ASU compressor aftercoolers resulting in increased steam turbine output. Additionally, the GPS consumes about 3% less power since less gas is processed in Updated Case-2.

Figure 5-13: Net Plant Output Comparison

Plant Thermal Efficiency:

Figure 5-14 shows a comparison of Plant Thermal Efficiency between Case 1 and Original Case 2 and Updated Case-2. These efficiency results reflect the combined impact of boiler efficiency, steam cycle efficiency, and plant auxiliary power on net plant thermal efficiency. As shown previously, the differences in plant auxiliary power represents the dominant factor for differences in overall plant thermal efficiency for the cases studied.

Figure 5-14: Plant Thermal Efficiency Comparison

The resulting energy penalty for Updated Case 2 is about 27 percent as compared to Case 1. There are two primary reasons for the energy penalty associated with Updated Case 2. First, the integration into the power plant of the Air Separation Unit (ASU) to provide combustion oxygen, and second, the Gas Processing System (GPS) to, compress, purify, and liquefy the CO₂ product. The oxygen-fired cases utilize a cryogenic based ASU system, which adds a significant load to the plant auxiliary power requirement (about 231 kWh/ton of oxygen supplied or about 18 percent of the steam turbine generator output). The GPS power requirements were calculated to be about 153 kWh/ton of CO₂ captured or about 12 percent of the steam turbine generator output. Both these systems (ASU and GPS) consume large quantities of auxiliary power as shown in Table 5-2.

Plant CO₂ Emissions:

Figure 5-15 compares overall plant CO₂ emissions on a normalized basis (lbm/kWh) among these three cases. Also shown in this figure are the quantities of captured CO₂ (normalized basis). The Base Case (Case-1) emits about 2.0 lbm/kWh of CO₂ as is typical for coal fired power plants with subcritical steam cycles. The oxygen-fired cases, which include CO₂ capture systems, show normalized CO₂ emissions of about 0.18 lbm/kWh of CO₂. Both of the oxygen-fired cases capture more than 93 percent of the CO₂ produced.

Figure 5-15: Plant CO₂ Emission Comparison

The upper bars (lighter shade) and the upper set of data labels (bold) shown on the figure indicate the normalized quantities of CO₂ captured. The captured quantities of CO₂ are about 2.62 and 2.46 lbm/kWh for Original Case-2 and Updated Case-2 respectively. The lower bars (darker shade) and the lower set of data labels show the normalized CO₂ emitted. The emitted quantity of CO₂ is about 0.18 and 0.17 lbm/kWh for the CO₂ capture cases. The sum of these two quantities (captured + emitted) represents the quantity of CO₂ produced (e.g., the Updated Case-2 power plant produces $2.46 + 0.17 = 2.63$ lbm/kWh of CO₂ on a normalized basis).

Figure 5-16 compares avoided CO₂ emissions on a normalized basis (lbm/kWh) among the three cases. The avoided CO₂ emissions are calculated relative to the Base Case (Case-1). The avoided quantities of CO₂ for both Original Case-2 and Updated Case-2 are about 1.8 lbm/kWh.

Figure 5-16: Avoided CO₂ Emission Comparison (Relative to Case-1)

6 COST ANALYSIS

The plant capital cost estimate summaries, including engineering, procurement, and construction (EPC basis), are shown in this section for the three (3) power plants included in this study. Costs for two of the cases shown (Case-1 and Original Case-2) are copied directly from the Phase I Topical Report (Marion, et al., 2003) for comparison purposes. Updated Case-2 provides the revised cost estimate for the new compact oxygen fired CFB with CO₂ capture (i.e. an update of Original Case-2 from the Phase I study). A detailed cost breakdown for each case is included in Appendix I. The EPC basis does not include owner's costs. Owner's costs are, however, included in the economic analysis (Refer to Section 7). All costs shown are expressed in July 2003 dollars. The level of accuracy for these conceptual level designs is expected to be about ± 30 percent. These plants are assumed to be constructed on a common Greenfield site in the Gulf Coast region of southeastern Texas.

Investment Cost Estimation Basis:

The EPC costs include all required equipment including the traditional Boiler Island equipment, and Balance of Plant equipment (steam turbine, condensate and feedwater, draft system, gas clean-up, material handling, cooling, electrical, instrumentation and control, and misc.). Additionally, for the CO₂ removal concepts (Original Case-2 and Updated Case-2) the non-traditional equipment required for CO₂ capture, compression and liquefaction is included, and the Air Separation Unit equipment is also included.

The boundary limit for these plants includes the complete plant facility within the “fence line”. It includes the coal receiving and water supply systems and terminates at the high-voltage side of the main power transformers. Also, for the cases with CO₂ capture, the boundary terminates at the outlet flange of the CO₂ product pipe (It does not include the CO₂ pipe offsite).

The costs include equipment, materials, labor, indirect construction costs, and engineering. The labor cost to install the equipment and materials was estimated on the basis of labor man-hours. The labor costing approach was a multiple contract labor basis with the labor cost including direct and indirect labor cost plus fringe benefits and allocations for contractor expenses and markup.

These costs include professional services and “other costs”. Professional services consist of the cost for engineering, construction management, and startup assistance. The engineering services include all preliminary and detailed engineering and design for the total plant scope. It includes specifying equipment for purchase, procurement, performing project scheduling and cost control services for the project; providing engineering and design liaison during the construction period; and providing startup support. Construction management services cost includes a field management staff capable of performing all field contract administration; field inspection and quality assurance; project construction control; safety and medical services as required; field and construction insurance administration, field office clerical and administrative support. The “other costs” category includes a cost allowance for freight costs, heavy haul, insurance and taxes, and indirect startup spares.

The capital cost estimates for these plants were calculated based on a combination of

vendor-furnished quotes, and cost estimating database values. The Boiler Island costs were estimated based on calculated material weights for all components. Conceptual equipment arrangement drawings and equipment lists were developed as a part of the conceptual design of the required equipment.

The following assumptions were made in developing the EPC cost estimates for each concept evaluated:

- Investment costs are expressed in July 2003 US dollars
- Construction labor rates are based on Gulf Coast non-union rates
- The plant is constructed on a Greenfield site in southeastern Texas
- All costs are based on mature level (nth plant) commercial design
- Owners costs (including interest during construction, start-up fuel, land, land rights, plant licensing, permits, etc.) are not included in the investment costs but are included in the Cost of Electricity analysis
- Ash is to be shipped off site with provisions for short-term storage only
- Outdoor installation for Gas Processing System (GPS)
- Investment in new utility systems is outside the scope
- CO₂ pipeline is outside the scope
- No special limitations for transportation of large equipment
- No protection against unusual airborne contaminants (dust, salt, etc.)
- No unusual wind storms
- No earthquakes
- No piling required
- All releases can go to atmosphere – no flare provided
- CO₂ Pump designed to API standards, all other pumps conform to ANSI
- All GPS heat exchangers designed to TEMA “C”
- All GPS vessels are designed to ASME Section VIII, Div 1.
- Annual operating time is 7008 h/yr (80 percent capacity factor).
- The investment cost estimate was developed as a factored estimate based on a combination of vendor quotes and in-house data for the major equipment. Such an estimate can be expected to have accuracy of ±30 percent.
- No purchases of utilities or charges for shutdown time have been charged against the project.

Other exclusions from the EPC investment cost estimate are as follows:

- CO₂ pipeline offsite

- Fuels required for startup
- Relocation or removal of buildings, utilities, and highways
- Permits
- Land and land rights
- Soil investigation
- Environmental Permits
- Disposal of hazardous or toxic waste
- Disposal of existing materials
- Custom's and Import duties
- Sales/Use tax.
- Forward Escalation
- Capital spare parts
- Chemical loading facilities
- GPS Buildings except for Compressor building and electrical substation.
- Financing cost
- Owners costs
- Guards during construction
- Site Medical and Ambulance service
- Cost & Fees of Authorities
- Overhead High voltage feed lines
- Cost to run a natural gas pipeline to the plant
- Excessive piling

Operating and Maintenance Cost Estimation Basis:

Operating and maintenance (O&M) costs were also calculated for all systems. The variable operating and maintenance (VOM) costs for the new equipment included such categories as chemicals and desiccants, waste handling, maintenance material and labor, supplemental fuel usage, and contracted services. The fixed operating and maintenance (FOM) costs for the new equipment includes operating labor only.

Operating and Maintenance (O&M) costs are calculated for each plant and are listed as either fixed or variable. The fixed costs are those costs, which are incurred irrespective of the number of hours of plant operation whereas the variable costs are directly proportional to the operating hours. These costs are calculated separately for the traditional power plant equipment, the oxygen supply systems (ASU), and the Gas Processing Systems (GPS's) where applicable. The O&M costs for the ASU was calculated by Praxair and the O&M costs for the GPS's were calculated by ABB

Lummus Global Inc. (Lummus).

The O&M costs for the traditional power plant equipment was developed quantitatively by Parsons and ALSTOM. Operating labor cost was calculated based on the number of operator jobs (O.J.) required. The average labor rate used to determine the annual cost was 30.90 \$/hr, with a labor burden of 30 percent. The labor administration and overhead cost was assessed at a rate of 25 percent of the O&M labor. Maintenance cost was evaluated as a percentage of the initial capital cost.

Consumable costs including fuel, limestone, water, and chemicals were determined on the basis of individual flow rates as listed in the material and energy balances, individual unit costs (listed below), and the plant annual operating hours. Waste disposal cost was also based on flow rates from the material and energy balances, unit costs, and operating hours.

- Coal cost - 1.25 \$/MM-Btu
- Natural Gas cost - 4.00 \$/MM-Btu
- Limestone cost - 10.00 \$/Ton
- Lime cost - 50.00 \$/Ton
- Water cost - 1.00 \$/1,000 gallons
- Water Treatment Chemicals cost - 0.16 \$/lbm
- Ash Disposal cost - 8.00 \$/Ton

By-product credits were not considered for these cases except in the economic sensitivity study (Refer to Section 7) where a credit for CO₂ product was calculated using values ranging from 0.0 to 20.0 \$/Ton of CO₂.

6.1 Plant Investment Cost and Operating and Maintenance Cost Summary:

A summary of costs (Capital and O&M) for the three cases are shown in Table 6-1. A major account level breakdown of the costs for each case is shown later in this section and a detailed cost breakdown for each case is included in Appendix I.

Table 6-1: EPC Plant Costs and O&M Costs Summary

Study Case	Total Investment Cost (EPC Basis)		Operating & Maintenance Costs				
	\$ x 1000	\$/kW	Fixed		Variable @80%CF		Total
			\$	\$/kW	\$	\$/kWh	
Case-1: Air-Fired CFB w/o CO ₂ Capture	251,804	1,304	5,657,635	29.31	5,587,188	0.0041	11,244,823
Case-2 Original: O ₂ -Fired CFB w/ASU & CO ₂ Capture	328,589	2,443	7,853,885	58.39	8,820,048	0.0094	16,673,933
Case-2 Updated: O ₂ -Fired CFB w/ASU & CO ₂ Capture	329,610	2,382	7,858,989	56.78	8,834,994	0.0091	16,693,983

Figure 6-1 shows the specific EPC costs (\$/kW-net and \$/kW-gross basis) for each case. The graph on the left shows the impact of the relatively costly and power consuming Air Separation Unit and Gas Processing System with the specific cost of the power plant (\$/kW-net) increasing by almost 80 percent as compared to the air fired Case-1 without CO₂ capture. The graph on the right shows the costs on a \$/kW-gross basis separated into three categories:

- Traditional Power Plant Equipment
- Air Separation Unit
- Gas Processing System

The use of the \$/kWgross units for the graph on the right clearly shows the savings incurred for the traditional power plant equipment as a result of oxygen firing. This savings totals to about 20 percent (1,205 \$/kW vs. 988 or 983 \$/kW) and is primarily a result of cost reductions in the Boiler Island for equipment where cost is strongly dependent on gas flow (combustor, cyclones, backpass heat exchangers, air heater, fans, ductwork, particulate removal system, etc.).

Figure 6-1: EPC Plant Costs – Summary of all Cases

As indicated, most of the cost savings incurred for the traditional power plant equipment are attributed to savings in the Boiler Island. Figure 6-2 shows the EPC cost savings for the CFB boiler in particular. Units of \$/kW-gross are used instead of \$/kW-net because of the large amount of auxiliary power associated with the oxygen fired case. This large auxiliary power difference between the cases would therefore distort the savings when using the more traditional \$/kW-net units. The oxygen fired boiler EPC cost is shown to be about 65% of that for the air fired boiler. It should be emphasized that the heat output of both boilers is nearly identical. This CFB boiler is the component where most of the traditional power plant savings described above came from.

Figure 6-2: EPC Boiler Costs (\$/kW-gross)

Overall plant costs and the associated specific plant costs (\$/kW) can vary quite significantly for any given plant design depending on several factors. Some of the more important factors are listed below.

- Plant Location and Site Conditions
- Construction Labor Basis
- Coal Analysis
- Ambient Conditions

For the cases in this study, the design coal analysis, design ambient conditions, plant location and site conditions are described under Plant Design Basis in the Phase I Topical Report (Marion, et al., 2003). The construction labor basis used is Gulf Coast non-union. The sensitivity of plant specific cost to construction labor basis is indicated by observing that for these studies, changing from Gulf Coast non-union to Ohio River Valley union basis, for example, would increase the EPC plant costs by about 20 percent.

6.1.1 Case-1: Investment Costs and Operating and Maintenance Costs (copied from Marion et al., 2003)

The capital cost estimate of the Case-1 plant (Table 6-2) without CO₂ capture was developed consistent with the approach and basis identified in the design basis. The capital cost estimate is expressed in July 2003 dollars. The production cost and expenses were developed on a first-year basis with a July 2003 plant in-service date.

The production costs consist of plant operating labor, maintenance (material and labor), an allowance for administrative and support labor, consumables, solid waste disposal and fuel costs. The costs were determined on a first-year basis that includes evaluation at an equivalent plant operating capacity factor of 80 percent. The results are summarized in Table 6-3.

Table 6-2: Case-1: Total Plant Investment Cost Summary (copied from Marion et al., 2003)

Acct. No.	Account Description	Total Cost	
		\$x1000	\$/kW
1	FUEL & SORBENT HANDLING	14,193	74
2	FUEL & SORBENT PREP. & FEED	5,626	29
3	FEEDWATER & MISC. BOP SYSTEMS	20,361	105
4	FLUIDIZED BED BOILER	79,409	411
5	FLUE GAS CLEANUP	8,781	45
6	COMBUSTION TURBINE ACCESSORIES		
7	HRSG DUCTING & STACK	13,306	69
8	STEAM TURBINE GENERATOR	37,654	195
9	COOLING WATER SYSTEM	11,835	61
10	ASH/SPENT SORBENT HANDLING SYSTEMS	9,157	47
11	ACCESSORY ELECTRIC PLANT	18,141	94
12	INSTRUMENTATION & CONTROL	11,045	57
13	IMPROVEMENT TO SITE	5,291	27
14	BUILDINGS & STRUCTURES	17,005	88
TOTAL COST		251,804	1,304

Table 6-3: Case-1: Total Plant Operating and Maintenance Costs (copied from Marion et al., 2003)

Client: ALSTOM Power Inc.	ANNUAL O&M EXPENSES	Cost Base: Jul-03
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers	Case 1 - 1x200 MW Air-Fired CFB w/o CO2 Capture	
	Net Plant Heat Rate (Btu/kWh): 9,611	
	Net Power Output (kW): 193,037	
	Capacity Factor (%): 80	
BOILER ISLAND AND BALANCE OF PLANT O&M COSTS		
<i>TOTAL FIXED O&M COSTS</i>	<u>Annual Cost, \$</u> 5,657,635	<u>Annual Unit Cost, \$/kW</u> 29.31
<i>TOTAL VARIABLE O&M COSTS</i>	<u>Annual Cost, \$</u> 5,587,188	<u>Annual Unit Cost, \$/kWh</u> 0.0041
AIR SEPARATION UNIT (ASU)		
<i>TOTAL FIXED O&M COSTS</i>	<u>Annual Cost, \$</u> -	<u>Annual Unit Cost, \$/kW</u> 0.00
<i>TOTAL VARIABLE O&M COSTS</i>	<u>Annual Cost, \$</u> -	<u>Annual Unit Cost, \$/kWh</u> 0.000000
GAS PROCESSING SYSTEM (GPS)		
<i>TOTAL FIXED O&M COSTS</i>	<u>Annual Cost, \$</u> -	<u>Annual Unit Cost, \$/kW</u> 0.00
<i>TOTAL VARIABLE OPERATING COST</i>	<u>Annual Cost, \$</u> -	<u>Annual Unit Cost, \$/kWh</u> 0.0000
TOTAL PLANT O&M COSTS		
<i>TOTAL FIXED OPERATING COSTS</i>	<u>Annual Cost, \$</u> 5,657,635	<u>Annual Unit Cost, \$/kW</u> 29.31
<i>TOTAL VARIABLE OPERATING COST</i>	<u>Annual Cost, \$</u> 5,587,188	<u>Annual Unit Cost, \$/kWh</u> 0.00413

6.1.2 Original Case-2: Investment Costs and Operating and Maintenance Costs (copied from Marion et al., 2003)

The capital cost estimate of the Original Case-2 plant (Table 6-4) with CO₂ capture was developed consistent with the approach and basis identified in the design basis. The capital cost estimate is expressed in July 2003 dollars. The production cost and expenses were developed on a first-year basis with a July 2003 plant in-service date.

The production costs consist of plant operating labor, maintenance (material and labor), an allowance for administrative and support labor, consumables, solid waste disposal and fuel costs. The costs were determined on a first-year basis that includes evaluation at an equivalent plant operating capacity factor of 80 percent. The results are summarized in Table 6-5.

Table 6-4: Original Case-2: Total Plant Investment Cost Summary (copied from Marion et al., 2003)

Acct. No.	Account Description	Total Cost	
		\$x1000	\$/kW
1	FUEL & SORBENT HANDLING	13,955	104
2	FUEL & SORBENT PREP. & FEED	5,526	41
3	FEEDWATER & MISC. BOP SYSTEMS	21,317	158
4	FLUIDIZED BED BOILER & AIR SEP. UNIT	114,291	850
5	FLUE GAS CLEANUP & GAS PROC. SYSTEM	60,137	447
6	COMBUSTION TURBINE ACCESSORIES		
7	HRSG DUCTING & STACK	965	7
8	STEAM TURBINE GENERATOR	37,777	281
9	COOLING WATER SYSTEM	11,869	88
10	ASH/SPENT SORBENT HANDLING SYSTEMS	8,045	60
11	ACCESSORY ELECTRIC PLANT	21,815	162
12	INSTRUMENTATION & CONTROL	10,545	78
13	IMPROVEMENT TO SITE	5,304	39
14	BUILDINGS & STRUCTURES	17,043	127
TOTAL COST		328,589	2,443

Table 6-5: Original Case-2: Total Plant Operating and Maintenance Costs (copied from Marion et al., 2003)

Client: ALSTOM Power Inc. Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers		ANNUAL O&M EXPENSES Case 2 - 1x200 gr. MW O ₂ -Fired CFB w/ASU & CO ₂ Capture Net Plant Heat Rate (Btu/kWh): 13,548 Net Power Output (kW): 134,514 Capacity Factor (%): 80	Cost Base: Jul-03
BOILER ISLAND AND BALANCE OF PLANT O&M COSTS			
<i>TOTAL FIXED O&M COSTS</i>		<u>Annual Cost, \$</u> 5,435,950	<u>Annual Unit Cost, \$/kW</u> 40.41
<i>TOTAL VARIABLE O&M COSTS</i>		<u>Annual Cost, \$</u> 5,786,394	<u>Annual Unit Cost, \$/kWh</u> 0.0061
AIR SEPARATION UNIT (ASU)			
<i>TOTAL FIXED O&M COSTS</i>		<u>Annual Cost, \$</u> 2,111,335	<u>Annual Unit Cost, \$/kW</u> 15.70
<i>TOTAL VARIABLE O&M COSTS</i>		<u>Annual Cost, \$</u> 397,253	<u>Annual Unit Cost, \$/kWh</u> 0.000421
GAS PROCESSING SYSTEM (GPS)			
<i>TOTAL FIXED O&M COSTS</i>		<u>Annual Cost, \$</u> 306,600	<u>Annual Unit Cost, \$/kW</u> 2.28
<i>TOTAL VARIABLE OPERATING COST</i>		<u>Annual Cost, \$</u> 2,636,401	<u>Annual Unit Cost, \$/kWh</u> 0.0028
TOTAL PLANT O&M COSTS			
<i>TOTAL FIXED OPERATING COSTS</i>		<u>Annual Cost, \$</u> 7,853,885	<u>Annual Unit Cost, \$/kW</u> 58.39
<i>TOTAL VARIABLE OPERATING COST</i>		<u>Annual Cost, \$</u> 8,820,048	<u>Annual Unit Cost, \$/kWh</u> 0.00936

6.1.3 Updated Case-2: Investment Costs and Operating and Maintenance Costs

The capital cost estimate of Updated Case-2 (Table 6-6) with CO₂ capture was developed consistent with the approach and basis identified in the design basis. The capital cost estimate is expressed in July 2003 dollars. The production cost and expenses were developed on a first-year basis with a July 2003 plant in-service date.

The production costs consist of plant operating labor, maintenance (material and labor), an allowance for administrative and support labor, consumables, solid waste disposal and fuel costs. The costs were determined on a first-year basis that includes evaluation at an equivalent plant operating capacity factor of 80 percent. The results are summarized in Table 6-7.

Table 6-6: Updated Case-2: Total Plant Investment Cost Summary

Acct. No.	Account Description	Total Cost	
		\$x1000	\$/kW
1	FUEL & SORBENT HANDLING	11,903	86
2	FUEL & SORBENT PREP. & FEED	4,467	32
3	FEEDWATER & MISC. BOP SYSTEMS	19,394	140
4	FLUIDIZED BED BOILER & AIR SEP. UNIT	113,643	821
5	FLUE GAS CLEANUP & GAS PROC. SYSTEM	67,538	488
6	COMBUSTION TURBINE ACCESSORIES		
7	HRSG DUCTING & STACK	965	7
8	STEAM TURBINE GENERATOR	38,131	276
9	COOLING WATER SYSTEM	11,782	85
10	ASH/SPENT SORBENT HANDLING SYSTEMS	7,213	52
11	ACCESSORY ELECTRIC PLANT	21,681	157
12	INSTRUMENTATION & CONTROL	10,545	76
13	IMPROVEMENT TO SITE	5,304	38
14	BUILDINGS & STRUCTURES	17,043	123
TOTAL COST		329,610	2,382

Table 6-7: Updated Case-2: Total Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc. Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers		ANNUAL O&M EXPENSES Case 2 - 1x200 gross MW O ₂ -Fired CFB w/ASU & CO ₂ Capture Net Plant Heat Rate (Btu/kWh): 13,152 Net Power Output (kW): 138,402 Capacity Factor (%): 80	Cost Base: Jul-03
BOILER ISLAND AND BALANCE OF PLANT O&M COSTS			
<i>TOTAL FIXED O&M COSTS</i>		<u>Annual Cost, \$</u> 5,441,054	<u>Annual Unit Cost, \$/kW</u> 39.31
<i>TOTAL VARIABLE O&M COSTS</i>		<u>Annual Cost, \$</u> 5,801,340	<u>Annual Unit Cost, \$/kWh</u> 0.0060
AIR SEPARATION UNIT (ASU)			
<i>TOTAL FIXED O&M COSTS</i>		<u>Annual Cost, \$</u> 2,111,335	<u>Annual Unit Cost, \$/kW</u> 15.26
<i>TOTAL VARIABLE O&M COSTS</i>		<u>Annual Cost, \$</u> 397,253	<u>Annual Unit Cost, \$/kWh</u> 0.000410
GAS PROCESSING SYSTEM (GPS)			
<i>TOTAL FIXED O&M COSTS</i>		<u>Annual Cost, \$</u> 306,600	<u>Annual Unit Cost, \$/kW</u> 2.22
<i>TOTAL VARIABLE OPERATING COST</i>		<u>Annual Cost, \$</u> 2,636,401	<u>Annual Unit Cost, \$/kWh</u> 0.0027
TOTAL PLANT O&M COSTS			
<i>TOTAL FIXED OPERATING COSTS</i>		<u>Annual Cost, \$</u> 7,858,989	<u>Annual Unit Cost, \$/kW</u> 56.78
<i>TOTAL VARIABLE OPERATING COST</i>		<u>Annual Cost, \$</u> 8,834,994	<u>Annual Unit Cost, \$/kWh</u> 0.00911

7 ECONOMIC ANALYSIS

This section shows the results of an economic evaluation that compares the two CO₂ capture concepts (Original Case-2 and Updated Case-2) with the Base Case study unit without CO₂ capture (Case-1). The results of an economic sensitivity analysis are also shown for the three cases. The economic results shown in this section for Case-1 and Original Case-2 were copied from the Phase I Topical Report (Marion et al., 2003).

The purpose of this evaluation is to update the economic analysis, which was done for Original Case-2. The updated results are labeled as Updated Case-2. Additionally a comparison between the three cases (Case-1, Original Case-2, and Updated Case-2) is also provided. The economic evaluation results are presented as both total and incremental Costs of Electricity (levelized basis). The incremental costs of electricity are incremental relative to the Base Case (Case-1). CO₂ mitigation costs (\$/Ton of CO₂ avoided) were also determined in this analysis for each capture case relative to Case-1. The comparisons shown quantify the impact of the changes that were made to Updated Case-2 on economic parameters of merit.

The model used to perform the economic evaluations was the proprietary ALSTOM Power Plant Laboratories' Project Economic Evaluation Pro-Forma. This cash flow model, developed by the Company's Project & Trade Finance group, has the capability to analyze the economic effects of different technologies based on differing efficiencies, investment costs, operating and maintenance costs, fuel costs, and cost of capital assumptions. Various categories of results are available from the model. In addition to cost of electricity, net present value, project internal rate of return, payback period, and other evaluation parameters are available.

Economic Analysis Assumptions:

Numerous financial assumptions were required in performing the economic evaluations. These assumptions are listed in Table 7-1. Items shaded in yellow represent parameters that were varied in the economic sensitivity study.

Table 7-1: Economic Evaluation Study Assumptions

	CFB System - Case 2		CFB System Case 2
POWER GENERATION		DEBT PORTFOLIO	
Net output (MW)	Case Sensitive	Interest Rates (Financed) ¹	
Capacity factor (%)	100%	During Construction	
Availability factor (%)	80%	Base Rate	1.32%
Net plant heat rate, HHV (Btu per kWh)	Case Sensitive	Swap/Reinvestment cushion	1.28%
Degradation factor (%)	0.0%	Fixed Rate Margin	3.00%
TIME FRAME		All-In Fixed Rate	5.60%
Construction period (months)	30	During Operation	
Depreciation Term (years)	30	Base Rate	1.32%
Analysis Horizon (years)	30	Swap/Reinvestment cushion	1.28%
PROJECT COSTS		Fixed Rate Margin	2.50%
EPC Price (\$1000s)	Case Sensitive	All-In Fixed Rate	5.10%
Fixed O&M costs (\$ per kW)	Case Sensitive	Up-front Fee (Financed)	2.0%
Variable O&M costs (cents per kWh)	Case Sensitive	Commitment Fee	1.0%
Owner's EPC Contingency	0.0%	Grace Period (months)	
Initial spares and consumables	1.0%	Loan Tenor (years after construction)	30
Insurance		TAXES	
Insurance during Construction	1.0%	Corporate Tax	20.0%
Insurance during first year of operation	0.5%	Tax holiday (years after commissioning)	0.0%
Development Costs		Customs Duty	0.0%
Development Costs & Fees	4.0%	Customs Clearance Fee	0.0%
Reimbursable Dev't Costs	3.0%	COST OF CAPITAL ASSUMPTIONS	
Advisory Fees	3.0%	Discount Factor	10.0%
Financial and Legal Fees	3.0%	PROGRESS PAYMENT SCHEDULES	
Start-up Fuel	0.5%		
Fuel Stock Pile	0.5%		
Other Costs	0.5%		
Total Initial Project Costs (% of EPC)	17.0%		
FUEL COST			
Coal Price (\$ per MMBtu)	1.25	Month	
Natural Gas Price (\$ per MMBtu)	4.00	1	10%
PROJECT CREDITS		8	15%
CO ₂ Sell Price (\$ per ton)	0.00	10	
N ₂ Sell Price (\$ per ton)	0.00	16	
ESCALATION FACTORS		17	25%
Coal Price	0.0%	20	
Variable O&M	0.0%	22	20%
Fixed O&M (including payroll)	0.0%	25	
Consumer Price Index	0.0%	26	20%
FINANCING ASSUMPTIONS		30	10%
Equity	50.0%	31	
Debt	50.0%	32	
		36	
		41	
		48	
		Total	100%

Cost of Electricity Calculation:

Levelized cost of electricity (COE) was used as one criterion to compare the systems in this study. The levelized cost of electricity result is comprised of the following four components: financial, fixed O&M, variable O&M, and fuel. The cash flow model used is structured to calculate the corresponding annual cash flows for each of these items over the evaluation life of the project. The annual expenses are distributed over the corresponding net annual electricity generated (kWh/year) in order to determine a unit cost (cents/kWh). These costs are subsequently levelized to get a corresponding value of each component over the plant life. In other words, each of the cash flow streams are converted to annuity payments corresponding to a constant value over the life of the study.

The **financial** component of the COE represents the costs which are associated with payment of the original engineered, procured and constructed (EPC) price, all associated owner's costs, custom's and financing fees, and interests accrued both during construction and during operation. The **fixed O&M** component represents the costs that occur regardless of whether the unit is in operation or not. The **variable O&M** component represents the incremental costs which occur when the unit is in operation. The **fuel** cost component represents the cost of the fuel, which is consumed.

Economic Analysis Results:

The economic analysis results are discussed in this section. The case studies are compared using several evaluation criterion including levelized cost of electricity (COE), incremental COE with respect to the reference plant without CO₂ capture (Case-1), and by mitigated costs of avoided CO₂, also with respect to Case-1.

The equation for the incremental COE is defined as:

$$\text{Incremental COE} = (\text{COE}_{\text{CP}} - \text{COE}_{\text{ref}})$$

Where:

COE $\equiv$ levelized Cost of Electricity (cents / kWh),

CP $\equiv$ Capture Plant, and

ref $\equiv$ Reference Plant.

The equation for the mitigation cost is defined as:

$$\text{Mitigation Cost} = \frac{\text{COE}_{\text{CP}} - \text{COE}_{\text{ref}}}{\text{CO2}_{\text{ref}} - \text{CO2}_{\text{CP}}}$$

Where:

Mitigation Cost $\equiv$ \$/ton of CO₂ Avoided,

COE $\equiv$ levelized Cost of Electricity (\$ / kWh),

CO₂ $\equiv$ Carbon dioxide emitted (ton / kWh),

CP $\equiv$ Capture Plant, and

ref $\equiv$ Reference Plant.

The levelized COE results for the three cases are summarized in Table 7-2. The air-fired CFB system (Case-1) is the reference plant against which the systems with CO₂ capture (Original Case-2 and Updated Case-2) are compared. The reference plant (Case-1) is included in Table 7-2 to provide a comparison to a “business as usual” scenario. The incremental COE and mitigation cost results for the CO₂ capture technologies are also provided in Table 7-2 relative to Case-1.

Table 7-2: Economic Analysis Summary (Case 1, Original Case-2, and Updated Case-2)

	Case 1 - Air Fired CFB w/o CO ₂ Capture	Case 2 - O ₂ Fired CFB w/ ASU & CO ₂ Capture	Case 2 Updated - O ₂ Fired CFB w/ ASU & CO ₂ Capture
Levelized Cost of Electricity at 80% Availability Factor (cents / kWh)			
Financial	2.5	4.7	4.5
Fixed O&M	0.4	0.8	0.8
Variable O&M	0.4	0.9	0.9
Fuel	1.2	1.7	1.6
Total	4.5	8.1	7.9
Incremental Cost of Electricity at 80% Availability Factor (cents / kWh)			
Reference Plant		3.6	3.4
Mitigated Cost (\$ / ton CO ₂ avoided at 80% Availability Factor)			
Reference Plant		40	37

The economic study results are summarized by comparing the levelized COE, incremental COE, and CO₂ mitigation cost results for these CO₂ capture technologies as shown in Figure 7-1-Figure 7-3, respectively. The levelized COE value is increased from about 4.5 cents/kWh without CO₂ capture (Case-1) to about 8.1 cents/kWh with the addition of the CO₂ capture system for Original Case-2 and 7.9 cents/kWh for Updated Case-2. The update of Case-2 shows a relatively minimal impact on levelized COE (about -2%) as shown by comparing Original Case-2 and Updated Case-2. The slight improvement in COE for Updated Case-2 is primarily the result of a small decrease in the specific plant cost (about 2.5% - \$/kW basis) and a small increase in the net plant output (about 2.9%). The incremental COE as compared to the air-fired CFB reference plant without CO₂ capture (Case 1) is about 3.6 cents / kWh for Original Case-2 and 3.4 cents/kWh for Updated Case-2. Similarly, the CO₂ mitigation costs are about 40 and 37 \$/ton of CO₂ avoided respectively.

Figure 7-1: Levelized Cost of Electricity Comparison

Figure 7-2: Incremental Cost of Electricity for CO₂ Capture

Figure 7-3: CO₂ Mitigation Costs (\$/ton of CO₂ avoided)

It should be noted that the economic results for Original Case-2 reported in the Phase I Topical Report (Marion, et al., 2003) were found to be slightly in error. These results used as a program input the net plant output value of 132,423 kW. The correct value for the net plant output is 134,514 kW (refer to Table 5-2). The impact of this error is that the levelized COE was decreased from 8.25 to 8.1 Cents/kWh. Similarly, the incremental COE was reduced from 3.7 to 3.6 Cents/kWh and the CO₂ mitigation cost was reduced from 41 to 40 \$/ton of avoided CO₂ (both relative to air fired Case-1).

Sensitivity Study Results:

Sensitivity analyses were conducted for these three case studies to determine the effect on levelized COE of variations of selected base parameter values by ± 25 percent and CO₂ by-product selling price up to \$20 per ton. These parameters (shaded in yellow in Table 7-1) are capacity factor, EPC plant price, coal price, CO₂ credit sell price, equity rate, corporate tax rate, and the discount rate for cost of capital. The base parameter values represent the point where all the sensitivity curves intersect (point 0, 0). Sensitivity analysis “spider plots” for air fired Case-1, Original Case 2, and Updated Case-2 are provided in this section. The tabulated results for all three cases are also provided.

In general, for the variable ranges studied, CO₂ selling price, capacity factor, plant investment cost, and discount rate, in order of decreasing significance, have the greatest effect on the COE.

Case 1 - Air-Fired CFB without CO₂ Capture

Results for the air fired Case 1 COE sensitivity study are shown in Figure 7-4 and Table 7-3. The levelized COE for the base parameter values is about 4.5 cents per kWh. Levelized COE ranges from a low of 3.9 to a high of 5.5 cents per kWh.

Figure 7-4: Case 1 - Air-Fired CFB without CO₂ Capture - Economic Sensitivity Analysis Results

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Table 7-3: Case 1 - Air-Fired CFB without CO₂ Capture - Economic Sensitivity Analysis
Results

Parameter	Units	Case 1 - Air-Fired CFB without CO2 Capture												
Power Generation														
Net Output	kW	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037
Availability Factor	%	80	60	70	90	100	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	5,256	6,132	7,884	8,760	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5
Net Plant Heat Rate, HHV	Btu / kWh	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611
Net Generation	MWh / year	1,352,803	1,014,602	1,183,703	1,521,904	1,691,004	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803
Costs														
EPC Price	\$ / kW	1,304	1,304	1,304	1,304	1,304	978	1,141	1,467	1,631	1,304	1,304	1,304	1,304
EPC Price	\$1000s	251,804	251,804	251,804	251,804	251,804	188,853	220,329	283,280	314,755	251,804	251,804	251,804	251,804
Fixed O&M Costs	\$1000 / year	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658
Fixed O&M Costs	\$ / kW	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31
Variable O&M Costs	\$1000 / year	5,587	4,190	4,889	6,286	6,984	5,587	5,587	5,587	5,587	5,587	5,587	5,587	5,587
Variable O&M Costs	cents / kWh	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41
Total O&M Costs	cents / kWh	0.83	0.97	0.89	0.78	0.75	0.83	0.83	0.83	0.83	0.83	0.83	0.83	0.83
Fuel Cost Calculation														
Coal Price	\$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	0.94	1.09	1.41	1.56
CO2 Emissions														
CO2 Produced	lbm / h	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427
CO2 Produced	lbm / kWh	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00
CO2 Emitted	lbm / h	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427
CO2 Emitted	lbm / kWh	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00
Financing Assumptions														
Equity	%	50	50	50	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	20	20	20	20	20	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	10	10	10	10	10
Levelized Cost of Electricity (cents / kWh)														
Financial Component		2.5	3.3	2.9	2.2	2.0	1.9	2.2	2.8	3.1	2.5	2.5	2.5	2.5
Fixed O&M		0.4	0.6	0.5	0.4	0.3	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4
Variable O&M		0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4
Fuel		1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	0.9	1.1	1.4	1.5
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		4.5	5.5	4.9	4.2	3.9	3.9	4.2	4.8	5.2	4.2	4.4	4.7	4.8

Parameter	Units	Case 1 - Air-Fired CFB without CO2 Capture												
Power Generation														
Net Output	kW	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037	193,037
Availability Factor	%	80	80	80	80	80	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5	35.5
Net Plant Heat Rate, HHV	Btu / kWh	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611	9,611
Net Generation	MWh / year	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803	1,352,803
Costs														
EPC Price	\$ / kW	1,304	1,304	1,304	1,304	1,304	1,304	1,304	1,304	1,304	1,304	1,304	1,304	1,304
EPC Price	\$1000s	251,804	251,804	251,804	251,804	251,804	251,804	251,804	251,804	251,804	251,804	251,804	251,804	251,804
Fixed O&M Costs	\$1000 / year	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658	5,658
Fixed O&M Costs	\$ / kW	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31	29.31
Variable O&M Costs	\$1000 / year	5,587	5,587	5,587	5,587	5,587	5,587	5,587	5,587	5,587	5,587	5,587	5,587	5,587
Variable O&M Costs	cents / kWh	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41	0.41
Total O&M Costs	cents / kWh	0.83	0.83	0.83	0.83	0.83	0.83	0.83	0.83	0.83	0.83	0.83	0.83	0.83
Fuel Cost Calculation														
Coal Price	\$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25
CO2 Emissions														
CO2 Produced	lbm / h	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427
CO2 Produced	lbm / kWh	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00
CO2 Emitted	lbm / h	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427	385,427
CO2 Emitted	lbm / kWh	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00
Financing Assumptions														
Equity	%	38	44	56	63	50	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	15	18	23	25	20	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	8	9	11	13	13
Levelized Cost of Electricity (cents / kWh)														
Financial Component		2.3	2.4	2.6	2.7	2.4	2.5	2.5	2.6	2.1	2.3	2.7	2.9	2.9
Fixed O&M		0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4
Variable O&M		0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4
Fuel		1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		4.4	4.4	4.6	4.7	4.5	4.5	4.6	4.6	4.1	4.3	4.7	5.0	5.0

Original Case 2 - Oxygen-Fired CFB with Air Separation Unit and CO₂ Capture

Results for the Original Case 2 COE sensitivity study are shown in Figure 7-5 and Table 7-4. The levelized COE for the base parameter values is 8.1 cents per kWh. Levelized COE ranges from a low of 5.5 to a high of 10.0 cents per kWh. CO₂ mitigation costs ranged from 11 to 60 \$/ton of CO₂ avoided (reference plant is Case 1) with the baseline value at 40 \$/ton of CO₂ avoided.

Figure 7-5: Original Case 2 - Oxygen-Fired CFB with ASU and CO₂ Capture - Economic Sensitivity Analysis Results

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Table 7-4: Original Case 2 - Oxygen-Fired CFB with ASU and CO₂ Capture - Economic
Sensitivity Analysis Results

Parameter	Units	Case 2 - Oxygen-Fired CFB with ASU and CO2 Capture													
Power Generation															
Net Output	kW	134,514	134,514	134,514	134,514	134,514	134,514	134,514	134,514	134,514	134,514	134,514	134,514	134,514	134,514
Availability Factor	%	80	60	70	90	100	80	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%
Net Plant Heat Rate, HHV	Btu / kWh	13,546	13,546	13,546	13,546	13,546	13,546	13,546	13,546	13,546	13,546	13,546	13,546	13,546	13,546
Net Generation	MWh / year	942,674	707,006	824,840	1,060,508	1,178,343	942,674	942,674	942,674	942,674	942,674	942,674	942,674	942,674	942,674
Costs															
EPC Price	\$ / kW	2,443	2,443	2,443	2,443	2,443	1,832	2,137	2,748	3,053	2,443	2,443	2,443	2,443	2,443
EPC Price	\$1000s	328,589	328,589	328,589	328,589	328,589	246,442	287,515	369,663	410,736	328,589	328,589	328,589	328,589	328,589
Fixed O&M Costs	\$1000 / year	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854
Fixed O&M Costs	\$ / kW	58.39	58.39	58.39	58.39	58.39	58.39	58.39	58.39	58.39	58.39	58.39	58.39	58.39	58.39
Variable O&M Costs	\$1000 / year	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820
Variable O&M Costs	cents / kWh	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94
Total O&M Costs	cents / kWh	1.77	1.77	1.77	1.77	1.77	1.77	1.77	1.77	1.77	1.77	1.77	1.77	1.77	1.77
Credits															
CO2 Price	\$ / ton	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Fuel Cost Calculation															
Coal Price	\$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	0.94	1.09	1.41	1.56	
CO2 Emissions															
CO2 Produced	lbm / h	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995
CO2 Produced	lbm / kWh	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85
CO2 Emitted	lbm / h	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618
CO2 Emitted	lbm / kWh	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19
Financing Assumptions															
Equity	%	50	50	50	50	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	20	20	20	20	20	20	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	10	10	10	10	10	10
Levelized Cost of Electricity (cents / kWh)															
Financial Component		4.7	6.2	5.3	4.1	3.7	3.5	4.1	5.2	5.7	4.7	4.7	4.7	4.7	4.7
Fixed O&M		0.8	1.1	1.0	0.7	0.7	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8
Variable O&M		0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9
Fuel		1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.3	1.5	1.9	2.1	2.1
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		8.1	10.0	8.9	7.5	7.0	7.0	7.6	8.7	9.2	7.7	7.9	8.3	8.5	8.5
CO2 Mitigation															
CO2 Mitigation Cost	\$ / ton	40	60	48	33	28	27	33	46	52	35	37	42	44	

Parameter	Units	Case 2 - Oxygen-Fired CFB with ASU and CO2 Capture													
Power Generation															
Net Output	kW	134,514	134,514	134,514	134,514	134,514	134,514	134,514	134,514	134,514	134,514	134,514	134,514	134,514	134,514
Availability Factor	%	80	80	80	80	80	80	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%	25.2%
Net Plant Heat Rate, HHV	Btu / kWh	13,546	13,546	13,546	13,546	13,546	13,546	13,546	13,546	13,546	13,546	13,546	13,546	13,546	13,546
Net Generation	MWh / year	942,674	942,674	942,674	942,674	942,674	942,674	942,674	942,674	942,674	942,674	942,674	942,674	942,674	942,674
Costs															
EPC Price	\$ / kW	2,443	2,443	2,443	2,443	2,443	2,443	2,443	2,443	2,443	2,443	2,443	2,443	2,443	2,443
EPC Price	\$1000s	328,589	328,589	328,589	328,589	328,589	328,589	328,589	328,589	328,589	328,589	328,589	328,589	328,589	328,589
Fixed O&M Costs	\$1000 / year	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854	7,854
Fixed O&M Costs	\$ / kW	58.39	58.39	58.39	58.39	58.39	58.39	58.39	58.39	58.39	58.39	58.39	58.39	58.39	58.39
Variable O&M Costs	\$1000 / year	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820	8,820
Variable O&M Costs	cents / kWh	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94
Total O&M Costs	cents / kWh	1.77	1.77	1.77	1.77	1.77	1.77	1.77	1.77	1.77	1.77	1.77	1.77	1.77	1.77
Credits															
CO2 Price	\$ / ton	0	0	0	0	0	0	0	0	0	0	0	0	10	20
Fuel Cost Calculation															
Coal Price	\$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25
CO2 Emissions															
CO2 Produced	lbm / h	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995
CO2 Produced	lbm / kWh	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85
CO2 Emitted	lbm / h	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618
CO2 Emitted	lbm / kWh	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19
Financing Assumptions															
Equity	%	38	44	56	63	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	15	18	23	25	20	20	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	8	9	11	13	10	10
Levelized Cost of Electricity (cents / kWh)															
Financial Component		4.2	4.5	4.8	5.0	4.5	4.6	4.7	4.8	3.9	4.3	5.1	5.5	4.7	4.7
Fixed O&M		0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8
Variable O&M		0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9
Fuel		1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-1.3	-2.6
Total		7.7	7.9	8.3	8.4	8.0	8.0	8.2	8.3	7.4	7.7	8.5	8.9	6.8	5.5
CO2 Mitigation															
CO2 Mitigation Cost	\$ / ton	35	37	42	43	38	39	41	41	31	35	44	49	25	11

Updated Case 2 - Oxygen-Fired CFB with Air Separation Unit and CO₂ Capture

Results for the Updated Case 2 COE sensitivity study are shown in Figure 7-6 and Table 7-5. The levelized COE for the base parameter values is 7.9 cents per kWh. Levelized COE ranges from a low of 5.4 to a high of 9.7 cents per kWh. CO₂ mitigation costs ranged from 10 to 57 \$/ton of CO₂ avoided (reference plant is Case 1) with the baseline value at 37 \$/ton of CO₂ avoided.

Figure 7-6: Updated Case 2 - Oxygen-Fired CFB with ASU and CO₂ Capture - Economic Sensitivity Analysis Results

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Table 7-5: Updated Case 2 - Oxygen-Fired CFB with ASU and CO₂ Capture - Economic Sensitivity Analysis Results

Parameter	Units	Case 2 Updated - Oxygen-Fired CFB with ASU and CO2 Capture													
Power Generation															
Net Output	kW	138,402	138,402	138,402	138,402	138,402	138,402	138,402	138,402	138,402	138,402	138,402	138,402	138,402	138,402
Availability Factor	%	80	60	70	90	100	80	80	80	80	80	80	80	80	80
Actual Operating Hours	hours / year	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008	7,008
Net Efficiency, HHV	%	25.9%	25.9%	25.9%	25.9%	25.9%	25.9%	25.9%	25.9%	25.9%	25.9%	25.9%	25.9%	25.9%	25.9%
Net Plant Heat Rate, HHV	Btu / kWh	13,152	13,152	13,152	13,152	13,152	13,152	13,152	13,152	13,152	13,152	13,152	13,152	13,152	13,152
Net Generation	MWh / year	969,921	727,441	848,681	1,091,161	1,212,402	969,921	969,921	969,921	969,921	969,921	969,921	969,921	969,921	969,921
Costs															
EPC Price	\$ / kW	2,382	2,382	2,382	2,382	2,382	1,786	2,084	2,679	2,977	2,382	2,382	2,382	2,382	2,382
EPC Price	\$1000s	329,610	329,610	329,610	329,610	329,610	247,208	288,409	370,811	412,013	329,610	329,610	329,610	329,610	329,610
Fixed O&M Costs	\$1000 / year	7,859	7,859	7,859	7,859	7,859	7,859	7,859	7,859	7,859	7,859	7,859	7,859	7,859	7,859
Fixed O&M Costs	\$ / kW	56.78	56.78	56.78	56.78	56.78	56.78	56.78	56.78	56.78	56.78	56.78	56.78	56.78	56.78
Variable O&M Costs	\$1000 / year	8,835	6,626	7,731	9,939	11,044	8,835	8,835	8,835	8,835	8,835	8,835	8,835	8,835	8,835
Variable O&M Costs	cents / kWh	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91	0.91
Total O&M Costs	cents / kWh	1.72	1.99	1.84	1.63	1.56	1.72	1.72	1.72	1.72	1.72	1.72	1.72	1.72	1.72
Credits															
CO2 Price	\$ / ton	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Fuel Cost Calculation															
Coal Price	\$ / MMBtu	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	1.25	0.94	1.09	1.41	1.56	1.56
CO2 Emissions															
CO2 Produced	lbm / h	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995	376,995
CO2 Produced	lbm / kWh	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85	2.85
CO2 Emitted	lbm / h	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618	24,618
CO2 Emitted	lbm / kWh	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19	0.19
Financing Assumptions															
Equity	%	50	50	50	50	50	50	50	50	50	50	50	50	50	50
Corporate Tax	%	20	20	20	20	20	20	20	20	20	20	20	20	20	20
Discount Factor	%	10	10	10	10	10	10	10	10	10	10	10	10	10	10
Levelized Cost of Electricity (cents / kWh)															
Financial Component		4.5	6.1	5.2	4.0	3.6	3.4	4.0	5.1	5.6	4.5	4.5	4.5	4.5	4.5
Fixed O&M		0.8	1.1	0.9	0.7	0.6	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8
Variable O&M		0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9
Fuel		1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.2	1.4	1.8	2.1	2.1
CO2 Credit		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		7.9	9.7	8.7	7.3	6.8	6.8	7.4	8.4	9.0	7.5	7.7	8.1	8.3	8.3
CO2 Mitigation															
CO2 Mitigation Cost	\$ / ton	37	57	46	31	26	25	31	43	49	33	35	40	42	42

[illegible]

8 SUMMARY AND RECOMMENDATIONS

Summary of Pilot Testing

Pilot-scale testing of the oxygen-fired CFB concept was performed at ALSTOM's 9.9 MMBtu/hr Multi-use Test Facility (MTF), located in Windsor, Connecticut. Key results from the testing are summarized below.

- The furnace was successfully operated on bituminous coal and petcoke in O₂/CO₂ combustion mediums containing local oxygen enrichments of up to 70%. There was no evidence of particle agglomeration or defluidization in the furnace.
- Because of the high CO₂ content of the flue gas, the furnace operated above 1650°F to ensure calcination of the limestone for sulfur capture. In regions where the temperature was much cooler, there was evidence of recarbonation.
- Because of the high temperature, the sulfur emissions from the bituminous coal were higher than normal for the same limestone feed rate. For pet coke, the optimum temperature for sulfur capture is higher, so the oxygen-fired emissions were very low.
- NO_x emissions were lower with oxygen firing. This is due to the reduction in prompt NO_x, and perhaps some reduction in thermal NO_x and fuel NO_x.
- Carbon monoxide emissions were higher with oxygen firing. This is likely due to the high CO₂ content of the flue gas, which hinders oxidation of the CO.
- Carbon heat loss in the fly ash was lower than, or comparable to, the levels with air firing. The carbon loss was lower for pet coke than for bituminous coal.
- There was no significant difference in heat transfer to the furnace waterwall test sections between air and oxygen firing. This heat transfer is dominated by solids effects, which do not depend on the gas composition.

These results are largely as expected based upon earlier test results, and did not identify any major technical showstoppers to the oxygen-fired CFB concept.

Summary of Design and Cost Updates

In a previous study (Marion, et al., 2003) ALSTOM compared the performance and economics of 13 combustion and gasification cases with and without CO₂ capture. Cases 1 and 2 were

Case-1: Air Fired CFB without CO₂ Capture (Base Case) Conventional air-fired CFB without CO₂ capture using 1,800 psig / 1,000°F / 1,000°F steam cycle.

Case-2: New Compact Oxygen-Fired CFB with CO₂ Capture, Purification, Compression and Liquefaction. Nearly identical steam cycle as Case-1 (same main and reheat steam conditions - flow, temperature, and pressure) but smaller Boiler Island equipment than Case-1. Oxygen is supplied from a Cryogenic Air Separation Unit (ASU) Plant. The CFB Boiler Island provides a concentrated CO₂ flue gas product stream to the Gas Processing System.

In the present study, Case-2 was updated to incorporate the results from the MTF pilot tests. This new design is Updated Case-2. The major changes in the design were the addition of the FDA system and some changes in the low-level heat recovery, especially

in the ASU.

- The plant efficiency of Updated Case-2 was 26.0 percent compared to 25.2 percent for Original Case-2. The heat penalty due to the FDA was more than compensated for by the improvement in low-level heat recovery. The efficiency for Case-1 was 35.5%.
- The capital cost of the Updated Case-2 was 2320 \$/kW(net) compared to 2443 \$/kW(net) for Original Case-2 - about a 5 percent decrease. The cost for Case-1 was 1304 \$/kW(net).
- Levelized cost of electricity (COE) is about 4.5 cents/kWh without CO₂ capture (Case-1). Original Case-2 had a levelized COE of about 8.1 cents/kWh with the addition of the CO₂ capture system (incremental COE of 3.6 cents/kWh). The Updated Case-2 has a levelized COE of about 7.9 cents/kWh (incremental COE of 3.4 cents/kWh). This 2.5% decrease in levelized COE for Updated Case-2 is due to a small decrease in the specific plant cost and a small increase in the net plant output.
- CO₂ mitigation costs are about 40 and 37 \$/ton of CO₂ avoided for Original Case-2 and Updated Case-2, respectively.

Recommendations

To prepare for a commercial demonstration of the oxygen-fired CFB concept, there remain several items to be investigated:

- the operation and performance of the moving bed heat exchanger (MBHE) to avoid recarbonation and also for cost savings compared to the standard fluid bed heat exchanger (FBHE);
- performance of the back-end Flash Dryer Absorber (FDA) for sulfur capture under high CO₂ / high moisture flue gas environment using calcined limestone in the fly ash and using fresh commercial lime directly in the FDA;
- determine the effect of recarbonation on fouling in the convective pass;
- assess the impact of oxygen firing on the mercury, trace elements, and VOC emissions;
- update the technical and economic analysis of the oxygen-fired CFB based on these further results; and
- develop a proposal-level design for a commercial demonstration of the oxygen-fired CFB concept.

ALSTOM has proposed to the DOE a project to address just these issues with further MTF pilot testing (at the full 9.9 MMBtu/hr firing rate) and subsequent update of the technical and economic analyses. This project has been selected and awarded by the DOE - Commercialization Development of Oxygen Fired CFB for Greenhouse Gas Control, Cooperative Agreement DE-FC26-04NT42205. The performance period for this project is from 1 October 2004 to 30 September 2005. In parallel with this new project, ALSTOM is actively seeking partners for the next step - a commercial demonstration.

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10 APPENDIX I: Detailed Power Plant Cost Estimates

Appendix I provides detailed investment cost breakdowns and operating & maintenance costs for each of the three plants studied. Costs for Case-1 and Original Case-2 are copied from (Marion et al., 2003). The costs are grouped into four separate areas: Boiler Island, Air Separation, Gas Processing, and Balance of Plant. Some of the cases do not have equipment in all four areas.

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CIRCULATING FLUIDIZED BED BOILERS – PHASE II

10.1 Case-1: Detailed Investment Costs and Operating and Maintenance Costs (copied from Marion et al., 2003)

Table 10-1: Case-1 Overall Power Plant Investment Costs

Client: ALSTOM Power Inc. Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers										Report Date: 4/18/2003	
TOTAL PLANT COST SUMMARY											
Case 1 - 1x200 MW Air-Fired CFB w/o CO2 Capture											
Net Output Power, kW				193,037		Estimate Type: Conceptual		Cost Base: Jul-03		(\$x1000)	
Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost	
				Direct	Indirect					(\$x1000)	\$/kW
1	FUEL & SORBENT HANDLING	6,760	1,657	3,765	-	-	12,182	1,463	548	14,193	74
2	FUEL & SORBENT PREP. & FEED	3,588	202	1,039	-	-	4,829	580	217	5,626	29
3	FEEDWATER & MISC. BOP SYSTEMS	11,721	-	5,757	-	-	17,478	2,097	786	20,361	105
4	FLUIDIZED BED BOILER										
	4.1 Fluidized Bed Boiler w/o Bhse. & Accessories									75,658	392
	4.2 Open										
	4.4 Open										
4.4 - 4.9	Boiler BOP (Fluidizing Air Fans)	2,502	-	718	-	-	3,220	386	145	3,751	19
	SUBTOTAL 4	2,502	-	718	-	-	3,220	386	145	79,409	411
5	FLUE GAS CLEANUP	3,873	-	3,664	-	-	7,537	905	339	8,781	45
6	COMBUSTION TURBINE ACCESSORIES										
	6.1 Combustion Turbine Generator										
	6.2-6.9 Combustion Turbine Accessories										
	SUBTOTAL 6	-	-	-	0	0	-	-	-	-	0
7	HRSG DUCTING & STACK										
	7.1 Heat Recovery Steam Generator										
7.2-7.9	ID Fans, Ductwork and Stack	6,851	415	4,155	-	0	11,421	1,370	515	13,306	69
	SUBTOTAL 7	6,851	415	4,155	-	-	11,421	1,370	515	13,306	69
8	STEAM TURBINE GENERATOR										
	8.1 Steam TG & Accessories	16,242	-	2,330	0	0	18,572	2,229	836	21,637	112
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	8,706	324	4,719	0	0	13,749	1,649	619	16,017	83
	SUBTOTAL 8	24,948	324	7,049	-	-	32,321	3,878	1,455	37,654	195
9	COOLING WATER SYSTEM	3,767	2,146	4,246	-	-	10,159	1,219	457	11,835	61
10	ASH/SPENT SORBENT HANDLING SYSTEMS	4,499	157	3,204	-	-	7,860	943	354	9,157	47
11	ACCESSORY ELECTRIC PLANT	5,545	2,366	7,661	-	-	15,572	1,869	700	18,141	94
12	INSTRUMENTATION & CONTROL	4,731	-	4,837	-	-	9,568	1,047	430	11,045	57
13	IMPROVEMENT TO SITE	1,389	798	2,355	-	-	4,542	545	204	5,291	27
14	BUILDINGS & STRUCTURES	-	7,307	7,290	-	-	14,597	1,750	658	17,005	88
	TOTAL COST	80,174	15,372	55,740	-	-	151,286	18,052	6,808	251,804	1,304

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
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Client: ALSTOM Power Inc. Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers										Report Date: 4/18/2003	
TOTAL PLANT COST SUMMARY											
Case 1 - 1x200 MW Air-Fired CFB w/o CO2 Capture											
Net Output Power, kW 193,037 Estimate Type: Conceptual Cost Base: Jul-03 (\$x1000)											
Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost	
				Direct	Indirect					\$	\$/kW
1	FUEL & SORBENT HANDLING										
1.1	Coal Receive & Unload	1,179		565			1,744	209	78	2,031	11
1.2	Fuel Stackout and Reclaim	1,524		398			1,922	231	87	2,240	12
1.3	Fuel Conveyors	1,417		358			1,775	213	80	2,068	11
1.4	Other Fuel Handling	371		83			454	54	20	528	3
1.5	Sorbent Receive & Unload	86		27			113	14	5	132	1
1.6	Sorbent Stackout and Reclaim	1,389		267			1,656	199	74	1,929	10
1.7	Sorbent Conveyors	495	100	127			722	87	33	842	4
1.8	Other Sorbent Handling	299	66	165			530	64	24	618	3
1.9	Fuel & Sorbent Hnd. Foundations		1,491	1,775			3,266	392	147	3,805	20
	SUBTOTAL. 1	6,760	1,657	3,765	-	-	12,182	1,463	548	14,193	74
2	FUEL & SORBENT PREP. & FEED										
2.1	Coal Crushing & Drying	652		133			785	94	35	914	5
2.2	Fuel Conveyor to Storage	2,086		477			2,563	308	115	2,986	15
2.3	Fuel Injection System						-			-	-
2.4	Misc. Fuel Prep. & Feed						-			-	-
2.5	Sorbent Prep. Equipment	567		124			691	83	31	805	4
2.6	Sorbent Storage & Feed	283		113			396	48	18	462	2
2.7	Sorbent Injection System						-			-	-
2.8	Booster Air Supply System						-			-	-
2.9	Fuel & Sorbent Feed. Foundations		202	192			394	47	18	459	2
	SUBTOTAL. 2	3,588	202	1,039	-	-	4,829	580	217	5,626	29
3	FEEDWATER & MISC. BOP SYSTEMS										
3.1	Feedwater System	3,591		1,360			4,951	594	223	5,768	30
3.2	Water Makeup & Pretreating	1,468		452			1,920	230	86	2,236	12
3.3	Other Feedwater Subsystems	2,053		903			2,956	355	133	3,444	18
3.4	Service Water System	282		150			432	52	19	503	3
3.5	Other Boiler Plant Systems	1,974		1,783			3,757	451	169	4,377	23
3.6	FO Supply System & Nat. Gas	88		112			200	24	9	233	1
3.7	Waste Treatment Equipment	1,075		543			1,618	194	73	1,885	10
3.8	Misc. Equip. (Cranes, AirComp.Comm.)	1,190		454			1,644	197	74	1,915	10
	SUBTOTAL. 3	11,721	-	5,757	-	-	17,478	2,097	786	20,361	105
4	FLUIDIZED BED BOILER										
4.1	Fluidized Bed Boiler, w/o Bhse & Accessories						-			75,658	392
4.2	Open						-			-	-
4.3	Open						-			-	-
4.4	Boiler BoP (Fluidizing Air Fans)	365		105			470	56	21	547	3
4.5	Primary Air System (Fans)	1,343		385			1,728	207	78	2,013	10
4.6	Secondary Air System (Fans)	794		228			1,022	123	46	1,191	6
4.7	Major Component Rigging						-			-	-
4.8	Boiler Foundation						-			-	-
	SUBTOTAL. 4	2,502	-	718	-	-	3,220	386	145	79,409	411

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

Client: ALSTOM Power Inc. Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers										Report Date: 4/18/2003	
TOTAL PLANT COST SUMMARY											
Case 1 - 1x200 MW Air-Fired CFB w/o CO2 Capture											
Net Output Power, kW 193,037 Estimate Type: Conceptual Cost Base: Jul-03 (\$x1000)											
Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost	
				Direct	Indirect					\$	\$/kW
5	FLUE GAS CLEANUP										
	5.1 Absorber Vessels & Accessories						-			-	-
	5.2 Other FGD						-			-	-
	5.3 Bag House & Accessories	3,500		3,289			6,789	815	305	7,909	41
	5.4 Other Particulate Removal Materials	373		375			748	90	34	872	5
	5.5 Gypsum Dewatering System						-			-	-
	5.6 Mercury Removal System						-			-	-
	5.9 Open						-			-	-
	SUBTOTAL. 5	3,873	-	3,664	-	-	7,537	905	339	8,781	45
6	COMBUSTION TURBINE ACCESSORIES						-			-	-
	6.1 Combustion Turbine Generator						-			-	-
	6.2 Combustion Turbine Accessories						-			-	-
	6.3 Compressed Air Piping						-			-	-
	6.9 Combustion Turbine Foundations						-			-	-
	SUBTOTAL. 6	-	-	-	-	-	-			-	-
7	HRSG DUCTING & STACK						-			-	-
	7.1 Heat Recovery Steam Generator						-			-	-
	7.2 ID Fans	1,408		404			1,812	217	82	2,111	11
	7.3 Ductwork	1,818		1,309			3,127	375	141	3,643	19
	7.4 Stack	3,625		1,997			5,622	675	253	6,550	34
	7.9 Duct & Stack Foundations		415	445			860	103	39	1,002	5
	SUBTOTAL. 7	6,851	415	4,155	-	-	11,421	1,370	515	13,306	69
8	STEAM TURBINE GENERATOR						-			-	-
	8.1 Steam TG & Accessories	16,242		2,330			18,572	2,229	836	21,637	112
	8.2 Turbine Plant Auxiliaries	112		226			338	40	15	393	2
	8.3 Condenser & Auxiliaries	2,880		694			3,574	429	161	4,164	22
	8.4 Steam Piping	5,714		3,277			8,991	1,079	405	10,475	54
	8.9 TG Foundations		324	522			846	101	38	985	5
	SUBTOTAL. 8	24,948	324	7,049	-	-	32,321	3,878	1,455	37,654	195
9	COOLING WATER SYSTEM						-			-	-
	9.1 Cooling Towers	2,642		1,336			3,978	477	179	4,634	24
	9.2 Circulating Water Pumps	665		55			720	86	32	838	4
	9.3 Circulating Water System Auxiliaries	178		22			200	24	9	233	1
	9.4 Circulating Water Piping		1,314	1,286			2,600	312	117	3,029	16
	9.5 Make-up Water System	151		197			348	42	16	406	2
	9.6 Component Cooling Water System	131		101			232	28	10	270	1
	9.9 Circ. Water System Foundations Structures		832	1,249			2,081	250	94	2,425	13
	SUBTOTAL. 9	3,767	2,146	4,246	-	-	10,159	1,219	457	11,835	61
10	ASH/SPENT SORBENT HANDLING SYSTEMS						-			-	-
	10.1 Ash Coolers						-			-	-
	10.2 Cyclone Ash Letdown						-			-	-
	10.3 HGPU Ash Letdown						-			-	-
	10.4 High Temperature Ash Piping						-			-	-
	10.5 Other Ash Recovery Equipment						-			-	-
	10.6 Ash Storage Silos	331		961			1,292	155	58	1,505	8
	10.7 Ash Transport & Feed Equipment	4,168		2,068			6,236	748	281	7,265	38
	10.8 Misc. Ash Handling Equipment						-			-	-
	10.9 Ash/Spent Sorbent Foundations		157	175			332	40	15	387	2
	SUBTOTAL. 10	4,499	157	3,204	-	-	7,860	943	354	9,157	47

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

Client: ALSTOM Power Inc. Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers										Report Date: 4/18/2003	
TOTAL PLANT COST SUMMARY											
Case 1 - 1x200 MW Air-Fired CFB w/o CO2 Capture											
Net Output Power, kW 193,037 Estimate Type: Conceptual Cost Base: Jul-03 (\$x1000)											
Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost	
				Direct	Indirect					\$	\$/kW
11	ACCESSORY ELECTRIC PLANT										
11.1	Generator Equipment	674		91			765	92	34	891	5
115.2	Station Service Equipment	1,497		411			1,908	229	86	2,223	12
11.3	Switchgear & Motor Control	1,722		244			1,966	236	88	2,290	12
11.4	Conduit & Cable Tray		1,038	3,115			4,153	498	187	4,838	25
11.5	Wire & Cable		1,226	3,282			4,508	541	203	5,252	27
11.6	Protective Equipment	81		230			311	37	14	362	2
11.7	Standby Equipment	602		11			613	74	28	715	4
11.8	Main Power Transformer	969		38			1,007	121	45	1,173	6
11.9	Electrical Foundations		102	239			341	41	15	397	2
	SUBTOTAL. 11	5,545	2,366	7,661	-	-	15,572	1,869	700	18,141	94
12	INSTRUMENTATION & CONTROL										
12.1	PC Control Equipment						-			-	-
12.2	Combustion Turbine Control						-			-	-
12.3	Steam Turbine Control						-			-	-
12.4	Other Major Component Control						-			-	-
12.5	Signal Processing Equipment						-			-	-
12.6	Control Boards, Panels & Racks	240		120			360	43	16	419	2
12.7	Distributed Control System Equipment	2,662		388			3,050	266	137	3,453	18
12.8	Instrument Wiring & tubing	1,145		3,034			4,179	501	188	4,868	25
12.9	Other I & C Equipment	684		1,295			1,979	237	89	2,305	12
	SUBTOTAL. 12	4,731	-	4,837	-	-	9,568	1,047	430	11,045	57
13	IMPROVEMENT TO SITE										
13.1	Site Preparation		23	393			416	50	19	485	3
13.2	Site Improvement		775	810			1,585	190	71	1,846	10
13.3	Site Facilities	1,389		1,152			2,541	305	114	2,960	15
	SUBTOTAL. 13	1,389	798	2,355	-	-	4,542	545	204	5,291	27
14	BUILDINGS & STRUCTURES										
14.1	FB Boiler Building Foundation		2,188	1,940			4,128	495	186	4,809	25
14.2	Turbine Building		3,257	3,060			6,317	758	284	7,359	38
14.3	Administration Building		415	442			857	103	39	999	5
14.4	Circulation Water Pumphouse		89	71			160	19	7	186	1
14.5	Water Treatment Building		289	240			529	63	24	616	3
14.6	Machine Shop		370	251			621	74	28	723	4
14.7	Warehouse		251	254			505	61	23	589	3
14.8	Other Buildings & Structures		154	132			286	34	13	333	2
14.9	Waste Treating Building & Structure		294	900			1,194	143	54	1,391	7
	SUBTOTAL. 14	-	7,307	7,290	-	-	14,597	1,750	658	17,005	88

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

Table 10-2: Case-1 Overall Power Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.		INITIAL & ANNUAL O&M EXPENSES		Cost Base: Jul-03	
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers		Case 1 - 1x200 MW Air-Fired CFB w/o CO2 Capture		Net Plant Heat Rate (Btu/kWh): 9,611	
				Net Power Output (kW): 193,037	
				Capacity Factor (%): 80	
OPERATING & MAINTENANCE LABOR					
<u>Operating Labor</u>					
Operating Labor Rate (Base):		30.90	\$ / hour		
Operating Labor Burden:		30.00	%		
Labor O-H change Rate:		25.00	%		
Operating Labor Requirements (O.J.) per shift		<u>1 unit/mod.</u>	<u>Total Plant</u>		
Skilled Operator		1.0	1.0		
Operator		7.0	7.0		
Foreman		1.0	1.0		
Lab Tech's, etc.		1.0	1.0		
TOTAL O.J.'s		10.0	10.0		
				Annual Cost	Annual Unit Cost
				<u>\$ / year</u>	<u>\$/kW-net</u>
Annual Operating Labor Costs (calc'd)				3,518,892	18.23
Maintenance Labor Costs (calc'd)				1,007,216	5.22
Administrative & Support Labor (calc'd)				1,131,527	5.86
TOTAL FIXED OPERATING COSTS				5,657,635	29.31
					<u>\$/kWh-net</u>
Maintenance Material Cost (calc'd)				1,208,659	0.00089
<u>Consumables</u>		<u>Consumption</u>	<u>Unit</u>	<u>Initial</u>	
		<u>Initial</u>	<u>Per Day</u>	<u>Cost</u>	<u>Cost</u>
Water (1000 gallons)			2,864	1.00	836,288 0.00062
Chemicals					
MU & WT Chem. (lbs.)		415,979	13,866	0.16	66,557 0.00048
Limestone (ton)		10,435	347.8	10.00	104,350 0.00075
Formic Acid (lbs.)				0.60	
Ammonia, NH3 (ton)				220	
Subtotal Chemicals				170,907	1,663,396 0.0012
Other Consumables					
Supplemental Fuel (MBtu)					
SCR Catalyst Replacement (MBtu)					
Emissions Penalties					
Subtotal Other					
Waste Disposal					
Fly Ash & Bottom Ash (ton)			804.3	8.00	1,878,845 0.0014
Subtotal Solid Waste Disposal					1,878,845 0.0014
By-Products & Emissions					
Gypsum (ton)					
Subtotal By-Products					
TOTAL VARIABLE OPERATING COST				5,587,188	0.0041

10.2 Original Case-2: Detailed Investment Costs and Operating and Maintenance Costs (copied from Marion et al., 2003)

Table 10-3: Original Case-2 Overall Power Plant Investment Costs

Client: ALSTOM Power Inc. Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers										Report Date: 6/20/2003
TOTAL PLANT COST SUMMARY Case 2 - 1x200 gr. MW O₂-Fired CFB w/ASU & CO₂ Capture										
Net Output Power, kW 134,514		Estimate Type: Conceptual		Cost Base: Jul-03 (\$x1000)						
Acct. No.	Account Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost \$/kW
1	FUEL & SORBENT HANDLING	6,647	1,630	3,703	-	-	11,980	1,436	539	13,955 104
2	FUEL & SORBENT PREP. & FEED	3,524	198	1,021	-	-	4,743	569	214	5,526 41
3	FEEDWATER & MISC. BOP SYSTEMS	12,319	-	5,978	-	-	18,297	2,196	824	21,317 158
4	FLUIDIZED BED BOILER									
4.1	Fluidized Bed Boiler w/o Bhse. & Accessories									50,036 372
4.2	Air Separation Unit (ASU)									64,000 476
4.4	Open									
4.4 - 4.9	Boiler BOP (Fluidizing Air Fans)	170	-	49	-	-	219	26	10	255 2
	SUBTOTAL 4	170	-	49	-	-	219	26	10	114,291 850
5	FLUE GAS CLEANUP									
5.1-5.8	Miscellaneous	1,328	-	1,260	-	-	2,588	311	116	3,015 22
5.9	Gas Processing System (GPS)									57,122 425
	SUBTOTAL 5	1,328	-	1,260	-	-	2,588	311	116	60,137 447
6	COMBUSTION TURBINE ACCESSORIES									
6.1	Combustion Turbine Generator									
6.2-6.9	Combustion Turbine Accessories									
	SUBTOTAL 6	-	-	-	0	0	-	-	-	- 0
7	HRSG DUCTING & STACK									
7.1	Heat Recovery Steam Generator									
7.2-7.9	ID Fans, Ductwork and Stack	514	10	305	-	0	829	99	37	965 7
	SUBTOTAL 7	514	10	305	-	-	829	99	37	965 7
8	STEAM TURBINE GENERATOR									
8.1	Steam TG & Accessories	16,296	-	2,338	0	0	18,634	2,236	839	21,709 161
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	8,734	325	4,734	0	0	13,793	1,655	620	16,068 119
	SUBTOTAL 8	25,030	325	7,072	-	-	32,427	3,891	1,459	37,777 281
9	COOLING WATER SYSTEM	3,777	2,152	4,258	-	-	10,187	1,223	459	11,869 88
10	ASH/SPENT SORBENT HANDLING SYSTEMS	3,935	137	2,801	-	-	6,873	825	347	8,045 60
11	ACCESSORY ELECTRIC PLANT	7,444	2,544	8,736	-	-	18,724	2,247	844	21,815 162
12	INSTRUMENTATION & CONTROL	4,475	-	4,577	-	-	9,052	1,086	407	10,545 78
13	IMPROVEMENT TO SITE	1,392	800	2,360	-	-	4,552	547	205	5,304 39
14	BUILDINGS & STRUCTURES	-	7,323	7,306	-	-	14,628	1,756	659	17,043 127
	TOTAL COST	70,555	15,119	49,426	-	-	135,099	16,212	6,120	328,589 2,443

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

Client: ALSTOM Power Inc.										Report Date: 6/20/2003	
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers											
TOTAL PLANT COST SUMMARY											
Case 2 - 1x200 gr. MW O2-Fired CFB w/ASU & CO2 Capture											
Net Output Power, kW				134,514		Estimate Type: Conceptual		Cost Base: Jul-03		(\$x1000)	
Acct. No.	Account Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost	
				Direct	Indirect					\$	\$/kW
1	FUEL & SORBENT HANDLING										
1.1	Coal Receive & Unload	1,160		555			1,715	206	77	1,998	15
1.2	Fuel Stackout and Reclaim	1,499		392			1,891	227	85	2,203	16
1.3	Fuel Conveyors	1,393		352			1,745	209	79	2,033	15
1.4	Other Fuel Handling	365		82			447	54	20	521	4
1.5	Sorbent Receive & Unload	84		27			111	13	5	129	1
1.6	Sorbent Stackout and Reclaim	1,365		262			1,627	195	73	1,895	14
1.7	Sorbent Conveyors	487	99	125			711	85	32	828	6
1.8	Other Sorbent Handling	294	65	162			521	62	23	606	5
1.9	Fuel & Sorbent Hnd. Foundations		1,466	1,746			3,212	385	145	3,742	28
	SUBTOTAL. 1	6,647	1,630	3,703	-	-	11,980	1,436	539	13,955	104
2	FUEL & SORBENT PREP. & FEED										
2.1	Coal Crushing & Drying	640		131			771	93	35	899	7
2.2	Fuel Conveyor to Storage	2,049		469			2,518	302	113	2,933	22
2.3	Fuel Injection System						-			-	-
2.4	Misc. Fuel Prep. & Feed						-			-	-
2.5	Sorbent Prep. Equipment	557		121			678	81	31	790	6
2.6	Sorbent Storage & Feed	278		111			389	47	18	454	3
2.7	Sorbent Injection System						-			-	-
2.8	Booster Air Supply System						-			-	-
2.9	Fuel & Sorbent Feed. Foundations		198	189			387	46	17	450	3
	SUBTOTAL. 2	3,524	198	1,021	-	-	4,743	569	214	5,526	41
3	FEEDWATER & MISC. BOP SYSTEMS										
3.1	Feedwater System	3,591		1,360			4,951	594	223	5,768	43
3.2	Water Makeup & Pretreating	1,798		553			2,351	282	106	2,739	20
3.3	Other Feeddwater Subsystems	2,053		903			2,956	355	133	3,444	26
3.4	Service Water System	345		183			528	63	24	615	5
3.5	Other Boiler Plant Systems	1,934		1,747			3,681	442	166	4,289	32
3.6	FO Supply System & Nat. Gas	89		112			201	24	9	234	2
3.7	Waste Treatment Equipment	1,316		664			1,980	238	89	2,307	17
3.8	Misc. Equip. (Cranes, AirComp.Comm.)	1,193		456			1,649	198	74	1,921	14
	SUBTOTAL. 3	12,319	-	5,978	-	-	18,297	2,196	824	21,317	158
4	FLUIDIZED BED BOILER										
4.1	Fluidized Bed Boiler, w/o Bhse & Accessories						-			50,036	372
4.2	Air Separation Unit (ASU)						-			64,000	476
4.3	Open						-			-	-
4.4	Boiler BoP (Fluidizing Air Fans)	34		10			44	5	2	51	0
4.5	Primary Air System (Fans)	136		39			175	21	8	204	2
4.6	Secondary Air System (Fans)						-			-	-
4.7	Major Component Rigging						-			-	-
4.8	Boiler Foundation						-			-	-
	SUBTOTAL. 4	170	-	49	-	-	219	26	10	114,291	850

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

Client: ALSTOM Power Inc. Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers										Report Date: 6/20/2003	
TOTAL PLANT COST SUMMARY											
Case 2 - 1x200 gr. MW O2-Fired CFB w/ASU & CO2 Capture											
Net Output Power, kW				134,514		Estimate Type: Conceptual		Cost Base: Jul-03		(\$x1000)	
Acct. No.	Account Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost	
				Direct	Indirect					\$	\$/kW
5	FLUE GAS CLEANUP										
5.1	Absorber Vessels & Accessories						-			-	-
5.2	Other FGD						-			-	-
5.3	Bag House & Accessories	1,150		1,081			2,231	268	100	2,599	19
5.4	Other Particulate Removal Materials	178		179			357	43	16	416	3
5.5	Gypsum Dewatering System						-			-	-
5.6	Mercury Removal System						-			-	-
5.1-5.6		1,328	-	1,260	-	-	2,588	311	116	3,015	22
5.9	Gas Processing System (GPS)									57,122	425
	SUBTOTAL. 5	1,328	-	1,260	-	-	2,588	311	116	60,137	447
6	COMBUSTION TURBINE ACCESSORIES						-			-	-
6.1	Combustion Turbine Generator						-			-	-
6.2	Combustion Turbine Accessories						-			-	-
6.3	Compressed Air Piping						-			-	-
6.9	Combustion Turbine Foundations						-			-	-
	SUBTOTAL. 6	-	-	-	-	-	-			-	-
7	HRSG DUCTING & STACK						-			-	-
7.1	Heat Recovery Steam Generator						-			-	-
7.2	ID Fans	172		49			221	26	10	257	2
7.3	Ductwork	342		246			588	71	26	685	5
7.4	Stack						-			-	-
7.9	Duct & Stack Foundations		10	10			20	2	1	23	0
	SUBTOTAL. 7	514	10	305	-	-	829	99	37	965	7
8	STEAM TURBINE GENERATOR						-			-	-
8.1	Steam TG & Accessories	16,296		2,338			18,634	2,236	839	21,709	161
8.2	Turbine Plant Auxiliaries	112		226			338	41	15	394	3
8.3	Condenser & Auxiliaries	2,889		696			3,585	430	161	4,176	31
8.4	Steam Piping	5,733		3,288			9,021	1,082	406	10,509	78
8.9	TG Foundations		325	524			849	102	38	989	7
	SUBTOTAL. 8	25,030	325	7,072	-	-	32,427	3,891	1,459	37,777	281
9	COOLING WATER SYSTEM						-			-	-
9.1	Cooling Towers	2,650		1,340			3,990	479	180	4,649	35
9.2	Circulating Water Pumps	667		56			723	87	33	843	6
9.3	Circulating Water System Auxiliaries	178		22			200	24	9	233	2
9.4	Circulating Water Piping		1,317	1,289			2,606	313	117	3,036	23
9.5	Make-up Water System	151		197			348	42	16	406	3
9.6	Component Cooling Water System	131		102			233	28	10	271	2
9.9	Circ. Water System Foundations Structures		835	1,252			2,087	250	94	2,431	18
	SUBTOTAL. 9	3,777	2,152	4,258	-	-	10,187	1,223	459	11,869	88
10	ASH/SPENT SORBENT HANDLING SYSTEMS						-			-	-
10.1	Ash Coolers						-			-	-
10.2	Cyclone Ash Letdown						-			-	-
10.3	HGCU Ash Letdown						-			-	-
10.4	High Temperature Ash Piping						-			-	-
10.5	Other Ash Recovery Equipment						-			-	-
10.6	Ash Storage Silos	290		840			1,130	136	51	1,317	10
10.7	Ash Transport & Feed Equipment	3,645		1,808			5,453	654	281	6,388	47
10.8	Misc. Ash Handling Equipment						-			-	-
10.9	Ash/Spent Sorbent Foundations		137	153			290	35	15	340	3
	SUBTOTAL. 10	3,935	137	2,801	-	-	6,873	825	347	8,045	60

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

Client: ALSTOM Power Inc. Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers										Report Date: 6/20/2003	
TOTAL PLANT COST SUMMARY											
Case 2 - 1x200 gr. MW O2-Fired CFB w/ASU & CO2 Capture											
Net Output Power, kW				134,514		Estimate Type: Conceptual		Cost Base: Jul-03		(\$x1000)	
Acct. No.	Account Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost	
				Direct	Indirect					\$	\$/kW
11	ACCESSORY ELECTRIC PLANT										
11.1	Generator Equipment	676		91			767	92	35	894	7
115.2	Station Service Equipment	2,383		654			3,037	364	137	3,538	26
11.3	Switchgear & Motor Control	2,662		378			3,040	365	137	3,542	26
11.4	Conduit & Cable Tray		1,130	3,392			4,522	543	204	5,269	39
11.5	Wire & Cable		1,312	3,512			4,824	579	217	5,620	42
11.6	Protective Equipment	148		421			569	68	26	663	5
11.7	Standby Equipment	603		11			614	74	28	716	5
11.8	Main Power Transformer	972		38			1,010	121	45	1,176	9
11.9	Electrical Foundations		102	239			341	41	15	397	3
	SUBTOTAL. 11	7,444	2,544	8,736	-	-	18,724	2,247	844	21,815	162
12	INSTRUMENTATION & CONTROL										
12.1	PC Control Equipment						-			-	-
12.2	Combustion Turbine Control						-			-	-
12.3	Steam Turbine Control						-			-	-
12.4	Other Major Component Control						-			-	-
12.5	Signal Processing Equipment						-			-	-
12.6	Control Boards, Panels & Racks	227		113			340	41	15	396	3
12.7	Distributed Control System Equipment	2,518		367			2,885	346	130	3,361	25
12.8	Instrument Wiring & tubing	1,083		2,871			3,954	474	178	4,606	34
12.9	Other I & C Equipment	647		1,226			1,873	225	84	2,182	16
	SUBTOTAL. 12	4,475	-	4,577	-	-	9,052	1,086	407	10,545	78
13	IMPROVEMENT TO SITE										
13.1	Site Preparation		23	394			417	50	19	486	4
13.2	Site Improvement		777	811			1,588	191	71	1,850	14
13.3	Site Facilities	1,392		1,155			2,547	306	115	2,968	22
	SUBTOTAL. 13	1,392	800	2,360	-	-	4,552	547	205	5,304	39
14	BUILDINGS & STRUCTURES										
14.1	FB Boiler Building Foundation		2,193	1,944			4,137	496	186	4,819	36
14.2	Turbine Building		3,265	3,068			6,333	760	285	7,378	55
14.3	Administration Building		416	443			859	103	39	1,001	7
14.4	Circulation Water Pumphouse		89	72			161	19	7	187	1
14.5	Water Treatment Building		289	240			529	64	24	617	5
14.6	Machine Shop		371	251			622	75	28	725	5
14.7	Warehouse		251	254			505	61	23	589	4
14.8	Other Buildings & Structures		154	132			286	34	13	333	2
14.9	Waste Treating Building & Structure		295	902			1,196	144	54	1,394	10
	SUBTOTAL. 14	-	7,323	7,306	-	-	14,628	1,756	659	17,043	127

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

Table 10-4: Original Case-2 Boiler and Balance of Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.	INITIAL & ANNUAL O&M EXPENSES	Cost Base: Jul-03
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers	Case 2 - 1x200 gr. MW O2-Fired CFB w/ASU & CO2 Capture	
	Net Plant Heat Rate (Btu/kWh): 13,548	
	Net Power Output (kW): 134,514	
	Capacity Factor (%): 80	
BOILER ISLAND AND BALANCE OF PLANT O&M COSTS		
OPERATING & MAINTENANCE LABOR		
<u>Operating Labor</u>		
Operating Labor Rate (Base):	30.90 \$/hour	
Operating Labor Burden:	30.00 %	
Labor O-H change Rate:	25.00 %	
Operating Labor Requirements (O.J.) per shift	<u>1 unit/mod.</u>	<u>Total Plant</u>
Skilled Operator	1.0	1.0
Operator	7.0	7.0
Foreman	1.0	1.0
Lab Tech's, etc.	1.0	1.0
TOTAL O.J.'s	10.0	10.0
Annual Operating Labor Costs (calc'd)		Annual Cost \$ / year
Maintenance Labor Costs (calc'd)		3,518,892
Administrative & Support Labor (calc'd)		829,868
TOTAL FIXED OPERATING COSTS		1,087,190
		5,435,950
		40.41
Maintenance Material Cost (calc'd)		995,842
		0.00106
<u>Consumables</u>	Consumption	Unit
	<u>Initial</u>	<u>Per Day</u>
Water (11000 gallons)		3,809
		1.00
		1,112,228
		0.00118
Chemicals		
MU & WT Chem. (lbs.)	553,200	18,440
		0.16
Limestone (ton)	10,156	339
		10.00
Formic Acid (lbs.)		0.60
Ammonia, NH3 (ton)		220
Subtotal Chemicals		187,702
		1,849,937
		0.0020
Other		
Supplemental Fuel (MBtu)		
SCR Catalyst Replacement (MBtu)		
Emissions Penalties		
Subtotal Other		
Waste Disposal		
Fly Ash & Bottom Ash (ton)	783	8.00
Subtotal Solid Waste Disposal		1,828,387
		0.00194
		1,828,387
		0.0019
By-Products & Emissions		
Gypsum (ton)		
Subtotal By-Products		
TOTAL VARIABLE OPERATING COST		5,786,394
		0.0061

Table 10-6: Original Case-2 Gas Processing System Operating and Maintenance Costs

Operating Costs (\$/yr)	Variable Costs	Fixed Costs
Chemical and Dessicant	52500	
Waste Handling	0	
Fuel Gas *	137921	
Electricity**	0	
Operating Labor	0	306600
Maintenance (Material & Labor)	1615980	
Contracted services	830000	
Column Total	2636401	306600
Grand Total (Fixed & Variable)	2943001	
* Based on \$4/ MMBU and 7000 hours/ yr.		
** Included in overall facility operating cost		

Table 10-7: Original Case-2 Air Separation Unit Operating and Maintenance Costs

Operating Cost (\$/yr)	Variable Costs	Fixed Costs
Minor Consumables	20,000	
Cooling Water*	0	
Natural Gas***	357,253	
Prepurified Adsorbent**	0	
Operating Labor		2,111,335
Column Total	377,253	2,111,335
Grand Total (Fixed + Variable)	2,488,588	
<p>* Cooling water is supplied by others; thus, major treatment chemicals are part of this supply</p> <p>** Prepurified adsorbent is included in the plant and is typically not replaced</p> <p>***Based on $\\$4.0/10^6$ Btu and 7008 hours/year</p>		

10.3 Updated Case-2: Detailed Investment Costs and Operating and Maintenance Costs

Table 10-8: Updated Case-2 Overall Power Plant Investment Costs

Client: ALSTOM Power Inc. Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers										Report Date: 10/7/2004	
TOTAL PLANT COST SUMMARY											
Case 2-Updated - 1x200 gross MW O2-Fired CFB w/ASU & CO2 Capture											
Net Output Power, kW				138,402		Estimate Type: Conceptual			Cost Base: Jul-03		(\$x1000)
Acct. No.	Account Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost	
				Direct	Indirect					\$	\$/kW
1	FUEL & SORBENT HANDLING	5,324	1,533	3,361	-	-	10,218	1,225	460	11,903	86
2	FUEL & SORBENT PREP. & FEED	2,802	198	834	-	-	3,834	460	172	4,467	32
3	FEEDWATER & MISC. BOP SYSTEMS	11,122	-	5,525	-	-	16,647	1,998	750	19,394	140
4	FLUIDIZED BED BOILER										
4.1	Fluidized Bed Boiler w/o Bhse. & Accessories									49,388	357
4.2	Air Separation Unit (ASU)									64,000	462
4.4	Open										
4.4 - 4.9	Boiler BOP (Fluidizing Air Fans)	170	-	49	-	-	219	26	10	255	2
	SUBTOTAL 4	170	-	49	-	-	219	26	10	113,643	821
5	FLUE GAS CLEANUP										
5.1-5.6	FDA, Bag House & Accessories	1,328	-	1,260	-	-	2,588	311	116	10,416	75
5.9	Gas Processing System (GPS)									57,122	413
	SUBTOTAL 5	1,328	-	1,260	-	-	2,588	311	116	67,538	488
6	COMBUSTION TURBINE ACCESSORIES										
6.1	Combustion Turbine Generator										
6.2-6.9	Combustion Turbine Accessories										
	SUBTOTAL 6	-	-	-	0	0	-	-	-	-	0
7	HRSG DUCTING & STACK										
7.1	Heat Recovery Steam Generator										
7.2-7.9	ID Fans, Ductwork and Stack	514	10	305	-	0	829	99	37	965	7
	SUBTOTAL 7	514	10	305	-	-	829	99	37	965	7
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	16,415	-	2,355	0	0	18,770	2,252	845	21,867	158
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	8,870	325	4,767	0	0	13,961	1,675	628	16,264	118
	SUBTOTAL 8	25,285	325	7,122	-	-	32,731	3,928	1,473	38,131	276
9	COOLING WATER SYSTEM	3,749	2,136	4,227	-	-	10,112	1,214	456	11,782	85
10	ASH/SPENT SORBENT HANDLING SYSTEMS	3,528	123	2,511	-	-	6,162	740	311	7,213	52
11	ACCESSORY ELECTRIC PLANT	7,480	2,505	8,625	-	-	18,610	2,232	839	21,681	157
12	INSTRUMENTATION & CONTROL	4,475	-	4,577	-	-	9,052	1,086	407	10,545	76
13	IMPROVEMENT TO SITE	1,392	800	2,360	-	-	4,552	547	205	5,304	38
14	BUILDINGS & STRUCTURES	-	7,323	7,306	-	-	14,628	1,756	659	17,043	123
	TOTAL COST	67,168	14,953	48,062	-	-	130,182	15,622	5,894	329,610	2,382

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

Client: ALSTOM Power Inc.										Report Date: 10/7/2004	
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers											
TOTAL PLANT COST SUMMARY											
Case 2 - 1x200 gross MW O2-Fired CFB w/ASU & CO2 Capture											
Net Output Power, kW				138,402		Estimate Type: Conceptual		Cost Base: Jul-03		(\$x1000)	
Acct. No.	Account Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost	
				Direct	Indirect					\$	\$/kW
1	FUEL & SORBENT HANDLING										
1.1	Coal Receive & Unload	1,160		555			1,715	206	77	1,998	14
1.2	Fuel Stackout and Reclaim	1,499		392			1,891	227	85	2,203	16
1.3	Fuel Conveyors	1,393		352			1,745	209	79	2,033	15
1.4	Other Fuel Handling	365		82			447	54	20	521	4
1.5	Sorbent Receive & Unload	34		11			45	5	2	52	0
1.6	Sorbent Stackout and Reclaim	555		107			662	79	30	771	6
1.7	Sorbent Conveyors	198	40	51			289	35	13	337	2
1.8	Other Sorbent Handling	120	26	66			212	25	9	246	2
1.9	Fuel & Sorbent Hnd. Foundations		1,466	1,746			3,212	385	145	3,742	27
	SUBTOTAL. 1	5,324	1,533	3,361	-	-	10,218	1,225	460	11,903	86
2	FUEL & SORBENT PREP. & FEED										
2.1	Coal Crushing & Drying	640		131			771	93	35	899	6
2.2	Fuel Conveyor to Storage	2,049		469			2,518	302	113	2,933	21
2.3	Fuel Injection System						-			-	-
2.4	Misc. Fuel Prep. & Feed						-			-	-
2.5	Sorbent Prep. Equipment	-		-			-	-	-	-	-
2.6	Sorbent Storage & Feed	113		45			158	19	7	185	1
2.7	Sorbent Injection System						-			-	-
2.8	Booster Air Supply System						-			-	-
2.9	Fuel & Sorbent Feed. Foundations		198	189			387	46	17	450	3
	SUBTOTAL. 2	2,802	198	834	-	-	3,834	460	172	4,467	32
3	FEEDWATER & MISC. BOP SYSTEMS										
3.1	Feedwater System	2,394		907			3,301	396	149	3,845	28
3.2	Water Makeup & Pretreating	1,798		553			2,351	282	106	2,739	20
3.3	Other Feedwater Subsystems	2,053		903			2,956	355	133	3,444	25
3.4	Service Water System	345		183			528	63	24	615	4
3.5	Other Boiler Plant Systems	1,934		1,747			3,681	442	166	4,289	31
3.6	FO Supply System & Nat. Gas	89		112			201	24	9	234	2
3.7	Waste Treatment Equipment	1,316		664			1,980	238	89	2,307	17
3.8	Misc. Equip. (Cranes, AirComp.Comm.)	1,193		456			1,649	198	74	1,921	14
	SUBTOTAL. 3	11,122	-	5,525	-	-	16,647	1,998	750	19,394	140
4	FLUIDIZED BED BOILER										
4.1	Fluidized Bed Boiler, w/o Bhse & Accessories						-			49,388	357
4.2	Air Separation Unit (ASU)						-			64,000	462
4.3	Open						-			-	-
4.4	Boiler BoP (Fluidizing Air Fans)	34		10			44	5	2	51	0
4.5	Primary Air System (Fans)	136		39			175	21	8	204	1
4.6	Secondary Air System (Fans)						-			-	-
4.7	Major Component Rigging						-			-	-
4.8	Boiler Foundation						-			-	-
	SUBTOTAL. 4	170	-	49	-	-	219	26	10	113,643	821

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

Client: ALSTOM Power Inc. Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers										Report Date: 10/7/2004	
TOTAL PLANT COST SUMMARY											
Case 2 - 1x200 gross MW O2-Fired CFB w/ASU & CO2 Capture											
Net Output Power, kW				138,402		Estimate Type: Conceptual		Cost Base: Jul-03		(\$x1000)	
Acct. No.	Account Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost	
				Direct	Indirect					\$	\$/kW
5	FLUE GAS CLEANUP										
5.1	Absorber Vessels & Accessories						-			-	-
5.2	Other FGD						-			-	-
5.3	FDA, Bag House & Accessories	1,150		1,081			2,231	268	100	10,000	72
5.4	Other Particulate Removal Materials	178		179			357	43	16	416	3
5.5	Gypsum Dewatering System						-			-	-
5.6	Mercury Removal System						-			-	-
5.1-5.6		1,328	-	1,260	-	-	2,588	311	116	10,416	75
5.9	Gas Processing System (GPS)						-			57,122	413
	SUBTOTAL 5	1,328	-	1,260	-	-	2,588	311	116	67,538	488
6	COMBUSTION TURBINE ACCESSORIES						-			-	-
6.1	Combustion Turbine Generator						-			-	-
6.2	Combustion Turbine Accessories						-			-	-
6.3	Compressed Air Piping						-			-	-
6.9	Combustion Turbine Foundations						-			-	-
	SUBTOTAL 6	-	-	-	-	-	-			-	-
7	HRSG DUCTING & STACK						-			-	-
7.1	Heat Recovery Steam Generator						-			-	-
7.2	ID Fans	172		49			221	26	10	257	2
7.3	Ductwork	342		246			588	71	26	685	5
7.4	Stack						-			-	-
7.9	Duct & Stack Foundations		10	10			20	2	1	23	0
	SUBTOTAL 7	514	10	305	-	-	829	99	37	965	7
8	STEAM TURBINE GENERATOR						-			-	-
8.1	Steam TG & Accessories	16,415		2,355			18,770	2,252	845	21,867	158
8.2	Turbine Plant Auxiliaries	112		226			338	41	15	394	3
8.3	Condenser & Auxiliaries	3,025		729			3,753	450	169	4,372	32
8.4	Steam Piping	5,733		3,288			9,021	1,082	406	10,509	76
8.9	TG Foundations		325	524			849	102	38	989	7
	SUBTOTAL 8	25,285	325	7,122	-	-	32,731	3,928	1,473	38,131	276
9	COOLING WATER SYSTEM						-			-	-
9.1	Cooling Towers	2,631		1,330			3,961	475	179	4,615	33
9.2	Circulating Water Pumps	662		56			718	86	33	837	6
9.3	Circulating Water System Auxiliaries	177		22			199	24	9	231	2
9.4	Circulating Water Piping		1,307	1,280			2,587	311	116	3,014	22
9.5	Make-up Water System	150		196			345	42	16	403	3
9.6	Component Cooling Water System	130		101			231	28	10	269	2
9.9	Circ. Water System Foundations Structures		829	1,243			2,072	248	93	2,413	17
	SUBTOTAL 9	3,749	2,136	4,227	-	-	10,112	1,214	456	11,782	85
10	ASH/SPENT SORBENT HANDLING SYSTEMS						-			-	-
10.1	Ash Coolers						-			-	-
10.2	Cyclone Ash Letdown						-			-	-
10.3	HGCU Ash Letdown						-			-	-
10.4	High Temperature Ash Piping						-			-	-
10.5	Other Ash Recovery Equipment						-			-	-
10.6	Ash Storage Silos	260		753			1,013	122	46	1,181	9
10.7	Ash Transport & Feed Equipment	3,268		1,621			4,889	586	252	5,727	41
10.8	Misc. Ash Handling Equipment						-			-	-
10.9	Ash/Spent Sorbent Foundations		123	137			260	31	13	305	2
	SUBTOTAL 10	3,528	123	2,511	-	-	6,162	740	311	7,213	52

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

Client: ALSTOM Power Inc.										Report Date: 10/7/2004	
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers											
TOTAL PLANT COST SUMMARY											
Case 2 - 1x200 gross MW O2-Fired CFB w/ASU & CO2 Capture											
Net Output Power, kW				138,402		Estimate Type: Conceptual		Cost Base: Jul-03		(\$x1000)	
Acct. No.	Account Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost	Professional Services	Other Costs	Total Cost	
				Direct	Indirect					\$	\$/kW
11	ACCESSORY ELECTRIC PLANT										
11.1	Generator Equipment	676		91			767	92	35	894	6
11.2	Station Service Equipment	2,345		644			2,989	359	135	3,482	25
11.3	Switchgear & Motor Control	2,716		386			3,101	372	140	3,613	26
11.4	Conduit & Cable Tray		1,112	3,338			4,450	534	201	5,185	37
11.5	Wire & Cable		1,291	3,456			4,748	569	214	5,531	40
11.6	Protective Equipment	148		421			569	68	26	663	5
11.7	Standby Equipment	603		11			614	74	28	715	5
11.8	Main Power Transformer	992		39			1,030	124	46	1,200	9
11.9	Electrical Foundations		102	239			341	41	15	397	3
	SUBTOTAL. 11	7,480	2,505	8,625	-	-	18,610	2,232	839	21,681	157
12	INSTRUMENTATION & CONTROL										
12.1	PC Control Equipment						-			-	-
12.2	Combustion Turbine Control						-			-	-
12.3	Steam Turbine Control						-			-	-
12.4	Other Major Component Control						-			-	-
12.5	Signal Processing Equipment						-			-	-
12.6	Control Boards, Panels & Racks	227		113			340	41	15	396	3
12.7	Distributed Control System Equipment	2,518		367			2,885	346	130	3,361	24
12.8	Instrument Wiring & tubing	1,083		2,871			3,954	474	178	4,606	33
12.9	Other I & C Equipment	647		1,226			1,873	225	84	2,182	16
	SUBTOTAL. 12	4,475	-	4,577	-	-	9,052	1,086	407	10,545	76
13	IMPROVEMENT TO SITE										
13.1	Site Preparation		23	394			417	50	19	486	4
13.2	Site Improvement		777	811			1,588	191	71	1,850	13
13.3	Site Facilities	1,392		1,155			2,547	306	115	2,968	21
	SUBTOTAL. 13	1,392	800	2,360	-	-	4,552	547	205	5,304	38
14	BUILDINGS & STRUCTURES										
14.1	FB Boiler Building Foundation		2,193	1,944			4,137	496	186	4,819	35
14.2	Turbine Building		3,265	3,068			6,333	760	285	7,378	53
14.3	Administration Building		416	443			859	103	39	1,001	7
14.4	Circulation Water Pumphouse		89	72			161	19	7	187	1
14.5	Water Treatment Building		289	240			529	64	24	617	4
14.6	Machine Shop		371	251			622	75	28	725	5
14.7	Warehouse		251	254			505	61	23	589	4
14.8	Other Buildings & Structures		154	132			286	34	13	333	2
14.9	Waste Treating Building & Structure		295	902			1,196	144	54	1,394	10
	SUBTOTAL. 14	-	7,323	7,306	-	-	14,628	1,756	659	17,043	123

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

Table 10-9: Updated Case-2 Boiler and Balance of Plant Operating and Maintenance Costs

Client: ALSTOM Power Inc.		INITIAL & ANNUAL O&M EXPENSES		Cost Base: Jul-03	
Project: Greenhouse Gas Emissions Control by Oxygen Firing in Circulating Fluidized Bed Boilers		Case 2 - 1x200 gross MW O2-Fired CFB w/ASU & CO2 Capture			
				Net Plant Heat Rate (Btu/kWh): 13,152	
				Net Power Output (kW): 138,402	
				Capacity Factor (%): 80	
BOILER ISLAND AND BALANCE OF PLANT O&M COSTS					
OPERATING & MAINTENANCE LABOR					
<u>Operating Labor</u>					
Operating Labor Rate (Base):		30.90 \$/hour			
Operating Labor Burden:		30.00 %			
Labor O-H change Rate:		25.00 %			
Operating Labor Requirements (O.J.) per shift		<u>1 unit/mod.</u>	<u>Total Plant</u>		
Skilled Operator		1.0	1.0		
Operator		7.0	7.0		
Foreman		1.0	1.0		
Lab Tech's, etc.		1.0	1.0		
TOTAL O.J.'s		10.0	10.0		
				Annual Cost	Annual Unit Cost
				<u>\$ / year</u>	<u>\$/kW-net</u>
Annual Operating Labor Costs (calc'd)				3,518,892	25.43
Maintenance Labor Costs (calc'd)				833,951	6.03
Administrative & Support Labor (calc'd)				1,088,211	7.86
TOTAL FIXED OPERATING COSTS				5,441,054	39.31
Maintenance Material Cost (calc'd)				1,000,741	0.00103
<u>Consumables</u>					
		<u>Consumption</u>	<u>Unit</u>	<u>Initial</u>	
		<u>Initial</u>	<u>Per Day</u>	<u>Cost</u>	<u>Cost</u>
Water (11000 gallons)			3,809	1.00	
					1,112,228
Chemicals					
MU & WT Chem. (lbs.)		553,200	18,440	0.16	86,146
Limestone (ton)		10,156	86.472	50.00	101,556
Formic Acid (lbs.)				0.60	
Ammonia, NH3 (ton)				220	
Subtotal Chemicals				187,702	2,124,008
					0.0022
Other					
Supplemental Fuel (MBtu)					
SCR Catalyst Replacement (MBtu)					
Emissions Penalties					
Subtotal Other					
Waste Disposal					
Fly Ash & Bottom Ash (ton)		670	8.00	1,564,362	0.00161
Subtotal Solid Waste Disposal				1,564,362	0.0016
By-Products & Emissions					
Gypsum (ton)					
Subtotal By-Products					
TOTAL VARIABLE OPERATING COST				5,801,340	0.0060

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Table 10-10: Updated Case-2 Gas Processing System Investment Costs

ABB LUMMUS GLOBAL HOUSTON								
Project : CO2 Plant - DOE			Location : GC - USA		Project start:		Rev. : 2.01	
Job/Prop # : 0-9484			Plant : CO2 Case 2		Mech.compl.:			
Scope : EPC			Capacity :					
Piece count:			Labor Prod.:				11-Feb-03	
Acc't Code	Description	Pieces	Direct Manhours	Labor (\$,000)	Material (\$,000)	Subcontract (\$,000)	Total (\$,000)	%
11000	Heaters						-	0.0%
11200	Exchangers & Aircoolers		8,952	139	5,670		5,809	10.2%
12000	Vessels / Filters		2,138	33	1,354		1,387	2.4%
12100	Towers / Internals		1,502	23	951		974	1.7%
12200	Reactors		-	-			-	0.0%
13000	Tanks		-	-			-	0.0%
14100	Pumps		323	5	205		210	0.4%
14200	Compressors		17,842	277	11,300		11,577	20.3%
18000	Special Equipment		-	-	-		-	0.0%
	Sub-Total Equipment	44	30,757	477	19,479	-	19,956	34.9%
21000	Civil		46,135	715	1,753		2,468	4.3%
21100	Site Preparation		-	-	-		-	0.0%
22000	Structures		10,765	167	877		1,043	1.8%
23000	Buildings		12,303	191	468		658	1.2%
30000	Piping		84,582	1,311	3,896		5,207	9.1%
40000	Electrical		43,572	675	1,558		2,234	3.9%
50000	Instruments		35,883	556	2,727		3,283	5.7%
61100	Insulation		23,068	358	584		942	1.6%
61200	Fireproofing		15,378	238	292		531	0.9%
61300	Painting		12,815	199	166		364	0.6%
	Sub-Total Commodities		284,502	4,410	12,321	-	16,731	29.3%
70000	Construction Indirects						7,093	12.4%
	Sub-Total Direct Cost		315,259	4,887	31,800	-	43,780	76.6%
	ASU TIC plant cost						-	0.0%
71000	Constr. Management						700	1.2%
80000	Home Office Engineering						4,488	7.9%
80000	Basic Engineering						600	1.1%
95000	License fee	Excluded						0.0%
19400	Vendor Reps						800	1.4%
19300	Spare parts						1,300	2.3%
80000	Training cost	Excluded						0.0%
80000	Commissioning	Excluded						0.0%
19200	Catalyst & Chemicals						100	0.2%
97000	Freight						954	1.7%
96000	CGL / BAR Insurance							0.0%
	Sub-Total						52,722	92.3%
91400	Escalation						1,500	2.6%
93000	Contingency	Excluded						0.0%
93000	Risk	Excluded						0.0%
	Total Base Cost						54,222	94.9%
	Contractors Fee						2,900	5.1%
	Grand Total						57,122	100.0%

Exclusions : Bonds,Taxes,Import duties , Hazardous material handling & disposal, Capital spare parts, Catalyst & Chemicals , Commissioning and Initial operations, Buildings other than Control room & MCC.

Table 10-11: Updated Case-2 Gas Processing System Operating and Maintenance Costs

Operating Costs (\$/yr)	Variable Costs	Fixed Costs
Chemical and Dessicant	52500	
Waste Handling	0	
Fuel Gas *	137921	
Electricity**	0	
Operating Labor	0	306600
Maintenance (Material & Labor)	1615980	
Contracted services	830000	
Column Total	2636401	306600
Grand Total (Fixed & Variable)	2943001	
* Based on \$4/ MMBU and 7000 hours/ yr.		
** Included in overall facility operating cost		

Table 10-12: Updated Case-2 Air Separation Unit Operating and Maintenance Costs

Operating Cost (\$/yr)	Variable Costs	Fixed Costs
Minor Consumables	20,000	
Cooling Water*	0	
Natural Gas***	357,253	
Prepurified Adsorbent**	0	
Operating Labor		2,111,335
Column Total	377,253	2,111,335
Grand Total (Fixed + Variable)	2,488,588	
<p>* Cooling water is supplied by others; thus, major treatment chemicals are part of this supply</p> <p>** Purpurified adsorbent is included in the plant and is typically not replaced</p> <p>***Based on \$4.0/10⁶ Btu and 7008 hours/year</p>		

11 APPENDIX II: Original Case–2: Oxygen Fired CFB with Moving Bed Heat Exchanger and CO₂ Capture

Section 2.2 of the Phase I Topical Report (Marion, et al., 2003) - which describes the Original Case 2 - is reproduced in its entirety below. The numbering of sections, figures and tables, and the references in this appendix all refer to the original report.

2.2 Case–2: Oxygen Fired CFB with Moving Bed Heat Exchanger and CO₂ Capture

This section describes a power plant comprised of oxygen-fired Circulating Fluidized Bed (CFB) boiler, a Moving Bed Heat Exchanger (MBHE), a cryogenic type Air Separation Unit (ASU), and a subcritical steam plant with reheat (1,800 psia / 1,000 °F / 1,000 °F). The plant is designed to produce a flue gas having a high concentration of CO₂. This stream is then further processed in a Gas Processing System to produce a CO₂ product suitable for usage or sequestration. The plant design configuration reflects current information and design preferences, the availability of a current generation steam turbine, and the design latitude offered by a Greenfield site.

The basic CO₂ capture concept behind Case-2 is to replace combustion air with oxygen thereby creating a high CO₂ content flue gas stream that can be further processed into a high purity CO₂ end product for various uses or sequestration. The replacement of combustion air with high purity oxygen can, as an additional benefit, significantly reduce the gas flow throughout the Boiler Island equipment. In this case a cryogenic Air Separation Unit (ASU) supplies the oxidant (99 percent pure oxygen) for the combustion of coal rather than direct utilization of ambient air as was done in Case-1. Since the size and cost of much of the equipment contained within the Boiler Island is strongly gas flow dependent (combustor, cyclones, backpass heat exchangers, air heater, fans, ductwork, baghouse, etc.) significant cost savings are anticipated for the Boiler Island of this concept as compared to the comparable air-fired CFB (Case-1).

A brief performance summary for this plant reveals the following information. The Case-2 plant produces a net plant output of about 134 MW. The net plant heat rate and thermal efficiency are calculated to be 13,546 Btu/kWh and 25.2 percent respectively (HHV basis) for this case. Carbon dioxide emissions are about 0.18 lbm/kWh on a normalized basis. A more detailed presentation of plant performance is shown in Section 2.2.5.

Agglomeration Concerns:

When firing with oxygen in coal fired equipment it is critical to be able to control temperature in the combustor region in order to avoid potential problems. High combustor temperatures in CFB units can lead to agglomeration of the bed. With oxygen firing in a Circulating Fluid Bed (CFB), agglomeration of bed material is a major operational concern. If the ash particles are raised to a high enough temperature they will become sticky and large masses may form, requiring the unit to be shut down and cleaned out. Therefore, fuels containing ash with low fusibility temperatures are more challenging to burn than high fusibility temperature fuels.

In pulverized coal (PC) units, high furnace temperatures can cause uncontrollable slagging and or fouling which also can require the unit to be shut down to correct. In order to avoid uncontrollable slagging and fouling in oxygen fired pulverized coal and Stoker units, large quantities of recirculated flue gas are used to maintain temperatures

throughout the unit at acceptable levels. The quantity of recirculated gas typically proposed for O₂ fired PC boilers in order to avoid these problems is that amount which makes the flue gas to coal flow ratio equivalent, or nearly equivalent, to that of air firing. Because of the large quantities of recirculated flue gas required, an oxygen fired pulverized coal unit is actually slightly more costly than the comparable air fired pulverized coal unit due to the additional cost of the gas recirculation system (additional ductwork, fans, and controls).

Similarly, it is possible in an oxygen fired CFB system, using the proper amount of recirculated gas, to obtain bed conditions within the combustor that are very similar to air firing. Systems can be designed to have the same flue gas flow to coal flow ratio as with air firing and the same volumetric fraction of oxygen in the flue gas leaving the bed as that of the air fired case. This scenario would represent the most conservative oxygen fired condition and is very similar to what is done in oxygen fired pulverized coal systems as described above. Because the flue gas flow to coal flow ratio would be the same as with air firing in this scenario, the size and cost of the Combustor, Cyclone, and Convection pass heat exchanger components and other equipment would also be nearly identical to the air fired case.

A potentially significant advantage for oxygen fired CFB type combustors as compared to oxygen fired pulverized coal fired or Stoker fired counterparts is the capability to reduce flue gas flow for a given coal input while maintaining combustor temperature. This capability is available to CFB units because there is an additional controllable variable available for CFB systems. Because within the CFB process (both air and oxygen fired) there is a large stream of recirculated cooled solids returning to the combustor from the external heat exchanger, the bed temperatures can theoretically be maintained at required levels in both the air and oxygen fired scenario thus avoiding any potential agglomeration. Combustor temperature can be maintained when firing with oxygen by simply cooling a greater portion of the large solids recirculation loop leaving the cyclone. Equipment designs developed for this case shows this type of system is technically feasible. Cost analysis, presented later in this report (Section 3), clearly quantifies the benefit of the reduced gas flow used for this case.

An extremely wide variety of fuels have been successfully used in ALSTOM-designed air-fired CFB units. Table 2.2.1 shows of a cross section of such fuels. They cover the complete coal rank spectrum from brown coal to anthracite, plus other waste fuels (e.g., delayed petroleum coke, anthracite culm). Table 2.2.1 demonstrates the extreme fuel-flexibility of air-fired CFB's, as they range in as-received ash contents from less than 1 percent to more than 39 percent, and higher heating values from 6,100 Btu/lbm to 12,960 Btu/lbm, and in moisture contents from 3 to 31 percent. These fuels also cover a wide range of sulfur contents, from less than 1 to more than 6 percent, indicating, correspondingly, high flexibility of CFB's in firing fuels ranging from low to high slagging potentials.

Table 2.2. 1: Range of Fuels Fired in Air Fired CFB's

Quantity	Sub-Bituminous Coal	Brown Coal	Lignite	High Volatile Bituminous Coal	High Volatile Bituminous Coal	Medium Volatile Bituminous Coal	Medium Volatile Bituminous Coal	Delayed Petroleum Coke	Anthracite Culm	Anthracite
Proximate & Ultimate Analyses, Wt.%										
Moisture	11.6	24.6	31.0	6.5	2.4	4.7	5.4	14.4	14.6	3.3
Ash	25.5	24.7	14.8	11.0	15.8	14.0	13.0	0.2	33.9	39.1
Volatile Matter	40.4	26.5	24.5	36.0	35.6	18.5	17.5	10.6	6.5	4.1
Fixed Carbon (Diff.)	22.6	24.2	29.7	46.5	46.2	62.8	64.1	74.8	45.0	53.5
Hydrogen	2.9	3.0	3.0	4.7	4.4	3.7	3.6	3.0	0.8	0.3
Carbon	43.3	35.0	40.6	65.5	66.1	70.2	74.2	74.2	45.7	52.9
Nitrogen	1.0	0.5	0.7	1.2	1.2	1.5	1.5	1.0	0.6	0.2
Sulfur	3.5	1.1	0.7	0.6	4.3	1.8	1.7	6.3	0.6	0.6
Oxygen (Diff.)	12.3	11.1	9.2	10.6	5.8	4.1	0.5	0.9	3.8	3.6
HHV, Btu/lb, As Received	5785	6121	6856	11675	12067	12221	12256	12963	7000	8152
Carbon, DAF	68.7	69.1	74.9	79.3	80.8	86.4	90.9	86.9	88.7	91.8
Volatile Matter, DAF	64.1	52.3	45.2	43.6	43.5	22.8	21.5	12.4	12.6	7.1

DAF = Dry-Ash-Free

Bed temperatures in these units are maintained at acceptable values by recirculation of cooled solids from the external heat exchanger to the combustor. Additionally, heat is transferred from the gas-solids mixture to the combustor waterwalls and other heat exchange surfaces located in the combustor. Proper balancing of the combustor heat absorption and the external heat exchanger heat absorption ensure the desired combustor bed temperature is achieved. The large inventory of solids within the bed with its associated large thermal inertia along with the high degree of mixing help to limit local temperature excursions within the bed. These favorable characteristics inherent within the CFB process help to make the compact oxygen fired CFB concept of Case-2 viable with respect to the mitigation of potential agglomeration.

Oxygen Fired Bench Scale FBC Testing:

Tests were performed in Task-2 of this project (Bench Scale FBC Testing) to specifically address the agglomeration concerns associated with oxygen fired CFB systems. A wide range of operating conditions was investigated in the testing as described below. These conditions covered a significant range from traditional air firing (representing a baseline condition) to 70 percent oxygen firing. Higher than 70 percent oxygen levels were not tested due to equipment limitations. Additionally, tests with various intermediate levels of oxygen were also investigated in the testing.

One of the test facilities used by ALSTOM to characterize fuels and limestones for commercial CFB design purposes is a bench-scale (four-inch inner diameter) fluidized bed combustor (FBC). This facility was used to combustion test three fuels along with a limestone sample from the Base Case CFB plant. Details of these tests are presented in Volume II of this report. A brief summary of the fuel characteristics, test conditions, results/observations, and concluding remarks are given below:

Fuel Characteristics:

Three fuels were tested in Task-2. The Base Case CFB coal has nitrogen and sulfur contents of 1.7 percent and 2.4 percent on a dry-ash-free (DAF) basis, respectively; these values are typical of those of bituminous coals. These values, combined with high ash fusibility temperatures (I.T. = 2,400 °F and F.T. = 2,800 °F) place this coal in a low slagging potential range. The second coal tested (Illinois #6) has nitrogen and sulfur contents of 1.5 percent and 4.9 percent. These values, combined with low ash fusibility temperatures (I.T. = 2,070 °F and F.T. = 2,260 °F) place this coal in a high slagging potential range. The third fuel tested, Petroleum Coke, has nitrogen and sulfur contents of 4.0 percent and 1.5 percent (DAF). Furthermore, it has a very low ash content of 0.3 percent. These conditions are indicative of a fuel with low slagging potential.

Test Conditions:

All three fuels were burned in air and 30 percent O₂/70 percent CO₂ (by volume) mediums. Additionally, the Base Case CFB and Illinois #6 coals were burned in O₂/CO₂ mediums containing from 21 to 70 percent O₂, and from 21 to 50 percent O₂, respectively. Bed temperatures ranged from about 1,635 °F to 1,900 °F. The limestone sample from the Base Case CFB plant was used selectively to determine its impact on the flue gas desulfurization potential. All tests were carried out in high excess O₂ (with

stoichiometric ratios ranging from about 2 to about 4.7). Table 2.2.2 summarizes these test conditions.

Table 2.2. 2: Test Matrix for Bench-Scale Fluidized Bed Combustion

Fuel	Series	Ca/S Mole Ratio	Combustion Gas Medium	Bed Temperature, °F	Superficial Gas Velocity, ft/sec.	Stoichiometry, λ	Measurements/Observations:
Base Case CFB Coal	I	---	Air	1635-1908	1.8 - 3.3	2.0 - 4.7	☐ Ash slagging/sintering potentials as a result of increased oxygen concentrations in the combustion gas, and/or use of a slagging coal ☐ Flue gas desulfurization potential by the limestone sample used ☐ NO _x emissions reduction by use of nitrogen-free O ₂ /CO ₂ combustion mediums ☐ Impact of various combustion mediums on: (1) overall fuel combustion efficiency; (2) unburned carbon emissions; and (3) CO emissions ☐ Impact of fuel nature on all the parameters specified above
	II	---	23% O ₂ -70% O ₂ /in CO ₂ Balance				
	III	3.5	Air				
		3.5	30% O ₂ /70% CO ₂				
Illinois #6 Coal	I	---	Air				
	II	---	21% O ₂ -50% O ₂ /in CO ₂ Balance				
	III	3.5	Air				
		3.5	30% O ₂ /70% CO ₂				
Delayed Petroleum Coke	I	---	Air				
	II	---	30% O ₂ /70% CO ₃				
	III	3.50	Air				
		3.50	30% O ₂ /70% CO ₃				

Results/Observations:

Testing the Base Case CFB coal in O₂/CO₂ mediums containing up to 70 percent O₂ caused, as expected, a bed temperature rise of up to about 250 °F. Nevertheless, it was possible to obviate bed slagging/de-fluidization problems as long as the bed was well fluidized. As a frame of reference, the superficial gas velocity had to be maintained at $\geq$ 2.5 ft/sec., whereas for combustion testing in air and O₂/CO₂ mediums containing $\leq$ 40 percent O₂ a superficial gas velocity of $\sim$ 1.7 ft/sec. was sufficient to keep the bed fluidized at all times. The Illinois #6 coal, which has high slagging potential, was successfully burned in the FBC in O₂/CO₂ mediums containing up to 50 percent O₂. Although the test in 70 percent O₂/30 percent CO₂ was not run, due principally to time and financial constraints, there was no reason to believe that it would not have been executed successfully.

Concluding Remarks:

The test conditions used here in the FBC facility are very aggressive for two reasons: (1) there is no means of recycling particles to control the bed temperature; and (2) given the prevailing stoichiometric ratios, the furnace outlet O₂ concentration ranged from 13 to 51 percent (Dry basis).

On the other hand, the operating conditions of a commercial oxy-fuel fired CFB would be much less aggressive for the following reasons: (1) the bed temperature is closely controlled through judicious recycling of cooled bed materials; (2) the superficial gas velocity is maintained at about 18 ft/sec.; and (3) the O₂ concentration in the furnace would rapidly decline from its initial value of 70 percent, for Case 2, to about 3 percent at the outlet.

Based on these findings, it was preliminarily concluded that the choice of 70 percent O₂/30 percent recycled flue gas (i.e., ~CO₂) as a combustion medium for study Case 2 (New Compact O₂-Fired CFB) was plausible. This oxygen content was also utilized in Cases 3, 4 and 6. Pilot-scale testing in Phase II will evaluate the Case-2 concept, among other things. Based on the pilot-scale testing the concept presented in Case 2 will be either affirmed for further commercial development or modified beforehand.

2.2.1. Case 2 Boiler Island Process Description and Equipment

As discussed above, the basic concept for Case-2 is to minimize the amount of gas recirculation such that the size and cost of the above mentioned components, and other components, can be reduced significantly. If oxygen firing can be used in a CFB system with minimal gas recirculation, the flue gas flow to coal flow ratio can be reduced. This ratio was reduced from about 11.9 lbm gas/lbm coal with air firing (Case-1) to about 4.2 lbm gas/lbm coal for Case-2 with oxygen firing. This gas to coal ratio equates to about 70 percent oxygen by volume in the oxidant feed streams to the combustor. Since the size of much of the Boiler Island system components is strongly dependent on gas flow, significant size and cost savings are anticipated for these components.

This section describes the Boiler Island processes for Case-2 and includes a simplified process flow diagram (PFD), material and energy balance and equipment description. The equipment description includes only the major components of the Boiler Island.

2.2.1.1. Process Description and Process Flow Diagrams

Figure 2.2.1 shows a simplified process flow diagram for the Boiler Island of the Case-2 oxygen-fired CFB concept. This process description briefly describes the function of the major equipment and systems included within the Boiler Island. Selected mass flow rates (lbm/hr) and temperatures (°F) are shown on this figure. Complete data for all state points are shown in Table 2.2.3. In this concept coal or another high carbon content fuel (Stream 1) is reacted with a preheated mixture of substantially pure oxygen and recirculated flue gas (Stream 18) in the Combustor section of the Circulating Fluidized Bed (CFB) system. The oxygen (Streams 16, 17, 18) is provided from a cryogenic Air Separation Unit (ASU).

The products of combustion, flue gas comprised of primarily CO₂ and H₂O vapor and un-reacted hot solids, flow through a cyclone, or another type of particulate removal device, where most of the hot solids are removed and recirculated to the combustor. The temperature in the combustor is controlled to the proper level by properly splitting the flow of hot recirculated solids leaving the cyclone, between an uncooled stream which flows directly back to the Combustor and a stream flowing through the External Heat Exchanger where the solids are cooled before returning to the Combustor. Exchanging heat with the power cycle working fluid cools the hot solids in the External Heat Exchanger.

Draining hot solids through water-cooled fluidized bed ash coolers (Stream 21) controls solids inventory in the system while recovering heat from the hot ash. The cooling water used for the ash coolers is feedwater from the final extraction feedwater heater of the steam cycle.

The flue gas leaving the cyclone (Stream 3) is cooled in heat exchanger sections located in the convection pass of the system, also by exchanging heat with the power cycle working fluid. The flue gas leaving the convection pass heat exchanger sections (Stream 5) is further cooled in an Oxygen Heater. The oxygen stream leaving the Air Separation Unit (Stream 16) is mixed with a small stream of recirculated flue gas (Stream 15) and the mixture is preheated in the Oxygen Heater. The quantity of recirculated flue gas is only the amount necessary to provide proper fluidization for the bed.

The flue gas leaving the Oxygen Heater (Stream 6) is cleaned of fine particulate matter in the baghouse and further cooled in a Parallel Feedwater Heater (PFWH) by transferring heat to feedwater in parallel with extraction feedwater heaters. Finally, a Gas Cooler is used to cool the gas before the flue gas enters the Induced Draft (ID) Fan (Stream 10). The gas cooler is used to cool the flue gas to the lowest temperature possible before recycling to minimize the power requirements for the draft system (induced draft fan, fluidizing air blower, and gas recirculation fan) and the product gas compression system. Some H₂O vapor is condensed in the Gas Cooler. The flue gas leaving the ID Fan (Stream 11), comprised of mostly CO₂ and H₂O vapor, is split with most of the flue gas going to the product stream (Stream 12) for further processing and the remainder recirculated to the CFB system.

By using oxygen instead of air for combustion, and by minimizing the amount of recirculated flue gas, the size and cost of the Combustor, Cyclone, and Convection Pass Heat Exchanger sections and other equipment can be reduced as compared to many other concepts for CO₂ capture with CFB systems.

Figure 2.2. 1: Case-2: Simplified Boiler Island Gas Side Process Flow Diagram

2.2.1.2. Material and Energy Balance

Table 2.2.3 shows the Boiler Island material and energy balance for Case-2. The stream numbers shown at the top of each column of the table refer to stream numbers shown in the Case-2 simplified PFD for the Boiler Island (Figure 2.2.1). This performance was calculated at MCR conditions for this unit.

The MCR condition is defined as high-pressure turbine inlet conditions of 1,400,555 lbm/hr, 1,815 psia, and 1,000 °F and intermediate-pressure turbine inlet conditions of 1,305,632 lbm/hr 469 psia 1,000 °F. These conditions were also used for the Base Case. The boiler was fired with enough oxygen to leave about 3 percent by volume in Stream 3, the same as for the Base Case. This oxygen requirement results in a stoichiometry of about 1.05 for Case-2. The resulting boiler efficiency calculated for Case-2 was 94.12 percent (HHV basis) with an oxygen heater exit gas temperature of 313 °F and the PFWH exit gas temperature of 136 °F.

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Table 2.2. 3: Case-2 Boiler Island Gas Side Material and Energy Balance

Constituent	(Units)	1	2	3	4	5	6	7	8	9	10	11
O2	(Lbm/hr)	5152		17814	3246	21060	21060	21060	21060		21060	21060
N2	"	2380		14156	10753	24909	24909	24909	24909		24909	24909
H2O	"	6505		65228	182	65409	65409	65409	65409	45826	19583	19583
CO2	"			570438		570438	570438	570438	570438		570438	570438
SO2	"			1159		1159	1159	1159	1159		1159	1159
H2	"	5821										
Carbon	"	101168										
Sulfur	"	3815										
CaO	"											
CaSO4	"											
CaCO3	"		26800									
Ash	"	38201	1411									
Total Gas	(Lbm/hr)	Coal	Limestone	Flue Gas	Infiltration Air	Flue Gas	Flue Gas	Flue Gas	Flue Gas	Condensate	Flue Gas	Flue Gas
Total Solids	"	163043	28210	668795	14180	682975	682975	682975	682975		637149	637149
Total Flow	"	163043	28210	668795	14180	682975	682975	682975	682975	45826	637149	637149
Temperature	(Deg F)	80	80	1600	80	587	313	313	136	100	100	112
Pressure	(Psia)	14.7	14.7	14.7	14.7	14.4	14.2	13.8	13.7	14.7	13.6	14.7
hsensible	(Btu/lbm)			439.702	0.000	128.882	56.045	56.045	12.794		4.228	6.826
									19.960			
Chemical	106 Btu/hr	1805.540										
Sensible	106 Btu/hr	0.0	0.000	294.071	0.000	88.023	38.277	38.277	8.738	0.915	2.694	4.349
Latent	106 Btu/hr	0.0	0.000	68.489	0.191	68.680	68.680	68.680	68.680	0.000	20.562	20.562
Total Energy(1)	106 Btu/hr	1805.540	0.000	362.560	0.191	156.703	106.957	106.957	77.418	0.915	23.256	24.911

Constituent	(Units)	12	13	14	15	16	17	18	19	20	21	22
O2	(Lbm/hr)	13848	7212	5977	5977	321334	327311	327311	1235	1235		
N2	"	16379	8530	7069	7069	3246	10315	10315	1460	1460		
H2O	"	12877	6706	5558	5558		5558	5558	1148	1148		
CO2	"	375095	195342	161899	161899		161899	161899	33443	33443		
SO2	"	762	397	329	329		329	329	68	68		
H2	"											
Carbon	"										2023	2023
Sulfur	"											0
CaO	"										9009	9009
CaSO4	"										14581	14581
CaCO3	"											0
Ash	"										39612	39612
Total Gas	(Lbm/hr)	Flue Gas	Recirc Gas	Recirc Gas	Recirc Gas	Oxygen	Oxy + Recirc	Oxy + Recirc	Grease Gas	Grease Gas	Hot Ash Drain	Cool Ash Drain
Total Solids	"	418962	218187	180833	180833	324580	505413	505413	37354	37354	65225	65225
Total Flow	"	418962	218187	180833	180833	324580	505413	505413	37354	37354	65225	65225
Temperature	(Deg F)	112	112	112	140	100	114	537	112	195	1600	520
Pressure	(Psia)	14.7	14.7	14.7	17.4	17.4	17.4	17.19	14.7	23.7	14.7	14.7
hsensible	(Btu/lbm)	6.826	6.826	6.826	12.934	4.407	7.458	105.883	6.826	25.005	407.729	95.391
Chemical	106 Btu/hr										28.515	28.515
Sensible	106 Btu/hr	2.860	1.489	1.234	2.339	1.431	3.770	53.515	0.255	0.934	26.594	6.222
Latent	106 Btu/hr	13.521	7.041	5.836	5.836	0.000	5.836	5.836	1.206	1.206	0.000	0.000
Total Energy(1)	106 Btu/hr	16.381	8.531	7.070	8.175	1.431	9.606	59.351	1.460	2.140	55.109	34.737

Notes:—
(1) Energy Basis; Chemical based on Higher Heating Value (HHV); Sensible energy above 80F; Latent based on 1050 Btu/Lbm of water vapor

Figure 2.2.2 shows a general comparison of overall boiler performance between the air firing of Case-1 and the oxygen firing of Case-2. The boiler performance is compared in terms of the heat absorption distribution within the boiler components. Excluding the parallel feedwater heater (PFWH) heat absorption of Case-2, the total heat absorption is the same in both Case-1 and Case-2. The primary differences occur in the Convection Pass, Combustor and External Heat Exchanger. The Convection Pass heat absorption for Case-2 is about 40 percent of the Case-1 value due to the reduced gas flow. Similarly, the Combustor heat absorption for Case-2 is about 32 percent of the Case-1 value due to the reduction in size. Finally, the External Heat Exchanger heat absorption for Case-2 is

about 3.5 times greater than the Case-1 value. Because of the very large heat duty of the external heat exchanger of Case-2 (about 70 percent of the total duty as compared to about 20 percent for Case-1) a moving bed design was selected based on cost considerations. The Case-1 external heat exchanger was a bubbling fluidized bed design.

Figure 2.2. 2: CFB Boiler Heat Absorption Comparison (Air and O₂ Firing)

A moving bed external heat exchanger design that was implemented in this case provides several advantages over a bubbling fluidized bed. A significant advantage of the moving bed is that a higher temperature difference is obtained between the bed material and the steam cycle working fluid thus reducing the quantity of heat exchanger pressure part surface that is required. This occurs because the moving bed can be designed as a counterflow heat exchanger. The bubbling fluid bed on the other hand is more of a “stirred” heat exchanger where the bed material is at a “stirred temperature”. The “stirred temperature” is much lower than the inlet solids temperature in the moving bed. Additionally, the moving bed does not require any fluidizing air, fluidizing nozzles, and fluidizing air piping thus providing a simpler system. With these advantages, the moving bed allows for a much more compact and less expensive design.

2.2.1.3. Boiler Island Equipment

This section describes major equipment included in the Boiler Island for Case-2. The major components included in the Boiler Island include the combustor, ash coolers, fuel feed system, fuel silos, sorbent feed system, sorbent silo, cyclones, seal pots, external moving bed heat exchanger (MBHE), convection pass, superheater, reheater, economizer, oxygen heater, baghouse, parallel feedwater heater (PFWH), gas cooler, and draft system.

Figures 2.2.3 and 2.2.4 show a general arrangement drawing of the Case-2 CFB boiler. The plan area for the Case-2 Boiler Island is about 52 percent of that for Case-1. Similarly, the building volume for Case-2 is about 56 percent of that for Case-1. The complete Boiler Island Equipment List for Case-2 is shown in Appendix I. Appendix II shows several additional drawings of the Boiler (key plan view, boiler plan view, side elevation, and various sectional views).

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Figure 2.2. 3: Case-2 Boiler Island General Arrangement Drawing – Side Elevation

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Figure 2.2. 4: Case-2 Boiler Island General Arrangement Drawing - Plan View

Combustor:

The combustor size is reduced significantly for Case-2. The combustor for Case-2 is about 29 ft wide, 14 ft deep and 100 ft high. Thus, the combustor plan area for Case-2 is about 30 percent of the Case-1 plan area for nearly equivalent fuel heat input quantities. Because of the relatively small combustor size, evaporator panels are not used in the Case-2 combustor as was done for Case-1. Crushed fuel, sorbent, and recycle solids are fed to the lower portion of the combustor. Primary “air” (actually a mixture of about 65 percent by weight oxygen and 35 percent recycled flue gas) is fed to the combustor bottom through a grid plate with secondary “air” supplied higher up in the lower combustor region. The combustor is constructed in the same fashion as the Case-1 combustor. The lower combustor region is a tapered rectangular section formed from fusion welded waterwall tubing with a refractory lining. Combustion occurs throughout both the lower and upper combustor, which are filled with bed material. The upper combustor section is a rectangular straight walled section formed from fusion welded waterwall tubing. The combustor is cooled with continuous forced circulation of water from the steam drum. Combustor bed temperature is maintained at an optimum level for sulfur capture and combustion efficiency by balancing heat absorption in the combustor walls and in the MBHE. No additional heat-absorbing surface (i.e. evaporator panels as were used in Case-1) is included in the combustor for Case-2.

Fuel Feed System:

The fuel feed system for Case-2 is very similar to the system used for Case-1. It is designed to transport prepared coal from the storage silos to the lower combustor. The system includes the storage silos and silo isolation valves, fuel feeders, feeder isolation valves, and fuel piping to the furnace. In Case-2 only three feed points are used as compared to six in Case-1 because of the reduced combustor plan area.

Sorbent Feed System:

The limestone feed system for Case-2 is the same as for Case-1. The limestone feed system pneumatically transports prepared limestone from the storage silos to the lower combustor. The system includes the storage silos and silo isolation valves, rotary feeders, blower, and piping from the blower to the furnace injection ports.

Ash Coolers:

The ash cooler design for Case-2 is the same as for Case-1 as the ash flow is nearly identical to Case-1. Draining hot solids through 2 water-cooled ash coolers controls solids inventory in the system while recovering heat from the hot ash. The cooling water used for the ash cooler is provided by feedwater from the final extraction feedwater heater of the steam cycle. The heated water leaving the ash cooler is then combined with water from the economizer located in the convection pass to feed the steam drum.

Cyclones:

Flue gas and entrained solids exit the upper combustor and enter the cyclones. Two relatively small cyclones were selected for Case-2. Although a single cyclone could have been selected for this case because of the reduced gas flow resulting from the oxygen firing, two smaller cyclones were selected to reduce overall unit height. The cost savings

associated with a height reduction were greater than the cost increase for two cyclones instead of one. The cyclones are shaped like cylindrical cones constructed from steel plate with a multiple layer refractory lining. Solids are separated from the flue gas in the cyclones and fall into seal pots. Well over 99 percent of the entrained solids are captured in the cyclones. Two 17 ft diameter cyclones were selected for this case.

Seal Pots:

The seal pots for Case-2 are of the same design as in Case-1 although smaller in size since fewer solids are recirculated due to the reduced gas flow for Case-2. The seal pot is a device that provides a pressure seal between the combustor, which is at relatively high pressure (~ 40 inwg at the bottom), and the cyclone that is at near atmospheric pressure. The seal pot is a non-mechanical valve, which moves solids collected in the cyclone back to the combustor. The seal pot is constructed of steel plate with a multiple layer refractory lining with fluidizing nozzles located along the bottom to assist solids flow. Some of the solids flow directly from the seal pot back to the combustor while other solids are diverted through a plug valve, through the external Moving Bed Heat Exchangers (MBHE), and then back to the combustor.

Moving Bed Heat Exchanger:

The external heat exchanger for Case-2 is a single moving bed rather than 2 fluidized bubbling bed units as were used in Case-1. The moving bed heat exchanger is not fluidized and contains several immersed tube bundles, which cool the hot solids leaving the seal pot before the cooled solids return to the lower combustor. The tube bundles in the MBHE utilize spiral-finned tubes and include superheater, reheater and evaporator sections. Very high heat transfer rates are obtained in the MBHE due to the conduction heat transfer mechanism between the solids and tube. The MBHE is bottom supported and is constructed using steel plate refractory lined enclosure walls. It is rectangular in cross section with a hopper shaped bottom. The solids move through the bed by gravity at a design velocity of about 150 ft/hr. The cooled solids leaving the MBHE are feed to the combustor.

Convection Pass:

The flue gas leaving the cyclones is ducted into the Convection Pass. The convection pass for Case-2 is much smaller than for Case-1 and includes only an economizer section (3 banks) for heat absorption. The cross sectional area of the Case-2 convection pass is about 30 percent of the area required for Case-1. The convection pass is constructed similar to those used for pulverized coal firing with fusion welded steam cooled enclosure walls and fly-ash hoppers located at the bottom. Soot-blowers are used to keep the heat transfer surfaces clean.

Superheater:

The superheater is divided into two major sections, both of which are located in the MBHE. Saturated steam leaving the steam drum first cools the combustor roof and convection pass walls before supplying the horizontal low temperature and finishing superheater sections located in the external moving bed heat exchanger. The superheater sections are both horizontal sections comprised of spiral-finned tubing. There are no superheater banks located in the convection pass for Case-2. The steam leaving the low

temperature section is piped to the de-superheater for superheater outlet temperature control and then to the finishing section. The steam leaving the finishing superheater is piped to the high-pressure turbine where it is expanded to reheat pressure and then returned to the reheat section located in the MBHE.

Reheater:

The reheater, located in the MBHE, is designed as a single section. The steam is supplied to the reheater inlet header from the de-superheating spray station, which is fed from the high-pressure turbine exhaust. The reheater is a horizontal section comprised of spiral-finned tubing and located between the superheat finishing section and the low temperature superheat section. There are no reheater banks located in the convection pass for Case-2. The steam leaving the de-superheating spray station is piped to the reheater located in the external moving bed heat exchanger. The steam leaving the reheater is returned to the intermediate pressure turbine where it continues its expansion through the intermediate and low-pressure turbines before being exhausted to the condenser.

Economizer:

The flue gas leaving the cyclones is then further cooled in an economizer section located in the convection pass. The economizer is comprised of four banks of horizontal tubes, which heats high-pressure boiler feedwater. The water exiting the economizer tube banks then cools the convection pass hanger tubes, which support the economizer sections, before it is supplied to the steam drum. The feedwater supplying the economizer is piped from the final extraction feedwater heater and the ash cooler.

Oxygen Heater:

A tubular regenerative oxygen heater is used to cool the flue gas leaving the economizer by heating both the primary and secondary “air” streams prior to combustion in the furnace. The primary and secondary “air” is actually a mixture of about 65 percent by weight oxygen with recycled flue gas.

Baghouse:

Particulate matter for Case-2 is removed from the cooled flue gas leaving the oxygen heater in a baghouse. The baghouse for Case-2 is much smaller than for Case-1 due to the reduced gas flow (about 30 percent of the Case-1 flow). The ash collected in the baghouse is supplied to the ash handling system.

Parallel Feedwater Heater:

The Parallel Feedwater Heater (PFWH) of Case-2 is used to recover additional heat in the steam cycle for this case as shown in Figure 2.2.5. The feedwater flow is in parallel with the bottom two extraction feedwater heaters included in the steam cycle. The PFWH is used because in Case-2 the gas temperature leaving the Oxygen Heater is significantly higher than the gas temperature leaving the Air Heaters of Case-1 (313 °F vs. 275 °F) and some of this energy can be economically recovered. This occurs because the ratio of air to gas in the air heater is higher in Case-1 than is the ratio of oxidant to gas in the oxygen heater of Case-2. This causes the Air heater of Case-1 to be more effective than the Oxygen Heater of Case-2. The PFWH heat exchanger is located in the flue gas stream following the baghouse and is constructed similarly to economizer heat exchanger banks

used in Heat Recovery Steam Generator units. The tubes used are heavily finned since the gas is clean. The enclosure walls are constructed with insulated steel liners.

Figure 2.2. 5: Case-2 Parallel Feedwater Heater Arrangement

Gas Cooler:

The gas cooler of Case-2 is used to cool the flue gas leaving the PFWH to as low a temperature as possible in order to minimize the power requirements of the Draft System of the boiler and the Gas Processing System which produces the CO₂ product. The Gas Cooler is a direct contact, water spray type of system. Some of the water vapor contained within the flue gas also condenses out in this cooler. This cooler is designed to cool the flue gas to 100 °F. A detailed description of this process and equipment is contained in Section 2.2.2 as the Gas Cooler System is considered part of the Gas Processing System.

Draft System:

The flue gas is moved through the Boiler Island equipment with the draft system. The draft system includes the gas recirculation fan, the fluidizing gas blower, the induced draft (ID) Fan, the associated ductwork, and expansion joints. The induced draft fan, gas recirculation fan, and fluidizing gas blower are driven with electric motors and are controlled to operate the unit in a balanced draft mode with the cyclone inlets maintained at a slightly negative pressure (typically, -0.5 inwg).

2.2.2. Case-2 Gas Processing System Process Description and Equipment

The purpose of this system is to process the flue gas stream leaving the oxygen-fired Boiler Island to provide a liquid CO₂ product stream of suitable purity for sequestration or usage.

The Case-2 CO₂ capture system is designed for about 94 percent CO₂ capture. Cost and performance estimates were developed for all the systems and equipment required to cool, purify, clean, compress and liquefy the CO₂, to a product quality acceptable for pipeline transport. The Dakota Gasification Company's CO₂ specification for EOR, given in Table 2.0.1, was used as the basis for the CO₂ capture system design.

A very low concentration of oxygen, in particular, is specified for meeting current pipeline operating practices, due to the corrosive nature of the oxygen. Hence, for Case-2, whereby the final CO₂ liquid product was found to contain about 11,400 ppmv of O₂, the design of the transport pipe to an EOR site for example would have to take this characteristic under consideration.

The nitrogen concentration specified in Table 2.0.1 is < 300 ppmv. It should be noted that according to Charles Fox of Kinder Morgan (Fox, 2002), this specification is very conservative as his company specifies a maximum nitrogen concentration of 4 percent (by volume) to control the minimum miscibility pressure. In Case 2 the nitrogen concentration in the liquid product was 11,800 ppmv. The exact reasoning behind the very low nitrogen specification listed in Table 2.0.1 is not clear.

2.2.2.1 Process Description

The following describes a CO₂ recovery system that cools and then compresses a CO₂ rich flue gas stream from an oxygen-fired CFB boiler to a pressure high enough so CO₂ can be liquefied. The resulting liquid CO₂ is passed through a CO₂ Stripper to reduce the N₂/O₂ content to a level that is optimum with respect to energy consumption. Then the liquid CO₂ is pumped to a high pressure (2,000 psig) so it can be economically transported for sequestration or usage. Pressure in the transport pipeline will be maintained above the critical pressure of CO₂ to avoid 2-phase flow. The overhead gas from the CO₂ Stripper which contains about 39 percent by volume CO₂, 36 percent N₂, and 25 percent O₂, is vented to atmosphere.

The key process parameters (pressures, temperatures, duties etc.) are shown in the material and energy balance tables shown in Section 2.2.2.3 and will not be repeated here except in selected instances.

Figure 2.2.6 (Refer to Section 2.2.2.2) shows the Flue Gas Cooling process flow diagram and Figure 2.2.7 shows the Flue Gas Compression and Liquefaction process flow diagram.

Flue Gas Cooling:

Please refer to Figure 2.2.6 (Drawing D 12173-02001-0).

The feed to the Gas Processing System is the flue gas stream that leaves the PFWH of the Boiler Island. At this point, the flue gas is near the dew point of H₂O. All of the flue gas leaving the boiler is cooled to 100 °F in Gas Cooler DA-101 which operates slightly below atmospheric pressure. A significant amount of water condenses out in this cooler. Excess condensate is blown down to the cooling water system. A single vessel has been provided for this cooler.

The Gas Cooler is configured in a packed tower arrangement where the flue gas is contacted with cold water in countercurrent fashion. Warm water from the bottom of the contactor is recycled back to the top of the contactor by Water Pump GA-101 after first cooling it in an external water cooled heat exchanger, Water Cooler EB-101 (plate and frame exchanger). The cooling water for this exchanger comes from the new cooling tower.

Because the flue gas may carry a small amount of fly ash, the circulating water is filtered

in Water Filter FD-101A-E to prevent solids build-up in the circulating water. Condensate blowdown is filtered and is taken out downstream of the filter. However, the stream is not cooled and is split off before EB-101. Thus the heat load to the cooling tower is minimized.

From the Gas Cooler the gas stream then is boosted in pressure by the ID fan followed by a split of the gas into two streams. This design was developed to minimize the length of ducting operating at a slight vacuum and to minimize the temperature of the gas being recycled back to the boiler. The mass flow rate of the gas recirculation stream is about 52 percent of the flow rate of the product gas stream, which proceeds to the gas compression area. The recycle stream is sized to provide an oxygen content of about 70 percent by volume in the oxidant stream supplying the boiler. The Gas Cooler minimizes the volumetric flow rate to and the resulting power consumption of the Flue Gas Compression equipment located downstream and the gas recirculation fans in the Boiler Island.

Three-Stage Gas Compression System:

Please refer to Figure 2.2.7 (Drawing D 12173-02002-0).

The compression section, where CO₂ is compressed to 365 psig by a three-stage centrifugal compressor, includes Flue Gas Compressor GB-101. After the aftercoolers, the stream is then chilled in a propane chiller to a temperature of -21 °F. Note that both the trim cooling water and water for the propane condenser comes from the cooling tower. At this pressure and temperature, about 80 mole percent of the stream can be condensed. The flash vapors contain approximately 80 weight percent of the inlet oxygen and nitrogen, but also about 13.7 weight percent of the CO₂. Therefore, a rectifier tower has been provided to reduce the loss of CO₂ to an acceptable level (about 6 weight percent). Then the pressure of the liquid is boosted to 2,000 psig by CO₂ Pipeline Pump GA-103. This stream is now available for sequestration or usage.

The volumetric flow to the compressor inlet is about 69,000 ACFM and only a single frame is required. The discharge pressures of the stages have been balanced to give reasonable power distribution and discharge temperatures across the various stages. They are:

- 1st Stage 28 psig
- 2nd Stage 108 psig
- 3rd Stage 365 psig

Power consumption for this large compressor has been estimated assuming adiabatic efficiency of 75 percent.

The hot gas from each stage is first cooled in an air cooler to 120 °F (Flue Gas Compressor 1st/ 2nd / 3rd Stage Aftercooler EC-101/2/3) and then further cooled by a water-cooled heat exchanger to 95 °F (Flue Gas Compressor 1st/ 2nd Stage Trim Cooler EA-101/2). The flue gas compressor 3rd stage cooler (EA-103) cools the gas to 90 °F to reduce the size of the dryers. Due to their large size, many of these heat exchangers consist of multiple shells. Because of highly corrosive conditions, the process side of the coolers must be stainless steel.

Because the flue gas stream leaving DA-101 is saturated, some water condenses out in the three aftercoolers. The sour condensate is separated in knockout drums (FA-100/1/2/3) equipped with mist eliminator pads. Condensate from these drums is drained to the cooling tower or to waste water treatment. To prevent corrosion, these drums have a stainless steel liner.

Gas Drying:

Please refer to Figure 2.2.7 (Drawing D 12173-02002-0).

It is necessary to dry the CO₂ stream to meet the product specification. Flue gas leaving the 3rd stage discharge knockout drum (FA-103) is fed to Flue Gas Drier FF-101 A/B where additional moisture is removed. An alumina drier has been selected.

The performance of a fixed-bed drier improves as pressure increases. This favors locating the drier at the discharge of the compressor. However, as the operating pressure of the drier increases, so does the design pressure of the equipment. This favors low-pressure operation. But, at low pressure the diameter or number of the drier vessels grows, increasing the cost of the vessel. Having to process the recycle gas from the rectifier condenser cooling would also increase the diameter of the vessel. However, this is less than 10 percent of the forward flow. For this design the drier has been optimally located downstream of the 3rd stage compressor. The CO₂ Drier system consists of two vessels; FF-101 A/B. One vessel is on line while the other is being regenerated. Flow direction is down during operation and up during regeneration.

The drier is regenerated with the non-condensable vent gas from the rectifier after it exits the third stage discharge trim cooler in a simple once through scheme. During regeneration, the gas is heated in Regeneration Heater FH-101 before passing it through the exhausted drier. After regeneration, heating is stopped while the gas flow continues. This cools the bed down to the normal operating range. The regeneration gas and the impurities contained in it are vented to the atmosphere.

Regeneration of an alumina bed requires relatively high temperature and, because HP steam pressure may fluctuate, a gas-fired heater has been specified for this service.

Flue Gas Filter FD-102 has been provided at the drier outlet to remove any fines that the gas stream may pick up from the desiccant bed.

CO₂ Condensation and Stripping:

Please refer to Figure 2.2.7 (Drawing D 12173-02002-0).

From the CO₂ Drier, the gas stream is cooled down further to -21 °F with propane refrigeration in CO₂ Condenser EA-104 A-F. From EA-104 the partially condensed flue gas stream continues onto CO₂ Rectifier DA-102.

At this pressure and temperature, 80 mole percent of the stream can be condensed. The flash vapors contain approximately 80 weight percent of the inlet oxygen and nitrogen, but also 12 weight percent of the CO₂. Therefore, as mentioned, a rectifier tower has been provided to reduce the loss of CO₂ to an acceptable level. The pressure of the liquid is boosted to 2,000 psig by CO₂ Pipeline Pump GA-103 for delivery to a sequestration or usage location.

The vapors in the feed to the rectifier contain the nitrogen and the oxygen that flashed from the liquid CO₂. To keep the CO₂ loss to the minimum, the rectifier also has an overhead condenser, CO₂ Rectifier Condenser EA-107. This is a floodback type condenser installed on top of the Rectifier. It cools the overhead vapor from the tower down to –48 °F. The condensed CO₂ acts as cold reflux in the CO₂ Rectifier.

Taking a slipstream from the inert-free liquid CO₂ from the Rectifier bottoms and letting it down to the Flue Gas Compressor 3rd stage suction pressure cools EA-107. At this pressure, CO₂ liquid boils at –55 °F thus providing the refrigeration necessary to condense some of the CO₂ from the Stripper overhead gas. The process has been designed to achieve at least 94 percent CO₂ recovery. The vaporized CO₂ from the cold side of EA-107 is fed to the suction of the Flue Gas Compressor 3rd stage.

Any system containing liquefied gas such as CO₂ is potentially subject to very low temperatures if the system is depressurized to atmospheric pressure while the system contains cryogenic liquid. If the CO₂ Rectifier (and all other associated equipment that may contain liquid CO₂) were to be designed for such a contingency, it would have to be made of stainless steel. However, through proper operating procedures and instrumentation such a scenario can be avoided and low temperature carbon steel (LTCS) can be used instead. Our choice here is LTCS. However, the condenser section will be made from stainless steel.

CO₂ Pumping and CO₂ Pipeline:

Please refer to Figure 2.2.7 (Drawing D 12173-02002-0).

The CO₂ product must be increased in pressure to 2,000 psig. A multistage heavy-duty pump (GA-103) is required for this service. This is a highly reliable derivative of an API-class boiler feed-water pump.

It is important that the pipeline pressure be always maintained above the critical pressure of CO₂ such that single-phase (dense-phase) flow is guaranteed. Therefore, pressure in the line should be controlled with a pressure controller and the associated control valve located at the destination end of the line.

Offgas:

Please refer to Figure 2.2.7 (Drawing D 12173-02002-0).

The vent gas from the CO₂ Rectifier overhead is at high pressure and there is an opportunity for power recovery using turbo-expanders. Because the gas cools down in the expansion process, there is also an opportunity for cold recovery. Heat recovery from the stream after let down via an expander was examined and it was determined that the amount of duty that could be recovered without the carbon dioxide in the stream freezing was small. Thus heat recovery could not be justified. The offgas leaves the Rectifier at –48 °F approximately. The refrigeration recovery to condense CO₂ was the best use for this cold since it also produces a reasonable temperature regeneration gas for the dryers.

2.2.2.2 Process Flow Diagrams

Two process flow diagrams are shown below for these systems:

- Figure 2.2.6 (Drawing D 12173-02001-0) Flue Gas Cooling PFD

- Figure 2.2.7 (Drawing D 12173-02002-0) CO₂ Compression and Liquefaction PFD

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Figure 2.2. 6: Case-2 Flue Gas Cooling Process Flow Diagram

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Figure 2.2. 7: Case-2 CO₂ Compression and Liquefaction Process Flow Diagram

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2.2.2.3. Material and Energy Balance

Table 2.2.4 contains the overall material and energy balance for the Flue Gas Cooling System and the CO₂ Compression and Liquefaction System for Case-2 described above. It is based on 94 percent recovery of CO₂. Please refer to the Process Flow Diagrams shown in the previous section for stream numbers shown in this table.

Table 2.2. 4: Case-2 Gas Processing System Material & Energy Balance

STREAM NAME	To quench column	From Quench column	Excess water	From Large blowers	Quench water out	Quench water in	To liquefaction train	To boiler	First water KO	To 2nd stage	2nd water KO	To 3rd stage	Recycle from condenser
PFID STREAM NO.	1	3a	6	3b	2	5	3c	3d	7	8	9	10	25
VAPOR FRACTION	Molar	0.989	1.000	0.000		0.000	0.000	0.997	0.997	0.000	1.000	0.000	1.000
TEMPERATURE	°F	136.0	100	118		118	90	100	95	95	86	86	-48
PRESSURE	PSIA	13.7	13.5	55		14	45	13.3	13.3	34	34	117	117
MOLAR FLOW RATE	lbmol/hr	18,158	15,628	2,552		96,583	94,053	10,241	5,388	451	9,789	13	10,400
MASS FLOW RATE	lb/hr	682,970	637,380	45,992		1,740,600	1,695,000	418,390	218,990	8,141	410,250	231	441,530
ENERGY	Btu/hr	8.14E+07	6.84E+07	-3.59E+07		-1.36E+09	-1.37E+09	4.42E+07	2.42E+07	-6.54E+06	4.21E+07	-1.87E+05	4.30E+07
COMPOSITION	Mol %												
CO ₂		71.38%	82.93%	0.02%		0.02%	83.20%	83.20%	0.08%	87.03%	0.30%	89.32%	97.54%
H ₂ O		4.90%	5.69%	0.00%		0.00%	6.75%	6.75%	99.92%	2.46%	99.68%	0.59%	0.00%
Nitrogen		3.62%	4.21%	0.00%		0.00%	5.71%	5.71%	0.00%	5.97%	0.00%	5.71%	1.18%
Ammonia		20.00%	7.05%	99.98%		99.98%	99.98%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Oxygen		0.10%	0.12%	0.00%		0.00%	0.00%	4.22%	4.22%	0.00%	4.42%	0.00%	4.25%
SO ₂		0.00%	0.00%	0.00%		0.00%	0.00%	0.12%	0.12%	0.01%	0.12%	0.02%	0.14%
VAPOR													
MOLAR FLOW RATE	lbmol/hr	17,949.8	15,628.1	-		-	10,208.8	5,419.4	-	9,789.2	-	10,399.8	790.0
MASS FLOW RATE	lb/hr	679,220	637,380	-		-	417,820	219,560.0	-	410,253	-	441,530	34,533
STD VOL. FLOW	MMSCFD	163.48	142.33	-		-	92.98	49.4	-	89.16	-	94.72	7.20
ACTUAL VOL. FLOW	ACFM	138,840.00	114,250.00	-		-	69,181.03	45,069.0	-	28,427	-	8,390.08	447.36
MOLECULAR WEIGHT	MW	37.84	40.78	-		-	40.93	40.93	-	41.91	-	42.46	43.71
DENSITY	lb/ft ³	0.08	0.09	-		-	0.10	0.10	-	0.24	-	0.88	1.29
VISCOSITY	cP	0.0145	0.0150	-		-	0.0150	0.0150	-	0.0155	-	0.0156	0.0114
LIGHT LIQUID													
MOLAR FLOW RATE	lbmol/hr	-	-	-		-	-	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	-	-		-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-		-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-		-	-	-	-	-	-	-	-
DENSITY	lb/ft ³	-	-	-		-	-	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	-	-	-		-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-		-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-		-	-	-	-	-	-	-	-
HEAVY LIQUID													
MOLAR FLOW RATE	lbmol/hr	208.09	-	2,552.00		96,583	94,053	31.72	451.34	-	12.77	-	-
MASS FLOW RATE	lb/hr	3,750	-	45,991.83		1,740,600	1,695,000	571.64	8,141.38	-	231.26	-	-
STD VOL. FLOW	BPD	257	-	3,156		119,430	116,300	39	559	-	16	-	-
ACTUAL VOL. FLOW	GPM	7.62	-	92.76		3,510.88	3,378.10	1.14	16.26	-	0.46	-	-
DENSITY	lb/ft ³	61.32	-	61.82		61.81	62.56	62.29	62.44	-	62.74	-	-
VISCOSITY	cP	0.4793	-	0.5654		0.5657	0.7606	0.6799	0.7185	-	0.8176	-	-
SURFACE TENSION	Dyne/Cm	66.34	-	68.11		68.12	70.83	69.86	70.31	-	70.97	-	-
STREAM NAME	To drier	3rd water ko	From drier/ Condenser inlet	Condenser outlet	Non-condensable vent	Rectifier bottoms to condenser	CO ₂ to pipeline	Refrig compressor discharge	Refrig condenser out	Refrig subcooler out	Refrig to CO ₂ condenser	Refrig from CO ₂ condenser	Warm non condensable
PFID STREAM NO.	12	11	14	15	24	22	21	100	101	102	103	104	26
VAPOR FRACTION	Molar	1.000	0.000	1.000	0.206	1.000	0.130	0.000	1.000	0.000	0.000	1.000	1.000
TEMPERATURE	°F	90	90	-23	-48	-56	82	167	110	44	-26	-23	69
PRESSURE	PSIA	372	372	369	364	361	120	2,015	222	215	212	23	356
MOLAR FLOW RATE	lbmol/hr	10,366.29	33.47	10,337.65	10,087.65	1,358.89	790.00	8,209.08	9,100.00	9,100.00	9,100.00	9,100.00	1,358.89
MASS FLOW RATE	lb/hr	440,920	612	440,410	429,760	47,924	34,533	358,830	401,280	401,280	401,280	401,280	47,924
ENERGY	Btu/hr	4.01E+07	-4.83E+05	4.00E+07	-1.67E+07	3.58E+06	-2.22E+06	-2.76E+06	7.00E+07	6.77E+06	-1.13E+07	-1.13E+07	4.45E+07
COMPOSITION	Mol %												
CO ₂		89.61%	0.87%	89.86%	89.86%	39.12%	97.54%	0.00%	0.00%	0.00%	0.00%	0.00%	39.12%
H ₂ O		0.28%	99.06%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Nitrogen		5.73%	0.00%	5.74%	5.74%	35.89%	1.18%	0.00%	0.00%	0.00%	0.00%	0.00%	35.89%
Ammonia		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	100.00%	100.00%	100.00%	100.00%	100.00%	0.00%
Oxygen		4.26%	0.00%	4.27%	4.27%	24.99%	1.14%	0.00%	0.00%	0.00%	0.00%	0.00%	24.99%
SO ₂		0.12%	0.08%	0.12%	0.12%	0.00%	0.14%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
VAPOR													
MOLAR FLOW RATE	lbmol/hr	10,366.3	-	10,337.7	2,075.3	1,358.9	102.5	-	9,100.0	-	-	2,044.7	9,100.0
MASS FLOW RATE	lb/hr	440,920	-	440,410	79,499	47,924	4,282	-	401,280	-	-	90,163	401,280
STD VOL. FLOW	MMSCFD	94.41	-	94.15	18.90	12.38	0.93	-	82.88	-	-	18.62	82.88
ACTUAL VOL. FLOW	ACFM	2,410.72	-	2,425.14	373.67	241.80	56.04	-	3,758.16	-	-	6,717.44	30,084.4
MOLECULAR WEIGHT	MW	42.53	-	42.60	38.31	35.27	41.77	-	44.10	-	-	44.10	44.10
DENSITY	lb/ft ³	3.05	-	3.03	3.55	3.30	1.27	-	1.78	-	-	0.22	0.22
VISCOSITY	cP	0.0165	-	0.0165	0.0145	0.0147	0.0117	-	0.0103	-	-	0.0066	0.0066
LIGHT LIQUID													
MOLAR FLOW RATE	lbmol/hr	-	-	-	8,012.37	-	687.48	8,209.08	-	9,100.00	9,100.00	7,055.35	-
MASS FLOW RATE	lb/hr	-	-	-	350,260	-	30,251	358,830	-	401,280	401,280	311,110	-
STD VOL. FLOW	BPD	-	-	-	28,971	-	2,506	29,682	-	54,230	54,230	42,045	-
ACTUAL VOL. FLOW	GPM	-	-	-	659.06	-	52.70	895.70	-	1,735.91	1,533.76	1,091.12	-
DENSITY	lb/ft ³	-	-	-	66.26	-	71.57	49.95	-	28.82	32.62	35.55	-
MOLECULAR WEIGHT	MW	-	-	-	43.72	-	44.00	43.71	-	44.10	44.10	44.10	-
VISCOSITY	cP	-	-	-	0.1641	-	0.2239	0.0559	-	0.0835	0.1198	0.1765	-
SURFACE TENSION	Dyne/Cm	-	-	-	15.36	-	20.19	0.88	-	4.81	9.14	14.09	-
HEAVY LIQUID													
MOLAR FLOW RATE	lbmol/hr	-	33.47	-	(0.00)	-	(0.00)	-	-	-	-	(0.01)	-
MASS FLOW RATE	lb/hr	-	611.58	-	-	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	42	-	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	1.21	-	-	-	-	-	-	-	-	-	-
DENSITY	lb/ft ³	-	62.84	-	-	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	0.7751	-	-	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	70.20	-	-	-	-	-	-	-	-	-	-

2.2.2.4. Gas Processing System Utilities

The following tables define the cooling water, natural gas, and electrical requirements for the Gas Processing System.

Table 2.2. 5: Case-2 Gas Processing System Cooling Water and Fuel Gas Requirements

COOLING WATER							
REV	Equipment TAG NO	SERVICE	No. Installed	DUTY MMBTU/HR	INLET TEMP, F	OUTLET TEMP, F	FLOWRATE LB/HR
	EA-101	FG Comp 1 stg trim cooler	1	7.09	85	103	393,939
	EA-102	FG Comp 1 stg trim cooler	1	3.73	85	103	207,071
	EA-103	FG Comp 1 stg trim cooler	1	3.82	85	103	212,121
	EA-201	Refrig Condenser	1	63.18	85	100	4,212,121
	EB-101	Water Cooler	1	50.00	85	105	2,500,000
TOTAL COOLING WATER				127.82			7,525,253

FUEL GAS FUEL GAS VALUE BASIS: 930 BTU/SCF (LHV)										
REV	Equipment TAG NO	SERVICE	ONLINE FACTOR	COMPR HP	HEAT RATE BTU/HP-HR	DUTY MMBTU/HR	EFFICIENCY %	FLOWRATE (Peak)		FLOW (Avg)
	FH-101	Mole sieve regeneration	61%			4.60	80%	MMSCFD	SCFH	MMSCFD
TOTAL FUEL GAS						4.60		0.148	6,183	0.091

Table 2.2. 6: Case-2 Gas Processing System Electrical Requirements

Number of Trains		Item Number	Service	Number Operating per train	Power (ea) including 0.95 motor eff	Total all trains
					(kW)	(kW)
1	EC-101	Flue Gas Compressor 1st Stage Aftercooler		1	78	78
1	EC-102	Flue Gas Compressor 2nd Stage Aftercooler		1	76	76
1	EC-103	Flue Gas Compressor 3rd Stage Aftercooler		1	67	67
1	GB-101	Stage 1		1	5434	5434
1		Stage 2		1	6531	6531
1		Stage 3		1	5997	5997
1	GB-102	Stage 1		1	4947	4947
1		Stage 2		1	2961	2961
1	GA-101	Water pump		1	130	130
1	GA-103	CO2 Pipeline pump		1	685	685
Total						26905

2.2.2.5. Gas Processing System Equipment

The equipment list for the Gas Processing System is provided in Appendix I, Section 9.1.2.2.

2.2.3. Case-2 Air Separation Unit Process Description and Equipment

This section presents the process requirements for the warm end and cold box for the air separation plant. It will be designed to produce nominally 4000 tons per day (TPD) of Oxygen. This same design was also used for Case 3 and Case 4 Air Separation Units. Case-3 is Identical to Case-2 and Case-4 operates at a slightly higher output level.

Assumptions:

The following basic assumptions have been used in this design:

- No special low voltage starting equipment.
- Design based on customer inquiry dated 10/11/02 from Alstom Power.

Design Basis:

The ambient conditions presented in Table 2.2.7 below were used to evaluate performance and to generate the utility summary.

Table 2.2. 7: Ambient Conditions

Item	Value	Units
Barometric Pressure	14.7	psia
Dry Bulb Temperature	80	°F
Hot Day Temperature	95	°F
Cold Day Temperature	20	°F
Wet Bulb Temperature	52	°F
Cooling Water Temperature	90	°F

Production Rates and Purities:

The production rate indicated in Table 2.2.8 below is the net flow-rate from the Air Separation Unit's Cold Box.

Table 2.2. 8: Production and Purity

Plant Site	Oxygen Flow	Pressure @ B/L	Purity
	(Metric T/D Contained O ₂)	(psia)	(%O ₂)
Southeast Texas	3,568	18.0	99.0

2.2.3.1. Process Description and Process Flow Diagrams

The process description below refers to the Process Flow Diagram shown below in Figure 2.2.8 below.

GREENHOUSE GAS EMISSIONS CONTROL BY OXYGEN FIRING IN
CIRCULATING FLUIDIZED BED BOILERS – PHASE II

Figure 2.2. 8: Case-2 Air Separation Unit Process Flow Diagram

Air Compression:

Ambient air is drawn through the air suction filter house (ASFH) for the removal of large airborne particles prior to entering the three stage main air compressor (MAC). The filtered air is compressed in the MAC and then flows through the two-stage direct contact aftercooler (DCA). Air is cooled by exchanging heat with cooling water in the first stage and with chilled water provided by a mechanical chiller in the second stage.

Pre-purification:

The after-cooled air is then passed through the pre-purification system. The pre-purification system uses a two bed temperature-swing adsorption (TSA) process that allows continuous operation. One bed purifies the feed air while the other bed is being regenerated with first hot then cool waste nitrogen. A natural gas regeneration heater provides regeneration energy. The pre-purifier beds utilize a split adsorbent design (molecular sieve and alumina) to remove water, carbon dioxide, and most of the hydrocarbons from the air stream. After pre-purification, the air stream is passed through a dust filter to remove any solid particles.

Air Feed Streams:

The cold box requires one air feed stream. This stream is sent through the Primary Heat Exchanger (PHX) and then split into three streams. One stream is fed to the bottom of the lower column. The second air stream is fed to the oxygen boiler. The third air stream (turbine air) is cooled partially in the PHX and fed to the turbine. Adjusting the turbine airflow can modulate the total amount of refrigeration generated by the cold box.

Cold Box:

The air stream to the oxygen boiler is cooled and condensed against product oxygen and sent to both the upper and lower column.

The turbine air stream is cooled against warming nitrogen and oxygen streams. It is drawn from an intermediate location between the warm leg and the cold leg of the PHX. It is then expanded and cooled in the upper column turbine. The UCT stream enters two thirds of the way down the upper (low-pressure) distillation column.

The air entering the lower column is separated into nitrogen at the top and oxygen-enriched air (kettle liquid) at the bottom. The nitrogen at the top of the column is condensed in the main condenser against boiling oxygen from the upper column. A portion of the condensed nitrogen from the main condenser is used as reflux for the lower column. The remainder is subcooled in the cross flow passages in the nitrogen superheater section of the PHX against warming gaseous nitrogen streams from the upper column. This subcooled liquid nitrogen stream then enters the top of the upper column as reflux. The kettle liquid is subcooled in the cross flow passes of the nitrogen superheater section of the PHX and then enters the upper about 2/3 of the way down the column.

The upper column produces high purity liquid oxygen (>99.0 percent O₂) in the bottom. The upper column also produces waste nitrogen from the top. The gaseous nitrogen stream is warmed in all sections of the PHX to near-ambient temperatures. The product

oxygen is boiled in the oxygen boiler against the condensing air stream and exits as product.

Products:

Gaseous oxygen is available at pressure directly from the cold box and delivered to the battery limit at 3 psig.

2.2.3.2. Material and Energy Balance

Table 2.2.9 shows the Air Separation Unit material and energy balance for Case-2. The stream numbers shown at the top of each column of the table refer to stream numbers shown in the Case-2 PFD for the Air Separation Unit (Figure 2.2.8). Specific energy consumption for this ASU is calculated to be about 231 kWh/ton of O₂.

Table 2.2.9: Case-2 Air Separation Unit Energy and Material Balance

Stream	A4	A5-1	A5-2	A6-1	A6-2
Name	Air to PHX	Air to Turbine	Turbine Air to UC	Air to O2 Boiler and LC	Air to LC
Vapour Fraction	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature (K)	300.0	185.0	128.8	104.9	104.9
Pressure (psia)	89.00	86.60	21.00	86.00	86.00
Molar Flow (kCFH-NTP)	19,050	857.0	857	18,194	12,860
Mass Flow (lb/h)	142,600	64,180	64,180	1,362,000	962,700
Liquid Volume Flow (USGPM)	3,275.0	147.4	147.4	3,128.1	2,211.0
Heat Flow (Btu/hr)	183,500,000	5,010,000	3,510,000	54,990,000	38,900,000
Comp Mole Frac (Nitrogen)	0.7811	0.7811	0.7811	0.7811	0.7811
Comp Mole Frac (Argon)	0.0093	0.0093	0.0093	0.0093	0.0093
Comp Mole Frac (Oxygen)	0.2096	0.2095	0.2096	0.2096	0.2096
Stream	A7-1	A7-2	A8-1	A8-2	K1
Name	Air to O2 Boiler	Boiler Air to LC	Boiler Air to N2SH	N2SH Air to UC	Ket Liq to N2SH
Vapour Fraction	1.0000	0.1279	0.0000	0.0000	0.0000
Temperature (K)	104.9	98.5	98.5	94.6	100.4
Pressure (psia)	86.00	86.20	86.20	86.20	84.50
Molar Flow (kCFH-NTP)	5,335	1,834	3,500	3,500	7,746
Mass Flow (lb/h)	399,300	137,100	262,300	262,300	595,600
Liquid Volume Flow (USGPM)	917.1	315.9	601.2	601.2	1,283.0
Heat Flow (Btu/hr)	16,090,000	-3,836,000	-9,187,000	-10,730,000	-23,270,000
Comp Mole Frac (Nitrogen)	0.7811	0.7911	0.7756	0.7756	0.5946
Comp Mole Frac (Argon)	0.0093	0.0090	0.0095	0.0095	0.0145
Comp Mole Frac (Oxygen)	0.2096	0.1999	0.2147	0.2147	0.3908
Stream	K2	S1	S2	W1	W2
Name	Ket Liq to UC	Shelf to N2SH	Shelf to UC	Waste N2 to N2SH	Waste N2 to PHX
Vapour Fraction	0.0000	0.0000	0.0000	13.2500	2.2780
Temperature (K)	94.6	95.6	81.3	80.3	99.9
Pressure (psia)	84.50	82.03	82.03	20.00	20.00
Molar Flow (kCFH-NTP)	7,746	6,947	6,947	15,050	15,050
Mass Flow (lb/h)	595,600	504,200	504,200	1,095,000	1,095,000
Liquid Volume Flow (USGPM)	1,283.0	1,244.0	1,244.0	2,695.0	2,695.0
Heat Flow (Btu/hr)	-26,190,000	-24,640,000	-24,640,000	36,910,000	47,140,000
Comp Mole Frac (Nitrogen)	0.5946	0.9916	0.9916	0.9886	0.9886
Comp Mole Frac (Argon)	0.0145	0.0034	0.0034	0.0094	0.0094
Comp Mole Frac (Oxygen)	0.3908	0.0050	0.0050	0.0020	0.0020
Stream	W3	O1	O2	O3	
Name	WN2 to Regen and Vent	O2 to Oxygen Boiler	O2 to PHX	Oxygen Product	
Vapour Fraction	1.0000	0.0000	1.0000	1.0000	
Temperature (K)	291.3	93.7	94.4	291.3	
Pressure (psia)	17.00	21.00	22.50	18.00	
Molar Flow (kCFH-NTP)	15,050	3,998	3,998	3,998	
Mass Flow (lb/h)	1,095,000	331,500	331,500	331,500	
Liquid Volume Flow (USGPM)	2,695.0	580.8	580.8	580.8	
Heat Flow (Btu/hr)	141,200,000	-18,080,000	11,660,000	37,530,000	
Comp Mole Frac (Nitrogen)	0.9886	0.0000	0.0000	0.0000	
Comp Mole Frac (Argon)	0.0094	0.0090	0.0090	0.0090	
Comp Mole Frac (Oxygen)	0.0020	0.9910	0.9910	0.9910	

2.2.2.3. Air Separation Unit Utility Summary

The following tables show the expected Electricity, Water and Natural Gas use for the ASU. The utilities presented here are for the quantity of oxygen specified in Table 2.2.9.

- Electric Power:

Components	kW
BLAC	36,600
Turbine	(445)
Water Chiller	1,400
DCA Pumps	170
Misc. (Incl. Lube Oil)	75
Total Average Power (+/- 5%)	37,800

- Estimated Cooling Water Flow:

Estimated Cooling Water Rise: 15 F	
Total Average CW Flow, gpm	23,150

- Estimated Natural Gas Flow:

Natural Gas used for 1/3 of time	
Natural Gas Use (peak), lbm/hr	1,600

2.2.3.4. Air Separation Unit Equipment

Equipment for the Air Separation Unit is described in Appendix I, Section 9.1.2.3.

2.2.3.5. Air Separation Unit Miscellaneous Items

This section covers items such as availability, chemical requirements and operating manpower for the Air Separation Unit.

Plant Availability:

In the broad sense, “availability” means the fraction of time a unit is able to supply product at various capacity levels. Availability has two components, those related to planned outages and those related to unplanned outages.

The historic scheduled outage rate for Praxair designed large plants is less than 0.01

planned outage hours/period hours. Planned outages are normally planned to coincide with customer outages and thereby avoiding any customer impact. This activity can be planned in advance and is typically every five years.

Unplanned outages can occur because of any number of causes such as equipment failure, control system component failures, or an unnoticed degradation of a process value, which has reached a shutdown set point. In most cases, the cause of a shutdown can be rectified quickly, and the plant can be restarted in 1 to 2 hours.

With a long history of operating air separation plants, Praxair has experienced an average unplanned (forced) outage rate of less than 0.01 unplanned outage hours/period hours - availability greater than 99 percent (without backup system and excluding planned maintenance and utility outages). Molecular sieve pre-purifier plant cycles are inherently more reliable than the older reversing heat exchanger plants. The molecular sieve pre-purifier plant cycle proposed herein will demonstrate availability at the higher end of this experience.

A recent snapshot of product availability data from large plants over past 2.5 years showed all product availability of 99.16 percent. This number is not just based upon primary product like oxygen, but includes secondary products like argon and/or merchant liquid product as well. The historical distribution of unplanned outage events from the same group of plants is given below. The distribution provides an indication of how soon the product production may be restarted following an unexpected shutdown.

<u>Duration (hours)</u>	<u>percent of Unplanned Outages</u>
< 4	50 to 70
4 - 8	15 to 25
8 - 12	5 to 10
12 - 16	3 to 8
16 - 24	0 to 5
> 24	0 to 5

The high reliability of Praxair's plants are achieved through the following:

- Selection of equipment/fabrication services from qualified suppliers.
- Application of controls strategies to eliminate unnecessary outages.
- Plant design, layout, and cold box packaging to allow easy maintenance access.
- Application of best practices for plant operation and maintenance to ensure that the facility meets high expectations of safety, availability, efficiency, and regulatory requirements.

Based upon our history and practices availability of this plant is expected to be 99 percent (considering unplanned outages).

Chemical Requirements:

There are no major on-going chemical requirements, as follows:

- Cooling Water is supplied by others, thus major treatment chemicals are part of this supply.
- With a small closed loop cooling system, some minor treatment chemicals will be required.
- Minor consumable items such as analyzer zero span and fuel gas cylinders, as well as, lube oil top-off will be required.
- Pre-purifier adsorbent is included in plant pricing and is typically not replaced.
- To cover minor consumables, approximately \$20,000/year is estimated.

Operating Manpower:

- Operating staff :

Supervisor	1
Plant Engineering/Assistant Manager	1
Operators	6
Maintenance (Mechanical & Instrumentation)	3

- Major maintenance would be staffed externally – either from power plant staff or contractors.

2.2.4. Case-2 Balance of Plant Equipment and Performance

The balance of plant equipment described in this section includes the steam cycle performance and equipment, the draft system equipment, the cooling system equipment, and the material handling equipment (coal, limestone, and ash). Refer to Appendix I for equipment lists and Appendix II for drawings.

2.2.4.1. Steam Cycle Performance

The steam cycle for the Case-2 is shown schematically in Figure 2.2.9. The Mollier diagram which illustrates the process on enthalpy - entropy coordinates is the same as for Case-1 and is not repeated here. The high-pressure turbine expands 1,400,555 lbm/hr of steam at 1,800 psia and 1,005 °F. Reheat steam (1,305,632 lbm/hr) is heated and returned to the intermediate pressure turbine at 469 psia and 1,005 °F. The condenser pressure used for Case-2 and all other cases in this study was 3.0 in. Hga.

The steam cycle starts at the condenser hot well, which is a receptacle for the condensed steam from the exhaust of the steam turbine. The condensate flows to the suction of the condensate pumps, which increase the pressure of the fluid by a nominal 250 psi to transport it through the piping system and enable it to enter the open contact heater, or deaerator. The condensate passes through a gland steam condenser (SPE) first, followed in series by two low-pressure extraction feedwater heaters. The heaters successively increase the condensate temperature to a nominal 221°F by condensing and partially sub-cooling steam extracted from the LP steam turbine section. Each heater receives a separate extraction steam stream at successively higher pressure and temperature. The

condensed steam (now referred to as heater drains) is progressively passed to the next lower pressure heater, with the drains from the lowest heater draining to the condenser. The Case-2 condensate and feedwater system differs from Case 1 in that there is a parallel low-pressure Feedwater Heater (PFWH - heated by flue gas) in a parallel feedwater stream with the two low-pressure extraction feedwater heaters as shown in Figure 2.2.9

The condensate entering the deaerator is heated and stripped of noncondensable gases by contact with the steam entering the unit. The steam is condensed and, along with the heated condensate, flows by gravity to a deaerator storage tank. The boiler feedwater pumps take suction from the storage tank and increase the fluid pressure to a nominal 2200 psig. Both the condensate pump and boiler feed pump are electric motor driven. The boosted condensate flows through three more high-pressure feedwater heaters, increasing in temperature to 470°F at the entrance to the boiler economizer section. Each heater receives a separate extraction steam stream at successively higher pressure and temperature. The condensed steam (drains) is progressively passed to the next lower pressure heater, with the drains from the lowest heater draining to the deaerator.

Within the CFB boiler the feedwater is evaporated and finally superheated. The high-pressure superheated steam leaving the finishing superheater (1,400,555 lbm/hr of steam at 1,815 psia and 1,000 °F) is expanded through the high-pressure turbine. Reheat steam (1,305,632 lbm/hr) is heated and returned to the intermediate pressure turbine at 469 psia and 1,000 °F. These conditions (temperatures, pressures) represent common steam cycle operating conditions for existing utility scale CFB power generation systems in use today. The reheated steam expands through the intermediate and low-pressure turbines before exhausting to the condenser. The condenser pressure used for Case-1 and all other cases in this study was 3.0 in. Hga.

The steam turbine performance analysis results show the generator produces 209,907 kW output and the steam turbine heat rate is about 8,256 Btu/kWh. The generator output, turbine heat rate and condenser losses are slightly higher for Case-2 than for Case-1. This is a result of the PFWH, which reduces extraction flows to the low-pressure extraction feedwater heaters and increases LP turbine power output.

Figure 2.2. 9: Case-2 Steam Cycle Schematic and Performance

2.2.4.2. Steam Cycle Equipment

This section provides a brief description of the major equipment (steam turbine, condensate and feedwater systems) utilized for the steam cycle of this case.

Steam Turbine:

The turbine consists of a high-pressure (HP) section, intermediate-pressure (IP) section, and one double-flow low-pressure (LP) section, all connected to the generator by a common shaft. Main steam from the boiler passes through the stop valves and control valves and enters the turbine at 1,800 psig / 1,000°F. The steam initially enters the turbine near the middle of the high-pressure span, flows through the turbine, and returns to the boiler for reheating. The reheated steam flows through the reheat stop valves and intercept valves and enters the IP section at 465 psig / 1,000°F. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. The steam is divided into two paths that flow through the LP section, exhausting downward into the condenser.

The turbine stop valves, control valves, reheat stop valves, and intercept valves are controlled by an electro-hydraulic control system.

The turbine is designed to operate at constant inlet steam pressure over the entire load range and is capable of being converted in the future to sliding pressure operation for

economic unit cycling.

Condensate and Feedwater Systems:

The function of the condensate system is to pump condensate from the condenser hot well to the deaerator, through the gland steam condenser and the LP feedwater heaters. The Case-2 condensate and feedwater system differs from Case 1 in that there is a parallel low-pressure Feedwater Heater (PFWH - heated by flue gas) in a parallel feedwater stream with the traditional extraction feedwater heaters. This PFWH is part of the Boiler scope of supply.

The system consists of one main condenser; two 50 percent capacity, motor-driven vertical condensate pumps; one gland steam condenser; two LP heaters, and one deaerator with a storage tank. Condensate is delivered to a common discharge header through two separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line, discharging to the condenser, is provided to maintain minimum flow requirements for the gland steam condenser and the condensate pumps.

The function of the feedwater system is to pump feedwater from the deaerator storage tank to the boiler economizer. Two motor-driven boiler feed pumps are provided to pump feedwater through the three stages of HP feedwater heaters. Pneumatic flow control valves control the recirculation flow. In addition, the suctions of the boiler feed pumps are equipped with startup strainers, which are utilized during initial startup and following major outages or system maintenance.

2.2.4.3. Other Balance of Plant Equipment

The systems for draft, solids handling (coal, limestone, and ash), cooling, electrical, and other BOP systems are described in this section for Case-2.

Draft System:

The flue gas is moved through the boiler, baghouse and other Boiler Island equipment with the draft system. The draft system includes the Gas Recirculation (GR) fans, the fluidizing gas blowers, the induced draft (ID) Fan, and the associated ductwork and expansion joints. This case has no traditional stack as the flue gas generated is supplied to the gas processing system where the CO₂ is purified and liquefied for sequestration or usage. The fans, and blowers are driven with electric motors and controlled to operate the unit in a balanced draft mode with the cyclone inlet maintained at a slightly negative pressure (typically, -0.5 inwg).

Recirculated flue gas from the GR fan is mixed with oxygen from the ASU to provide a combustion oxidant stream, which is split into several flow paths.

Combustion gases exit the furnace and flow through two cyclones, which separate out ash and partially burned fuel particles. These solids are recycled back to the furnace, passing through J-valves, or seal pots, located below the cyclones. The solids leaving the seal pots are then split into two streams. The first stream is uncooled and flows directly to the combustor. The second stream flows through the moving bed heat exchanger where it is cooled before re-entering the furnace at the back wall.

The gas exiting the cyclones passes to the convection pass of the CFB, flowing through

only an economizer section. The gases leaving the convection pass flow through the tubular oxygen preheater and then exit the CFB steam generator to the baghouse for particulate capture. The flue gas leaving the baghouse are further cooled in a PFWH which is a low temperature economizer section and finally in a spray water cooler to about 100 °F. The gases are drawn through the CFB, baghouse, PFWH, and spray cooler with the Induced Draft Fan and then are recirculated to the CFB or discharged to the Gas Processing System.

The following fans and blowers are provided with the scope of supply of the Oxygen-fired CFB steam generator:

- Gas Recirculation fan, which provides recirculated flue gas to be mixed with oxygen from the ASU such that the mixed oxidant stream contains about 70 percent oxygen. This fan is a centrifugal type unit, supplied with electric motor drive, inlet screen, inlet vanes, and silencer (see Table 2.2.10). The electric power required for the electric motor drive is 341 kW.

Table 2.2. 10: Gas Recirculation Fan Specification

<u>Gas Analysis</u>			
Oxygen	(wt percent)	3.31	
Nitrogen	"	3.91	
Water Vapor	"	3.07	
Carbon Dioxide	"	89.53	
Sulfur Dioxide	"	0.18	
Total	"	100.00	
<u>Operating Conditions</u>			<u>Design Spec</u>
Mass Flow Rate	(lbm/hr)	180833	217000
Gas Inlet Temperature	(Deg F)	112.2	
Inlet Pressure	(psia)	14.70	
Outlet Pressure	(psia)	17.41	
Pressure Rise	(in wg)	75.0	97.5

- Induced draft fan, a centrifugal unit supplied with electric motor drive and inlet damper (see Table 2.2.11). The electric power required for the electric motor drive is 511 kW.

Table 2.2. 11: Induced Draft Fan Specification

<u>Gas Analysis</u>			
Oxygen	(wt percent)	3.31	
Nitrogen	"	3.91	
Water Vapor	"	3.07	
Carbon Dioxide	"	89.53	
Sulfur Dioxide	"	0.18	
Total	"	100.00	
<u>Operating Conditions</u>			<u>Design Spec</u>
Mass Flow Rate	(lbm/hr)	637149	764579
Gas Inlet Temperature	(Deg F)	100.0	
Inlet Pressure	(psia)	13.64	
Outlet Pressure	(psia)	14.70	
Pressure Rise	(in wg)	29.5	38.4

- Fluidizing gas blowers, centrifugal units that provide recirculated flue gas for cooling and sealing the seal pots, and for assisting in the conveyance of cyclone bottoms (see Table 2.2.12). The electric power required for the electric motor drive is 209 kW.

Table 2.2. 12 Fluidizing Gas Blower Specification

<u>Gas Analysis</u>			
Oxygen	(wt percent)	3.31	
Nitrogen	"	3.91	
Water Vapor	"	3.07	
Carbon Dioxide	"	89.53	
Sulfur Dioxide	"	0.18	
Total	"	100.00	
<u>Operating Conditions</u>			<u>Design Spec</u>
Mass Flow Rate	(lbm/hr)	37354	44825
Gas Inlet Temperature	(Deg F)	112.2	
Inlet Pressure	(psia)	14.70	
Outlet Pressure	(psia)	23.70	
Pressure Rise	(psia)	9.0	11.7

Ducting and Stack:

There is no stack included in Case-2. The flue gas product leaving the Boiler Island, which is rich in CO₂, is delivered to the Gas Processing System (GPS) where the CO₂ stream is further purified for sequestration or usage. The impurities removed in the GPS, primarily nitrogen and oxygen are vented to atmosphere.

Coal Handling and Preparation:

The function of the coal handling and preparation system is to unload, convey, prepare, and store the coal delivered to the plant. The scope of the system is from the trestle bottom dumper and coal receiving hoppers up to the inlets of the prepared fuel silos.

The bituminous coal is delivered to the site by unit trains of 100-ton rail cars. Each unit train consists of 100, 100-ton rail cars. The unloading is done by a trestle bottom dumper, which unloads the coal to two receiving hoppers. Coal from each hopper is fed directly into a vibratory feeder. The 6" x 0 coal from the feeder is discharged onto a belt conveyor (No. 1). The coal is then transferred to a conveyor (No. 2) that transfers the coal to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron and then to the reclaim pile.

Coal from the reclaim pile is fed by two vibratory feeders, located under the pile, onto a belt conveyor (No. 3) that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3" x 0. The coal then enters a second crusher that reduces the coal size to 1/4" x 0. Conveyor No. 4 then transfers the coal to the transfer tower. In the transfer tower the coal is routed to the tripper that loads the coal into one of the three silos.

Technical Requirements and Design Basis:

- Coal burn rate:
 - Maximum coal burn rate = 163,043 lbm/h = 81.6 tph plus 10 percent margin = 90 tph (based on the 100 percent MCR rating for the plant, plus 10 percent design margin)
 - Average coal burn rate = 139,000 lbm/h = 70 tph (based on MCR rate multiplied by an 85 percent capacity factor)
 - Coal delivered to the plant by unit trains:
 - One and one-half unit trains per week at maximum burn rate
 - One unit train per week at average burn rate
 - Each unit train shall have 10,000 tons (100-ton cars) capacity
 - Unloading rate = 9 cars/hour (maximum)
 - Total unloading time per unit train = 11 hours (minimum)
 - Conveying rate to storage piles = 900 tph (maximum, both conveyors in operation)
 - Reclaim rate = 300 tph
 - Storage piles with liners, run-off collection, and treatment systems:
 - Active storage = 6,600 tons (72 hours at maximum burn rate)
 - Dead storage = 50,000 tons (30 days at average burn rate)

Table 2.2. 13: Coal Receiving Design Summary

Design Parameter	Value
Coal Receiving, tph	90
Active Storage, tons	6,600
Dead Storage, tons	50,000

Limestone Handling and Preparation System:

The function of the balance-of-plant limestone handling system is to receive and store prepared limestone on an as-needed delivery basis. The system consists of a receiving station, unloading system with blowers, and a 1,000-ton silo to accommodate 3 days operation.

Bottom Ash Removal:

Bottom ash, or bed drain material, constitutes approximately two-thirds of the solid waste material discharged by the CFB steam generator. This bottom ash is discharged through a complement of two bed coolers (any one of which must be able to operate at 100 percent load on the design coal). The stripper/coolers cool the bed material to a temperature in the range of 300 °F (design coal) to a maximum of 500 °F (worst fuel) prior to discharge via rotary valves to the bed material conveying system. The steam generator scope terminates at the outlets of the rotary valves.

Fly Ash Removal:

Fly ash comprises approximately one-third of the solid waste discharged from the CFB steam generator. Approximately 8 percent of the total solids (fly ash plus bed material) is separated out in the economizer and oxygen heater hoppers; 25 percent of the total solids is carried in the gases leaving the steam generator en route to the baghouse. Fly ash is removed from the stack gas through a baghouse filter. Particulate conditions are as follows:

Design Specification for Particulate Removal System:

- Total solids to particulate removal system (stream 6, Figure 2.2.1) = 11,998 lbm/h
- Particle size distribution of particulate matter leaving cyclone (streams 5, 6, Figure 2.2.1), see Table 2.2.14.

Table 2.2. 14: Particle Size Distribution

% Wt. Less	Diameter (Micron, μ)
100	192
99	160
90	74
80	50
70	37
60	30
50	24
40	16
30	12
20	8
10	4
1	< 4

- Solids leaving particulate removal system (stream 7, Figure 2.2.1) meet applicable environmental regulations, see Table 2.2.15.

Table 2.2. 15: Fly Ash Removal Design Summary

Design Parameter	Value
Flue Gas Temperature, °F	313
Flue Gas Flow Rate, lbm/h	682,975
Flue Gas Flow Rate, acfm	170,561
Particulate Removal, lbm/h	11,998
Particulate Loading, grains/acf	8207

Ash Handling:

The function of the ash handling system is to provide the equipment required for conveying, preparing, storing, and disposing the bottom ash and fly ash that is produced on a daily basis by the boiler. The scope of the system is from the bag filter hoppers, oxygen heater hopper collectors, and bottom ash hoppers to the truck filling stations.

The fly ash collected in the bag filter and the oxygen heater is conveyed to the fly ash storage silo. A pneumatic transport system using low-pressure air from a blower provides the transport mechanism for the fly ash. Fly ash is discharged through a wet unloader, which conditions the fly ash and conveys it through a telescopic unloading chute into a truck for disposal.

The bottom ash from the boiler is drained from the bed, cooled in a stripper cooler, and discharged to a drag chain type conveyor for transport to the bottom ash silo. Ash from the fluidized bed ash coolers is drained to a complement of screw coolers, which discharge the cooled ash to a drag chain conveyor for transport to a surge bin. The latter is within the boiler scope of supply.

The cooled ash is pneumatically conveyed to the bottom ash silo from the surge bin. The silo is sized for a nominal holdup capacity of 36 hours of full-load operation (1,200 tons capacity). At periodic intervals, a convoy of ash hauling trucks will transit the unloading station underneath the silos and remove a quantity of ash for disposal. Approximately 30 truck loads per day are required to remove the total quantity of ash produced by the plant operating at nominal rated power.

Table 2.2.16: Ash Handling System Design Summary

Design Parameter	Value
Flyash from Baghouse, lbm/h	11,998
Ash from Boiler, lbm/h	65,225
Ash temperature, °F	520

Circulating Water System:

The function of the circulating water system is to supply cooling water to condense the main turbine exhaust steam. The system consists of two 50 percent capacity vertical circulating water pumps, a multi-cell mechanical draft evaporative cooling tower, and carbon steel cement-lined interconnecting piping. The condenser is a single-pass, horizontal type with divided water boxes. There are two separate circulating water circuits in each box. One-half of each condenser can be removed from service for cleaning or plugging tubes. This can be done during normal operation at reduced load.

Condenser Analysis:

The condenser system analysis is detailed in Table 2.2.17.

Table 2.2.17: Condenser Analysis

Item	Value	Units
Pressure	3.0	in. Hga
M stm-in	1,016,890	lbm/h
T stm-in	115.1	°F
P stm-in	1.474	psia
H stm-in	1051.7	Btu/lbm
M drain-in	81,274	lbm/h
H drain-in	89.7	Btu/lbm
H condensate	83.0	Btu/lbm
M condensate	1,098,164	lbm/h
Q condenser	986.3	10 ⁶ Btu/h

Waste Treatment System:

An onsite water treatment facility treats all runoff, cleaning wastes, blowdown, and backwash to within U.S. Environmental Protection Agency (EPA) standards for suspended solids, oil and grease, pH, and miscellaneous metals. All waste treatment equipment is housed in a separate building. The waste treatment system consists of a water collection basin, three raw waste pumps, an acid neutralization system, an oxidation system, flocculation, clarification/thickening, and sludge de-watering. The water collection basin is a synthetic-membrane-lined earthen basin, which collects rainfall runoff, maintenance cleaning wastes, and backwash flows.

The raw waste is pumped to the treatment system at a controlled rate by the raw waste pumps. The neutralization system neutralizes the acidic wastewater with hydrated lime in a two-stage system, consisting of a lime storage silo/lime slurry makeup system with 50-ton lime silo, a 0 – 1,000 lbm/hour dry lime feeder, a 5,000-gallon lime slurry tank, slurry tank mixer, and 25 gpm lime slurry feed pumps.

Miscellaneous systems:

Miscellaneous systems consisting of fuel oil, service air, instrument air, and service water are provided. A 200,000-gallon storage tank provides a supply of No. 2 fuel oil used for startup and for a small auxiliary boiler. Fuel oil is delivered by truck. All truck roadways and unloading stations inside the fence area are provided.

Accessory Electric Plant:

The accessory electric plant consists of all switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, all wire and cable. It also includes the main power transformer, all required foundations, and standby equipment.

Instrumentation and Control:

An integrated plant-wide distributed control and monitoring system (DCS) is provided. The DCS is a redundant microprocessor-based, functionally distributed system. The control room houses an array of multiple video monitor (CRT) and keyboard units. The CRT/keyboard units are the primary interface between the generating process and

operations personnel. The DCS incorporates plant monitoring and control functions for all the major plant equipment. The DCS is designed to provide 99.5 percent availability. The plant equipment and the DCS are designed for automatic response to load changes from minimum load to 100 percent. Startup and shutdown routines are implemented as supervised manual with operator selection of modular automation routines available.

Buildings and Structures:

A soil-bearing load of 5,000 lb/ft² is used for foundation design. Foundations are provided for the support structures, pumps, tanks, and other plant components. The following buildings are included in the design basis:

- Steam turbine building
- Boiler building
- Administration and service building
- Makeup water and pretreatment building
- Pump house and electrical equipment building
- Fuel oil pump house
- Continuous emissions monitoring building
- Coal crusher building
- River water intake structure
- Guard house
- Runoff water pump house
- Industrial waste treatment building

Plant Layout and Plot Plan:

The Case-2 plant is arranged functionally to address the flow of material and utilities through the plant site. A plan view of the boiler, power-generating components, and overall site plan for the entire plant is shown in Appendix II.

2.2.5. Case-2 Overall Plant Performance and CO₂ Emissions

Overall plant performance and emissions for Case-2 are summarized in Table 2.2.18. The Case-1 (Base Case) values are also listed along side for comparison purposes.

Boiler efficiency for Case-2 is calculated to be 94.12 percent (HHV basis) as compared to 89.46 percent for the Base Case. There is a significant improvement in boiler efficiency with oxygen firing as compared to air firing. The improvement is primarily due to the reduced dry gas loss resulting from the oxygen firing. With air firing about 3.3 lbm of nitrogen are carried into the system at ambient temperature with each 1.0 lbm of oxygen. The nitrogen and other combustion products leave the boiler at the exit flue gas temperature. The sensible energy in this exit flue gas stream represents an energy loss to the boiler. For Case-1 the sensible energy leaving the boiler with the nitrogen is nearly 4 percent of the coal heat input. With oxygen firing this loss associated with inert nitrogen is eliminated. Other attributes associated with oxygen firing also contribute to the

improved boiler efficiency but the elimination of nitrogen provides the large majority of the improvement.

The steam cycle thermal efficiency including the boiler feed pump debit is about 41.3 percent as compared to 41.9 percent for Case-1. The slight reduction is due to some low level heat recovery, which is required with the oxygen-fired system. The generator output is 209,907 kW.

The net plant heat rate and thermal efficiency for Case-2 are calculated to be 13,546 Btu/kWh and 25.2 percent respectively (HHV basis).

Auxiliary power for Case-2 is 75,393 kW (about 35.9 percent of generator output). The large auxiliary power increase, as compared to the Base Case, is due primarily to the large power requirement of the cryogenic based ASU and the gas compression requirement in the Gas Processing System of Case-2.

The resulting net plant output for Case-2 is 134,514 kW or about 70 percent of the Base Case output.

Carbon dioxide emissions for Case-2 are 24,618 lbm/hr or about 0.18 lbm/kWh on a normalized basis. This represents about 9 percent of the Case-1 normalized CO₂ emissions and a CO₂ avoided value of 1.81 lbm/kWh.

Table 2.2.18: Case-2 Overall Plant Performance and Emissions

		CFB Air Fired (Case 1)	CFB Cryogenic O ₂ Fired (Case 2)
<u>Auxiliary Power Listing</u>			
	(Units)		
Induced Draft Fan	(kW)	2285	511
Primary Air Fan	(kW)	2427	n/a
Secondary Air Fan	(kW)	1142	n/a
Fluidizing Air Blower	(kW)	920	209
Transport Air Fan	(kW)	n/a	n/a
Gas Recirculation Fan	(kW)	n/a	341
Coal Handling, Preparation, and Feed	(kW)	300	292
Limestone Handling and Feed	(kW)	200	195
Limestone Blower	(kW)	150	146
Ash Handling	(kW)	200	195
Particulate Removal System Auxiliary Power (baghouse)	(kW)	400	151
Boiler Feed Pump	(kW)	3715	3715
Condensate Pump	(kW)	79	79
Circulating Water Pump	(kW)	1400	1877
Cooling Tower Fans	(kW)	1400	1877
Steam Turbine Auxiliaries	(kW)	200	206
Misc. Auxiliary Power (Controls, Lighting, HVAC etc.)	(kW)	719	719
Transformer Loss	(kW)	470	472
Subtotal	(kW)	16007	10983
	(frac. of Gen. Output)	0.077	0.052
<u>Auxiliary Power Summary</u>			
Traditional Power Plant Auxiliary Power	(kW)	16007	10983
Air Separation Unit or Fuel Compressor	(kW)	n/a	37505
OTM System Compressor Auxiliary Power	(kW)	n/a	n/a
CO ₂ Removal System Auxiliary Power	(kW)	n/a	26905
Total Auxiliary Power	(kW)	16007	75393
	(frac. of Gen. Output)	0.077	0.359
<u>Output and Efficiency</u>			
Main Steam Flow	(lbm/hr)	1400555	1400555
Steam Turbine Heat Rate	(Btu/kwhr)	8147	8256
OTM System Expander Generator Output	(kW)	n/a	n/a
Gas Turbine Generator Output		n/a	n/a
Steam Turbine Generator Output	(kW)	209041	209907
Net Plant Output	(kW)	193034	134514
	(frac. of Case-1 Net Output)	1.00	0.70
Boiler Efficiency (HHV) ¹	(fraction)	0.8946	0.9412
Coal Heat Input (HHV)	(10 ⁶ Btu/hr)	1855	1806
Natural Gas Heat Input (HHV) ²	(10 ⁶ Btu/hr)	n/a	16.5
Total Fuel Heat Input (HHV)	(10 ⁶ Btu/hr)	1855	1822
¹ Boiler Heat Output / (Q _{coal} -HHV + Q _{credits})			
² Required for GPS Desiccant Regeneration in Cases 2-7, 13 and ASU in Cases 2-4			
Net Plant Heat Rate (HHV)	(Btu/kwhr)	9611	13546
Net Plant Thermal Efficiency (HHV)	(fraction)	0.3551	0.2520
Normalized Thermal Efficiency (HHV; Relative to Base Case)	(fraction)	1.00	0.71
<u>CO₂ Emissions</u>			
CO ₂ Produced	(lbm/hr)	385427	376995
CO ₂ Captured	(lbm/hr)	0	352377
Fraction of CO ₂ Captured	(fraction)	0.00	0.93
CO ₂ Emitted	(lbm/hr)	385427	24618
Specific CO ₂ Emissions	(lbm/kwhr)	2.00	0.18
Normalized Specific CO ₂ Emissions (Relative to Base Case)	(fraction)	1.00	0.09
Avoided CO ₂ Emissions (as compared to Base Case)	(lbm/kwhr)	0.00	1.81