**CIRCULATING MOVING BED
COMBUSTION PROOF OF CONCEPT – PHASE II**

FINAL REPORT

SUBMITTED BY

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ABSTRACT

Circulating Moving Bed (CMB) combustion technology has its roots in traditional circulating fluidized bed technology and involves a novel method of solid fuel combustion and heat transfer. CMB technology represents a step change in improved performance and cost relative to conventional PC and FBC boilers. The CMB heat exchanger preheats the energy cycle working fluid, steam or air, to the high temperature levels required in systems for advanced power generation. Unique features of the CMB are the reduction of the heat transfer surfaces by about 60% as a result of the enhanced heat transfer rates, flexibility of operation, and about 30% lower cost over existing technology.

The CMB Phase I project ran from July 2001 through March 2003. Its objective was to continue development of the CMB technology with a series of proof of concept tests. The tests were conducted at a scale that provided design data for scale up to a demonstration plant. These objectives were met by conducting a series of experiments in ALSTOM Power's Multi-use Test Facility (MTF). The MTF was modified to operate under CMB conditions of commercial interest. The objective of the tests were to evaluate gas-to-solids heat transfer in the upper furnace, assess agglomeration in the high temperature CMB bubbling bed, and evaluate solids-to-tube heat transfer in the moving bed heat exchanger.

The Phase I program results showed that there are still some significant technical uncertainties that needed to be resolved before the technology can be confidently scaled up for a successful demonstration plant design. Work remained in three primary areas:

- scale up of gas to solid heat transfer
- high temperature finned surface design
- the overall requirements of mechanical and process design.

The CMB Phase II workscope built upon the results of Phase I and specifically addressed the remaining technical uncertainties. It included a scaled MTF heat transfer test to provide the necessary data to scale up gas-to-solids heat transfer. A stress test rig was built and tested to provide validation data for a stress model needed to support high temperature finned surface design. Additional cold flow model tests and MTF tests were conducted to address mechanical and process design issues. This information was then used to design and cost a commercial CMB design concept. Finally, the MBHE was reconfigured into a slice arrangement and tested for an extended duration at a commercial CFB plant.

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1. EXECUTIVE SUMMARY

1.1. Background

Circulating Moving Bed (CMB) combustion technology has its roots in traditional circulating fluidized bed technology and involves a novel method of solid fuel combustion and heat transfer. CMB technology represents a step change in improved performance and cost relative to conventional PC and FBC boilers. The CMB heat exchanger preheats the energy cycle working fluid, steam or air, to the high temperature levels required in systems for advanced power generation. Unique features of the CMB are the reduction of the heat transfer surfaces by about 60% as a result of the enhanced heat transfer rates, flexibility of operation, and about 30% lower cost over existing technology. Specific plant systems have been conceived which would utilize the CMB technology for near-term repowering applications as well as the basis for achieving future advanced plant performance.

The CMB is expected to provide significant advantages in advanced power cycles and enable the achievement of improved efficiency, environmental, and cost goals, as targeted for new energy plants. ALSTOM Power's strategic objectives in developing CMB technology are outlined below:

- Enable high temperature Rankine cycles
 - Both greenfield plants and repowering
- Develop a low-cost CFB alternative for conventional steam cycles
 - Target of 30% lower material cost than current CFB designs
- Develop near-term spin-offs
 - Replace CFB FBHEs with a moving bed heat exchanger
- Enable advanced power cycles and CO₂ capture
 - O₂-fired CFB
 - Chemical looping

The CMB Phase I project ran from July 2001 through March 2003. Its objective was to continue development of the CMB technology with a series of proof of concept tests. The tests were conducted at a scale that provided design data for scale up to a demonstration plant. These objectives were met by conducting a series of experiments in ALSTOM Power's Multi-use Test Facility (MTF). The MTF was modified to operate under CMB conditions of commercial interest. Tests were conducted to evaluate gas-to-solids heat transfer in the upper furnace, assess agglomeration in the high temperature CMB bubbling bed, and evaluate solids-to-tube heat transfer in the moving bed heat exchanger.

The Phase I program results showed that there were still some significant technical uncertainties that needed to be resolved before the technology could be confidently scaled up for a successful demonstration plant design. Work remained in three primary areas:

- scale up of gas-to-solid heat transfer
- high temperature finned surface design
- the overall areas of mechanical and process design.

The CMB Phase II workscope built upon the results of Phase I and specifically addressed the remaining technical uncertainties. It included a scaled MTF heat transfer test to provide the necessary data to scale up gas-to-solids heat transfer. A stress test rig was built and tested to provide validation data for a stress model needed to support high temperature

finned surface design. Additional cold flow model tests and MTF tests were conducted to address mechanical and process design issues. This information was then used to design and cost a commercial CMB design concept. Finally, the MBHE was reconfigured into a slice arrangement and tested for an extended duration at a commercial CFB plant.

The U.S. Department of Energy (DOE) and ALSTOM Power jointly funded the CMB Proof of Concept project to evaluate and develop Circulating Moving BedTM combustion system technology. The objective of this project was to demonstrate the “proof of concept” for CMBTM combustion system technology, enabling it to be used as a key building block of new energy plants. The total Phase II project budget was \$4,998,000 with the DOE’s National Energy Technology Laboratory (NETL) providing 60% of the funding under Cooperative Agreement No. DE-FC26-01NT41223.

Circulating Moving Bed technology (illustrated in Figure 1-1) is a new method for solid fuel combustion and heat transfer, which has roots in the traditional circulating fluidized bed (CFB) technology. The CMB technology completely separates the combustion process from the heat transfer surfaces. Fuel, including coal and a wide range of alternate fuels, is burned in the bubbling bed region of the combustor to produce combustion temperatures approaching 2000°F. The heat from the products of combustion (upward flow) is exchanged (recuperated) with a flow of high-density solid particles falling downward through the upper region of the combustor. The solids have recuperated the heat of combustion upon reaching the bottom of the combustor and are then fluidized and transferred to a moving bed heat exchanger (MBHE) through standpipes.

Heat from moving particles is captured in a series of tubular moving bed heat exchangers that heat a working fluid, such as steam or compressed air, to the required process temperatures. From the bottom of the moving bed heat exchanger, the cooled solids are transported back to the top of the combustor to restart the lower recycle loop. Sintered bauxite particles can be used as the heat transfer solids media because they have a high density, have good bulk flow properties, are chemically inert, and are readily available. Figure 1-1 also shows an enlarged photo of a group of bauxite particles.

The flyash entrained in the flue gas flows upward and is captured by a cyclone. A portion of the solids is returned to the bubbling bed with the balance going to a cooler and disposal. This flyash is used to reduce carbon loss and to give additional control of the resultant bed temperature. The SO₂ emissions will be controlled primarily by a backend cleanup system such as ALSTOM Power’s NIDTM system. Limestone is calcined in the combustor for use in the backend desulfurization system. Additional sulfur capture is achieved in the combustor.

Figure 1-1: CMB™ Combustion System Schematic

The high CMB combustion temperature makes it possible to achieve working fluid temperatures as high as 1750°F. In addition to enabling high-temperature, high-efficiency power plant cycles, the combustion temperature offers better carbon burnout, virtual elimination of N₂O emission, and low carbon monoxide emissions.

1.2. Project Objectives and Workscope

The objective of the CMB Proof of Concept project was to identify the technical, design, and performance challenges that need to be met to make a commercial CMB system. The project objectives were initially achieved in Phase I through a series of experiments that were conducted in ALSTOM Power's Multi-use Test Facility. A number of technical challenges were identified that were the basis for continuing this work in Phase II.

The specific Phase II project objectives were set by ALSTOM Power, in concert with the U.S. DOE, and are summarized below:

- Improve the understanding of gas-to-solids heat transfer and then apply it to scale-up predictions for commercial sized units
- Develop recommendations for the tubing in the hottest parts of the MBHE
- Evaluate alternate mechanical designs and develop designs for critical components to be included in the commercial conceptual design

- Evaluate alternative techniques to enhance process performance
- Conduct MTF pilot plant tests to improve overall CMB performance with several fuels and to gather scaled heat transfer data for CFD validation
- Evaluate gas-to-solids heat transfer in a scaled MTF test and gather heat transfer validation data for the CFD process model. Apply the model to make scaleup predictions of CMB performance in a commercial scale combustor.
- Evaluate solids-to-tube heat transfer and solids distribution in the moving bed heat exchanger
- Evaluate factors to control ash agglomeration in the bubbling bed
- Evaluate CMB process performance, including carbon burnout, sulfur capture in the furnace and in the backend FDA system, and other gaseous emissions, including CO, NO_x, and N₂O
- Develop a commercial CMB boiler concept and cost evaluation
- Conduct extended duration MBHE field tests to evaluate long-term heat transfer performance and MBHE operational reliability

The Phase II project changed significantly during the contract period as new information became available, with two contract scope modifications. A key Phase II objective was to improve the understanding of gas-to-solids heat transfer and then apply it to scale-up predictions for commercial sized units. This objective was originally to have been met through a combination of CFD studies and heat transfer tests in a new large-scale test facility. However, the improved CFD model showed that the predicted solids flow distribution in a commercial CMB combustor was more uniform than originally believed and that the model was able to accurately model the heat transfer performance in the MTF. Confidence was also gained in the ability to use scaling criteria to scale the results from the MTF to commercial sized CMB combustors. The Phase II workscope was therefore modified to conduct gas-to-solid heat transfer tests in a scaled MTF test rather than in a separate large-scale heat transfer test facility.

The CMB was originally proposed as a cost effective design for high temperature Rankine cycles. The concept was also adopted for near term applications as a low cost CFB alternative with conventional steam conditions. ALSTOM Power established a goal of developing a near term CMB that had boiler island material costs that were 30% less expensive than a conventional CFB. The Phase II study showed that the CMB as currently designed would not meet the 30% cost reduction target for the near term CFB replacement. Although CMB's long range potential was for high temperature Rankine cycles, its near-term use as a lower cost CFB was considered to be a necessary stepping stone in its development path. ALSTOM Power plans to continue development of the MBHE technology as a lower cost replacement for a CFB MBHE. The MBHE eliminates the use of high pressure fluidizing air and results in significant auxiliary power savings in addition to capital cost reductions. It is also an enabling technology for both chemical looping and O₂-fired CFBs. ALSTOM Power therefore requested and received the DOE's agreement to redirect the CMB program to continue development of the MBHE only. The focus of this work was directed towards a long-term field test of the MBHE at a commercial CFB plant.

The Phase II project objectives were achieved through a series of experiments that were conducted at a scale that provided the design data for scale up to a demonstration plant. The proof of concept tests were conducted in ALSTOM Power's 3Mw_{th} Multi-use Test Facility (MTF). The MTF was modified to operate with all CMB subsystems fully integrated and operating at commercially significant temperatures. The CMB Proof of Concept workscope also included process evaluation and component development tasks in support of pilot plant testing. The MTF MBHE was also modified into a wider arrangement and installed at the Grant Town CFB for extended duration field tests.

Overall Phase II project results confirmed high heat transfer rates in the combustor and MBHE, along with a methodology for predicting scale up performance. Combustion performance was very good and bed ash agglomeration was controllable. Emissions were close to state of the art levels and can be further optimized to meet the DOE's Advanced Combustion Systems targets. Mechanical systems were developed for the MBHE and the MBHE was then tested at a commercial CFB site for 121 days. The following section summarizes the progress that was made against the specific Phase II project objectives for the CMB Proof of Concept project.

1.3. Gas to Solid Heat Transfer

The objective of this task was to improve the understanding of gas-to-solids heat transfer and then apply it to scale-up predictions for commercial sized units. The Phase I modelling work identified issues with grid independence, number of particle trajectories, and scale up methodology. All three issues were satisfactorily addressed in Phase II.

A grid independence study compared two different CFD models with the total number of grids ranging from 310,00 up to 2,800,000 cells. Both CFD models showed essentially the same particle concentration profiles at all grid levels, indicating that the models were 'grid independent'. This was a significant finding; since the larger models took weeks to get results even when using a computer cluster of ten state-of-the art CPUs.

A similar study investigated the number of particle tracks that were required for the CFD simulations. The particle tracks determine the position in the flow field that the particles exchange momentum and energy with the gas flow. The same two CFD models were run with the number of particle tracks varying from 240 to 240,000 tracks. The study again showed that results (see Figure 1-2) were essentially the same regardless of the number of particle tracks. In all cases, the model results showed that the particle distribution was fairly well dispersed.

Figure 1-2: Particle Track Comparison for 25Mw and MTF Models

A scale up methodology was developed based upon a simplified set of scaling relationships that were proposed by Dr. Leon Glicksman. The simplified technique required matching the Froude number, the solids to gas ratio, the ratio of the gas superficial velocity to particle terminal velocity, the ratio of linear bed dimensions, the dimensionless solid flow rate, the particle sphericity, and the dimensionless particle size distribution. The scaling relationship indicated that the MTF could be operated under scaled conditions to model a small 25Mw commercial CMB combustor. A CFD model was then made of a 25Mw CMB combustor and of the MTF operating in scaled mode. The model results were very close and validated the use of a scaled MTF test for gathering CFD validation data.

Cold flow validation tests were also conducted on the MSE Meili Labasys 100 measuring system to check its feasibility for solids concentration and velocity measurements. Tests were conducted over a range of bauxite particle sizes, solid flow rates, and gas superficial velocities. The results confirmed that the MSE Meili system was a feasible tool for characterizing gas/solids flow during the CMB hot gas scaling test campaign.

A further verification of the CFD modelling technique was made by operating the MTF with an intentionally disrupted solids flow field, then trying to simulate the results in the model. A deflector plate was installed under the MTF solids injector to artificially induce a poor solid distribution. The CFD correctly modelled the overall trends observed in the MTF for the deflected solids case.

1.4. Extended Tube Evaluation

The objective of this task was to develop recommendations for the tubing in the hottest parts of the MBHE. A stress model was developed to accurately determine the design limits of spiral-finned tubing, specifically the limitation imposed on the fin tip to fin base temperature difference. The analyses demonstrated that the stresses in the tube and the fin can be predicted, including cyclic response, but the lack of information about the strength of the fin-

to-tube weld hampered quantitative life estimates. A stress test rig was therefore built to experimentally determine the strength of the fin-to-tube weld under realistic loading. The upper two photos in Figure 1-3 show a close-up view of the test section and the test section just before it was raised into the oven.

The finned tube thermal cycling test rig automatically moved a test sample with either air or water-cooling into and out of a heated furnace on an automatic basis. The first sample experienced over 2000 cycles without any indication that the thermal cycling impaired heat transfer. Exposure testing on the second sample ended after 3600 cycles with a fin tip to tube temperature differential of 360°F, again with no signs of fin separation from the tube.

The finite element analysis stress model was made of the stainless steel fin on carbon steel tube shown in the bottom left photo of Figure 1-3. The model was validated with test results from the thermal cycling test rig. The model temperature predictions shown in the lower right photo closely matched the measured temperatures for both the fin tip and the tube. Based on the validated model, tip-to-base temperature differences of up to 400°F are predicted to be permissible for cyclic operation where long-term creep failure is not a concern.

Figure 1-3: Fin Integrity Assessment

1.5. Mechanical Evaluation

The objective of this task was to evaluate alternate mechanical designs and develop designs for critical components to be included in the commercial conceptual design. Mechanical designs were developed for various aspects of the MBHE design, including ash control valves, a solids distributor, a distribution orifice plate, and a vent system. All of these designs were tested in cold flow models, followed by hot testing in both the MTF pilot plant and in the Grant Town CFB MBHE.

The MBHE ash control valve was developed to control the solids flow rate and also designed to pass oversized material through the valve. The pneumatic ash flow control valve has no moving parts and uses low-pressure control air from the PA fan to activate the valve and to transport the solids to the combustor. The ash flow control valves can transport up to 400,000 lb/hr/ft² of ash with an ash to air ratio of 2000 or greater. The clearances of the valve were designed to pass a rubble piece of the same size as the valve inlet. The two left photos in Figure 1-4 show the ash control valve and some of the rubble that passed through it during the cold flow tests. The right figure shows the MTF MBHE that used these ash control valves to control solids flow rate.

FLOW CONTROL SHOE

RUBBLE FLOW THRU SHOE

Figure 1-4: Solids Flow Control in the MBHE

An active grid flow distributor was developed to provide uniform solids distribution control and the ability to pass rubble without plugging. The MTF active grid consists of 16 pyramidal hoppers in a 4x4 array with a flow control shoe under each hopper. Solids flow is controlled by actuating the flow control shoe air for 10-15 seconds and then stopping the air flow. The air flow is altered among the various shoes of the grid in a prescribed pattern so that each shoe in succession is actuated. The total flow is controlled by the dwell time between activation.

A MBHE distribution orifice plate was developed to ensure uniform solids flow throughout the MBHE. The multiple orifice plate design is simple and requires a 35-40° slope with less elevation requirement, resulting in a 75% height reduction as compared to using mass flow hoppers with 70° angles at the bottom of the MBHE.

A vent system was also developed to direct transport air from the active solids distribution grid to the vent at the top of the heat exchanger.

1.6. Process Evaluation

The objective of this task was to evaluate alternative techniques to enhance process performance. Techniques to enhance process performance were evaluated for agglomeration mitigation, CO emissions control, alternate methods for sulfur capture, along with an evaluation of CMB fuel flexibility.

A review was completed of the causes of rubble accumulation in the MBHE discharge during the Phase I testing, plus an analysis of agglomeration experience in commercial BFB boilers. The analysis showed that a properly designed MBHE could be operated with reliable solids flow and with an absence of solids agglomeration. Most of the early MBHE operation problems were addressed through proper refractory curing and vent design.

The CMB pilot plant tests have shown that a layer of ash can build up on the sintered bauxite used as the solids circulating material. With time, the ash accumulation will impact the size and aerodynamic properties of the bauxite and the particles will need to be gradually withdrawn from the system and replaced. Since the bauxite is quite expensive, an effort was made to identify alternative candidates for the CMB heat transfer media. Eleven alternative candidates were identified that were all commonly available worldwide and relatively inexpensive. A laboratory-screening test was developed to study the thermal stability of the candidate materials in a simulated CMB environment with respect to melting, softening, and tendency to agglomerate. The samples were exposed in an oven at temperatures between 2000°F – 2200°F in the presence of coal ash, lime, and a mix of coal ash components. The top two photos in Figure 1-5 show one sample mixture as it's being placed into the furnace. The bottom two figures show the bauxite and dolomite samples after being heated at 2200°F. Although each of the samples showed varying degrees of ash wetting, none of them exhibited behaviour that eliminated them from consideration as alternate heat transfer media. Further evaluation would be needed to identify a preferred alternate material for the heat transfer media.

Sample boat filled with dolomite and ash

Sample boat being placed into 2200°F furnace

Bauxite at 2200°F - 7X magnification

Dolomite at 2200°F - 7X magnification

Figure 1-5: Laboratory Thermal Screening Tests

The original CMB concept was based upon concentrating all of the heat transfer surface in the MBHE, thus eliminating the need for any combustor or backpass surface. This approach would result in the combustion gas being cooled to about 700°F leaving the combustor. MTF pilot plant tests showed that CO emissions increased as the combustor outlet temperature decreased below 1300°F. Chemkin and Fluent models showed that char particles continued to react and produce CO at temperatures as low as 800°F. However, the models also predicted that CO conversion was essentially quenched as temperatures dropped below 1200°F.

Various options were considered for mitigation of CO emissions. CMB designs were developed that significantly reduced the combustor height, while correspondingly increasing the combustor gas outlet temperature. This approach would reduce CO emissions, but the cost savings from the reduced combustor height and smaller MBHE were overwhelmed by the added cost of a high temperature cyclone, additional refractory, and a backpass. In-furnace and backend additives were also considered. These methods would reduce CO emissions somewhat, but the additive costs were quite high. A backend CO oxidation catalyst was also considered. Only one catalyst vendor was willing to propose a system that would survive in this environment, due to concerns of catalyst erosion and deactivation from sulfur poisoning. The catalyst would be able to make significant reductions in the CO emissions, but the equipment cost was quite high.

The CMB combustor meets SO₂ emission requirements through a combination of in-furnace sulfur capture and sulfur capture in the backend NID system. The NID system captures the majority of the SO₂ because the high combustion temperatures limit in-furnace capture. Depending on the coal sulfur content, the limited in-furnace sulfur capture could also have a

significant negative impact on the CMB boiler efficiency and overall economics. Sorbent consumption was minimized during the MTF pilot plant tests by operating the NID system with higher humidity levels than normal commercial practice. A decision was made for the CMB commercial evaluation to limit NID operation to commercially demonstrated humidity levels despite the positive MTF results. This approach would require significantly higher sorbent feed rates. Alternative sulfur capture methods were therefore assessed for the commercial plant evaluation.

A wide range of alternate approaches to sulfur capture were considered, including both commercially ready technologies and some under development. They included wet and dry scrubbers, steam reactivation to enhance in-furnace sulfur capture, and a cyclone based alternative to the NIDTM system. Capital costs were developed for each system. The lowest cost system was the cyclone NID alternative, followed by steam reactivation. The wet scrubber was clearly the most expensive design.

An assessment of fuel flexibility showed that the CMB is capable of handling a wide range of fuels similar to conventional CFBs, although some limitations may result due to the higher temperatures in the lower furnace. The best fuels for CMB are high reactivity, low sulfur fuels, including many subbituminous coals, high volatile bituminous coals, lignites, brown coals, and biomass. These attributes aid carbon burnout, CO emissions, low sulfur capture equipment requirements, and are attractive for ultra high efficiency and potential CO₂ control schemes. Low reactivity, high sulfur content, very high ash, and very low melting ash fuels may require design compromises that increase costs.

1.7. MTF Campaigns

The objective of this task was to conduct MTF pilot plant tests to improve overall CMB performance with several fuels and to gather scaled heat transfer data for CFD validation. Four MTF test campaigns were conducted during Phase II. The first test campaign used Pittsburgh #8 coal and a range of different limestone sizes. The main objectives were to evaluate MBHE performance with the improved solids distributor, evaluate process improvements to combustion and environmental performance, and to evaluate gas-to-solids heat transfer with poor solids distribution. The second test campaign continued the objectives of the first campaign, but fired a West Virginia bituminous coal and a petroleum coke. The objective of the third campaign was to gather gas-to-solids heat transfer in a scaled MTF test while also gathering validation data to calibrate a CFD model of the CMB process. The objective of the final test campaign was to collect moving bed heat transfer data when the MTF was operating in CFB mode rather than CMB mode.

Gas-to-solids heat transfer and scaleup predictions

Six scaled heat transfer tests were conducted with solids-to-gas ratios ranging from 0.5 to 4. Detailed temperature, solids concentration, and solids velocity profiles were collected for each test. The results showed that the Nusselt Number was generally lower than predicted for single particles due to particle-to-particle interactions. Particles in close proximity had a tendency to follow in each other's wake due to drag reduction phenomena. Particles following a lead particle were exposed to gas cooled by the lead particle, and therefore did not have the full gas-particle temperature difference available to the lead particle. A multi-particle model developed from earlier warm air tests gave a good fit to the whole range of MTF data.

The CFD model did a reasonable job modelling the MTF gas and solids flows when the heat transfer coefficient was adjusted to about 0.1 - 0.2 times the default Fluent single particle heat transfer coefficient. The calibrated CFD model was then used to model a commercial scale CMB combustor. The commercial model results illustrated in Figure 1-6 show that the solids are flowing downward through the combustor in a gently expanding cone pattern. The gas has a higher velocity in the center of the combustor but has to travel through the falling solids to exit at the top of three cyclone outlet ducts. This flow pattern results in very good heat transfer even with the use of a low heat transfer coefficient. The results are greatly improved as compared to the previous results from the original Phase I commercial CFD models, which showed varying degrees of gas and solids flow separation.

Figure 1-6: CFD Commercial CMB Plant Model – Solids Concentration and Velocity Profiles

Solids-to-tube heat transfer

Cold flow model heat transfer tests were conducted in a MBHE slice model to determine the effect of tube bundle depth, particle composition, transverse tube spacing, and ash velocity on MBHE heat transfer performance. There was little heat transfer impact with bed depth at solid velocities below 100 ft/hr and only a small decrease in heat transfer at the bottom of the tube bundle at velocities between 100-300 ft/hr.

There was little difference in heat transfer performance between using bauxite particles and CFB ash, despite the wide differences in particle size distribution and chemical composition. The tube spacing tests showed that there was little transverse mixing of the ash as it passed through the MBHE. The tube transverse spacing should be arranged such that ash flowing downward will touch a fin or tube surface – otherwise, the ash will just pass straight through the bundle without touching any heat transfer surface. MBHE heat transfer was measured at ash velocities from 15 to 300 ft/sec and showed a slight increase with velocity.

The MTF MBHE test results showed that the solids-to-tube heat transfer of the tightened pitch tube bundles exceeded the performance of the previous inline tube bundles used in Phase I by up to 60%. It is possible that the tube bundle performance could have been even higher than observed since the tube bundle showed signs of significant solids flow maldistribution even when the best heat transfer rates were obtained. One promising result was that the tube bundle fins were largely rock-free, with only a few instances of rocks lodged in the fins. The staggered tube bundle performance could now be predicted for both hot and cold conditions as well as for variations in tube bundle configuration.

Ash agglomeration in the bubbling bed

Results indicate that bed ash agglomeration is controllable over the range of CMB commercial temperatures. Two bituminous coals and a petroleum coke were fired for 268 hours of operation at 2000°F or above. No agglomerates were formed in the bed and the bed remained well fluidized throughout the operation. These results indicate that the CMB fluidized bed is quite robust and can tolerate temporary temperature excursions or temperature maldistributions.

The bauxite particles did develop an ash coating on its surface that slowly accumulated during operation. The ash coating was primarily a mixture of calcium, aluminum, silica, and iron oxides in varying proportions. The coating rate was retarded by the mechanical abrasion and attrition as the particles passed through the transport system. Although the growth rate was slowing, test duration was not long enough to conclude if the particle ash coating would eventually reach a steady state level, where any further growth was offset by attrition. There was some evidence from prior work that some of the coating would spall off when the coating became thicker.

If not controlled, the ash buildup on the particle surface could have a significant impact on gas-to-solids heat transfer and bed pressure drop. The larger particle size will increase the combustor gas outlet temperature due to the reduced particle surface area at a given solids circulation rate. Furthermore, bed inventory will need to be controlled to maintain a constant bed pressure drop as the ash coating increases. The bauxite particle size will thus need to be managed to some maximum size level to maintain proper boiler performance. Particle size can be controlled through several techniques, including attrition, modifying local bed conditions to reduce the coating growth rate, and solids replenishment. An alternative approach is to replace the bauxite with one of the less expensive heat transfer particles evaluated in this study. These particles will require additional testing in a hot MTF test to confirm that they are suitable alternatives for the heat transfer media.

CMB process performance, including carbon burnout and emissions

The MTF results demonstrated good combustion performance and gaseous emissions that could be controlled to current state-of-the-art CFB levels and could be further optimized to meet the DOE Advanced Combustion System targets. Tests were conducted with two different coals, four different sorbents with six different size distributions, and with ammonia and steam injection.

The combustion performance was as good as CFB performance with the same coals. The combustion efficiency for the two bituminous coals ranged from 96.1 to 99.6% over both the Phase I and II MTF test campaigns. The combustion efficiency of the petroleum coke ranged from 97.1 to 97.6%. One test condition was run at reduced load without any ash recycle from the cyclone. The combustion performance during this test was slightly better

than the performance at full load with ash recycle. Further work needs to be done to assess this condition as operation without a cyclone represents a considerable cost savings for a commercial CMB boiler.

Figure 1-7 and Figure 1-8 summarize the emission results that were obtained while burning the West Virginia bituminous coal and a petroleum coke in the MTF. The tests included a wide range of parametric variations. These figures identify those periods where ammonia injection was used to further control NO_x emissions.

Figure 1-7: MTF Emission Results with West Virginia Bituminous Coal

Figure 1-8: MTF Emission Results with Petroleum Coke

NO_x emissions were reduced to below 100 ppm without the addition of ammonia by low excess air and deeper staging. NO_x levels with the Pittsburgh #8 coal were as low as 68 ppm or 0.08 lb NO_x/MBtu without ammonia injection and averaged about 100 ppm. NO_x emissions were somewhat higher with both the West Virginia bituminous coal and pet coke, averaging about 150 ppm for each fuel. NO_x reductions of up to 40% were achieved when aqueous ammonia was injected. The reduction level was affected by the injection location and the local temperature and oxidizing conditions. No ammonia slip was detected during this testing. No further reduction in NO_x was observed with the addition of steam.

N₂O emissions were not measured during Phase II testing, but they were extremely low during all Phase I tests because of the high temperatures in the lower furnace. Emissions were always less than 5 ppm.

CO emissions were generally higher than observed during CFB operation. CO is apparently generated in the upper furnace where the gas temperatures are low. Emissions tended to increase as the temperature in the upper furnace decreased. CO emissions typically ranged from 180 to 400 ppm during these tests. CO emissions tended to increase slightly with the addition of ammonia, while there was little apparent impact from the steam injection.

The sulfur capture results from the Phase I program showed that sulfur capture close to 100% could be achieved at a Ca/S mole ratio of 2 with fine limestones and high NID humidity levels. However, the sulfur capture performance in the NID system was limited by how much calcium oxide was carried over to the backend equipment. An objective of the Phase II test campaign was therefore to try to improve the overall sulfur capture performance by increasing the carryover of calcium to the NID system. This was first

accomplished by reducing the size of the initial sorbent from a D_{50} of 50 microns for the EcoCall 50/50 limestone down to a D_{50} of 9 microns for the Marblewhite 325 limestone. Further attempts to increase the sorbent carryover were also made by intentionally disrupting the cyclone to decrease its collection efficiency.

All of the tests were conducted with a Ca/S mole ratio of 2. The in-furnace sulfur capture with the Pittsburgh #8 coal was roughly 40% to 50% for most of the tests. The overall sulfur capture was 60% to 70% for the early tests and then increased to over 90% when the NID relative humidity was increased to 80%. The sulfur capture with the finer Marblewhite limestone reached 98% with a relative humidity of 80%. The in-furnace sulfur capture with the West Virginia coal was 30% to 40% and averaged 40% with the petroleum coke. The total sulfur capture was about 70% with the West Virginia coal, largely because it used the coarser limestone and calcium oxide wasn't reaching the NID system. The total sulfur capture with the petroleum coke averaged 99% with a NID system relative humidity of 80%. These tests were conducted with the finer Marblewhite limestone, which allowed more of the sorbent to pass through the cyclone to the NID system.

1.8. Commercial Plant concept

The objective of this task was to develop a commercial CMB boiler concept and cost evaluation. Two CMB boiler designs and cost comparison studies were completed. The first was a preliminary 300Mw_e CMB design, followed by a more detailed 2x150Mw_e CMB design. The preliminary 300Mw_e study included the evaluation of a wide range of CMB arrangements, with the intent of identifying a configuration that would then be used for a more detailed cost study. Costs were then developed for the 300Mw_e CMB and for a 300Mw_e CFB. The study showed that the CMB boiler material costs were between 19-31% lower than the CFB design. The CMB auxiliary power was about 0.6% less than for the CFB and the O&M costs were about the same for both designs.

The more detailed 2x150Mw_e CMB design study was for steam conditions of 2130psig/1009°F/1009°F. Five CMB designs were considered with combustor gas outlet temperatures varying from 630°F to 1300°F. As the combustor gas outlet temperature increased, the duty in the MBHE decreased and the size of the backpass increased. At the same time, the height of both the MBHE and combustor decreased. Seven different sulfur capture technologies were also evaluated for each CMB configuration. Costs were developed for each CMB design and then compared against a 2x150Mw_e CFB design, both without and with a backend NID system for sulfur control. Figure 1-9 shows an isometric view of four of the CMB designs and illustrates how the equipment size changed as the combustor gas outlet temperature was increased.

Figure 1-9: 150Mw_e CMB Isometric View

The comparison study showed that the CMB combustor had significantly lower weights than the CFB design, with the largest weight savings in pressure parts and also in refractory for the low temperature CMB cases. The CMB boiler design with a 630°F combustor outlet temperature weighed 32.8% less than the CFB design. The weight savings in the CMB with the 1300°F combustor outlet temperature was reduced to 21.1% because of the much larger requirement for refractory.

The cost study showed that the CMB with the 630°F combustor outlet temperature had boiler island material costs that were 22.7% less than the CFB design. The CMB material savings for the 1300°F gas outlet temperature were only 15.1% less than the CFB design, largely because of the large increase in refractory weight. None of the CMB designs evaluated in this study met the 30% materials costs savings target against the CFB design.

Total boiler island costs were developed for both the 2x150Mw_e CMB and CFB designs, including all materials, services, and erection costs for the boiler island and gas cleanup systems. The original CMB design (with NID system operating at high humidity levels) cost \$17.4 million less than the CFB design with NID system. The lowest cost CMB design used a CycNID system for sulfur capture and cost \$29.0 million less than the CFB design with NID system.

Total evaluated or life cycle costs were also developed for the CMB and CFB designs for a moderate and a high sulfur fuel. The evaluated costs included boiler and gas cleanup system capital costs and life cycle costs for auxiliary power and all operating costs. The evaluated costs were expressed as a delta cost compared to the base CFB with NID design. They did not include any CO mitigation costs, which could add another \$3-5 million cost to the CMB cases with low combustor gas outlet temperatures.

The results showed that several CMB configurations looked promising compared to the CFB with NID design. The CMB with steam reactivation and CycNID showed a \$21 million evaluated cost savings compared to the CFB with NID design, while the original CMB

design with NID system at high humidity showed savings of \$5.3 to \$10.5 million with the high and moderate sulfur fuels, respectively.

The CMB cost study showed that the CMB as currently designed would not meet the 30% cost reduction target (for the near term design) even if the high CO emissions could be resolved with no additional cost. There were some scenarios where the total evaluated costs of the CMB design were favourable compared with the CFB with NID design. However, these scenarios all relied upon environmental control technologies that were also still under development. Further development to address both CMB and advanced environmental control technical issues would be costly and outside the scope of this CMB program. Although CMB's long range potential was for high temperature Rankine cycles, its near-term use as a lower cost CFB for conventional steam cycles was considered to be a necessary stepping stone in its development path. The total evaluated cost of a near term CMB design was not significantly lower than ALSTOM Power's conventional CFB designs. Given the technical risks associated with CMB, these potential savings did not make a compelling argument for continued development of CMB as a low cost CFB replacement.

1.9. MBHE Commercialization

The objective of this task was to conduct extended duration MBHE field tests to evaluate long-term heat transfer performance and MBHE operational reliability. ALSTOM Power plans to continue development of the MBHE technology as a lower cost replacement for a CFB FBHE. The MBHE eliminates the use of high pressure fluidizing air and results in significant auxiliary power savings in addition to capital cost reductions. From the DOE's perspective, the MBHE is an enabling technology for both chemical looping and O₂-fired CFBs. Continued development of this technology will reduce the risk of future demonstrations of both chemical looping and O₂-fired CFBs. ALSTOM Power therefore requested and received the DOE's agreement to redirect the CMB program to continue development of the MBHE only. The focus of this work was directed towards a long-term test of the MBHE at a commercial CFB plant.

Five different CFB plants were visited in an attempt to find a suitable host site to test the slipstream MBHE. The first four CFB sites had various arrangement issues that interfered with the proposed MBHE installation. The final CFB site was a 2x40Mw_e CFB in Grant Town, West Virginia that was owned by American Bituminous Power Partners. The plant had sufficient room for the MBHE installation, plus the CFB was small enough that the MBHE could potentially provide some process benefits to the plant.

The MTF MBHE was reconfigured from a 32" x 32" plan area to a 16" by 84" slice plan area to better evaluate the effect of solids distribution on MBHE heat transfer performance. This modification increased the width of the MBHE to almost half of the width of a commercial 250Mw MBHE.

The MBHE installation at the Grant Town CFB was completed in August 2008. Additional modifications were required, but could not be completed until the September 2009 plant outage. Once the modifications were completed, the MBHE performance was measured at commercial CFB operating conditions over 121 days of cumulative MBHE operation. During the test period, there were several periods where the MBHE was out of operation due either to power plant outages or to MBHE problems. The MBHE operation was limited to less than 50% of the solids design flow rate throughout most of the testing due to various mechanical

issues. One test was conducted on just the left side of the MBHE with solids flow up to 18,000 lb/hr or 90% of the design solids flow rate.

Operation was finally suspended in May 2010 following a failure of the cyclone sealpot return chute. Grant Town Plant Operations decided to isolate the MBHE to take the load of the MBHE ash supply line off of the loop seal by unbolting the downstream side of the MBHE inlet isolation valve. Subsequent inspection showed that the MBHE box inlet was free and able to expand. It is highly unlikely that the MBHE contributed to the cyclone sealpot return chute failure as there was no evidence that the MBHE ash supply line was not able to move thermally with the boiler and loop seal. However, Plant Operations still decided to leave the MBHE ash supply line disconnected from the loop seal until they had completed an investigation of the failure. At the time of this report, ALSTOM Power was still waiting for a decision from Grant Town on whether MBHE operation could be resumed.

The MBHE test results indicated that the heat transfer performance did not deteriorate with time. As shown in Figure 1-10, the overall heat transfer performance was slightly less than predicted at higher solids flow rates. This was most probably due to ash bypassing the finned region of the tube bundles in the return bend region and because of agglomeration build-up in the tube bundle during the initial start-up. Analysis of the agglomerates indicated that they were caused by excessive moisture from the CFB flue gas reacting with the calcium oxide in the ash.

Figure 1-10: Grant Town MBHE Measured Heat Flow vs Ash Flow

The results also demonstrated that maldistribution of the solids through the tube bundle had a significant influence on performance. The results demonstrated that the solids flow distribution baffle could be changed to improve the distribution at fixed flows but was susceptible to plugging. Pneumatic ash control valves (ACV) at the bottom of the MBHE were used to control the ash flow for an extended period of time.

Heat transfer prediction methods for a moving bed heat exchanger were verified to be consistent with operating time and ash flow. Also, the depth effect of the tube bundle did not appear to be significant.

The ash exit temperature profile at the bottom of the MBHE had more variation than expected. A passive ash-distributing grid at the bottom of the MBHE was prone to pluggage from agglomerates formed in the ash inlet pipe and top of the tube bundle. Orifices were installed to reduce the hole size and resulted in an improved distribution on the right side. The distribution on the right side showed that the orifices did change the temperature distribution, and further adjustments may improve the temperature distribution.

The total transport pressure drop necessary to return MBHE solids back to the combustor was higher than designed. This was because the existing plant layout required a difficult transport pipe layout to return the solids back to the combustor, including bends, reducers, and vertical pipe sections. Ideally, a commercial MBHE design would use only horizontal or slightly downward angled return piping. As shown in the blue trend line in Figure 1-11, the pressure drop through the horizontal portion of the transport line was well below the available PA fan head and thus supported the design philosophy of using only PA Fan air for the transport air.

Figure 1-11: MBHE Ash Return Pressure Drop

1.10. Recommendations

Although CMB's long-range potential was for high temperature Rankine cycles, ALSTOM Power considered CMB's near-term use as a lower cost CFB for conventional steam cycles to be a necessary stepping-stone in its development path. Since the total evaluated cost of the near term CMB design was not significantly less than a CFB for conventional steam cycles, ALSTOM Power decided to refocus its CMB development effort to accelerate the commercialization of the MBHE. The following recommendations are therefore made for further MBHE development:

- Previous work on MBHE tube-fin integrity limited the tip-to-base temperature difference to 250°F due to concerns of fatigue failure (e.g. separation of the fin from the tube). Subsequent work showed that tip-to-base temperatures of up to 400°F are permissible for cyclic operation where long-term creep failure is not a concern. Further work is needed to establish a temperature difference bound to ensure that creep-fatigue interaction does not result in premature failure of the fin or the tube.
- Additional MBHE testing should be continued at the Grant Town CFB after the plant evaluation of the cyclone sealpot return chute failure is completed. Testing needs to be completed at full solids flow through the MBHE to confirm performance across the full load range and to also assess the MBHE's impact on the overall CFB combustor performance.
- The heat transfer performance calculation techniques utilized for the Grant Town test series should be used to calculate performance of future MBHE applications. The MBHE design needs to incorporate baffles or tube bundle arrangements that minimize the ability of ash to bypass the finned portion of the tube bundle.
- The use of orifices in the distributing grid at the bottom the MBHE should be applied during an initial MBHE start up followed by selective orifice installation based on exit ash temperature profiles.
- The ACV provides a means of controlling ash flow and distribution with no moving parts in the ash flow stream. The ACV air actuating solenoid valves need to be selected for long-term high cycle operation. The metal seal legs need to be heat traced to prevent metal contraction during cool down for hot standby conditions.

2. INTRODUCTION

ALSTOM Power Inc. has been developing a novel Circulating Moving BedTM (CMB) combustion system. The CMB is expected to provide significant advantages in advanced power cycles and enable the achievement of improved efficiency, environmental, and cost goals, as targeted for new energy plants. The technology was envisioned as the successor to Circulating Fluidized Bed (CFB) technology. Specific plant systems were conceived which would utilize the CMB technology for near-term repowering applications as well as for the basis for achieving future advanced plant performance.

The U.S. Department of Energy (DOE) and ALSTOM Power jointly funded the CMB Proof of Concept program to evaluate and develop Circulating Moving BedTM combustion system technology. The objective of this program was to demonstrate the “proof of concept” for CMBTM combustion system technology, enabling it to be used as a key building block of new energy plants. Phase I of this project was initiated in July 2001 and completed in March 2003. The Phase I program cost was \$2,797,468 with the DOE’s National Energy Technology Laboratory (NETL) providing 60% of the funding under Cooperative Agreement No. DE-FC26-01NT41223. Participants included the U.S. DOE, ALSTOM Power, the University of Massachusetts, and the Massachusetts Institute of Technology.

A number of technical challenges were identified in Phase I that was the basis for continuing this work in Phase II. Phase II was initiated in April 2003 and completed in June 2010. The Phase II program cost was \$4,998,000 with the DOE’s National Energy Technology Laboratory (NETL) providing 60% of the funding under Cooperative Agreement No. DE-FC26-01NT41223. Participants included the U.S. DOE, ALSTOM Power, the University of Massachusetts, and the Massachusetts Institute of Technology. This report documents the results from Phase II of the CMBTM combustion system Proof of Concept program.

2.1. Circulating Moving Bed Technology Description

Circulating Moving Bed technology (illustrated in Figure 2-1) is a new method for solid fuel combustion and heat transfer, which has roots in the traditional circulating fluidized bed (CFB) technology. The CMB technology completely separates the combustion process from the heat transfer surfaces. Fuel is burned in the lower region of the combustor using bubbling fluid bed technology. The combustor burns coal or alternative fuels, such as petroleum coke, anthracite, sludge, and a variety of wood waste and biomass products to produce combustion temperatures approaching 1095°C (2000°F). These temperatures are higher than the combustion temperatures of 845 to 900°C (1550 to 1650°F) generated in traditional CFB boiler designs. The upper zone of the combustor is a relatively long residence time reactor that exchanges (recuperates) the heat from the products of combustion (upward flow) to a flow of high-density solid particles flowing downward. The solids have recuperated the heat of combustion upon reaching the bottom of the combustor and are then fluidized and transferred to a lower chamber by means of standpipes. The lower chamber contains a counterflow, direct contact “moving bed” heat exchanger. Heat from moving solid particles is captured in a series of tubular heat exchangers that preheat a working fluid such as steam or compressed air to the required process temperatures.

The moving bed heat exchanger uses a simple mass flow of solids that move downward at very low velocity. Sintered bauxite particles can be used as the heat transfer solids media because they have a high density, have good bulk flow properties, are chemically inert,

and are readily available. Figure 2-1 also shows an enlarged photo of a group of bauxite particles. The solids flow over tube circuits that contain the working fluid(s) of the cycle. The heat exchange is 100% counter-flow and, because the solids velocity is constant, there is no need to vary tube spaces.

The solids in the fluidized bed are primarily bauxite and contain a minimum amount of ash due to the high bauxite feed rates to the bed and because the fluidized bed acts as an effective classifier. The mass flow of bauxite into the fluidized bed is greater than the ash input. Furthermore, the ash and sorbent particles are considerably finer than the bauxite particles and are mostly entrained upward by the high bed fluidizing velocity. The larger and denser bauxite particles are retained in the bed and are then passed into the moving bed heat exchanger. At the bottom of the moving bed heat exchanger, the cooled solids are transported back to the top of the combustor to restart the "lower recycle loop".

The flyash entrained in the flue gas flows upward and is captured by a cold cyclone. Portions of the solids are returned to the bubbling bed with the balance of the solids going to a cooler and disposal (not shown). This flyash recycle is used to reduce carbon loss and to give additional control of the resultant bed temperature. The SO₂ emissions will be controlled primarily by a backend cleanup system such as ALSTOM's NID™ system. Limestone is calcined in the combustor for use in the backend desulfurization system and additional sulfur capture can be achieved in the combustor also.

Figure 2-1: CMB Combustion System Schematic

The CMB combustion temperature makes it possible to develop a design in which the working fluid can reach temperatures as high as 955°C (1750°F). In addition to enabling high temperature, high efficiency power plant cycles, the combustion temperature offers better carbon burnout, virtual elimination of N₂O emission, low carbon monoxide emissions, and hence, increased combustion efficiency with reduced pollutant emission.

2.2. MBHE vs. FBHE Comparison

An integral part of the Phase II study involved the development of the moving bed heat exchanger as a low cost replacement for a fluidized bed heat exchanger in a conventional CFB boiler. The MBHE has many attractive attributes including lower capital costs and lower power consumption. Figure 2-2 shows a comparison of a FBHE (left figure) and a MBHE (right figure). The FBHE requires a separate sealpot and uses an ash control valve before the FBHE to control solids flow. High pressure fluidizing air is required to fluidize the solids and to move them laterally through the FBHE. The FBHE uses bare heat transfer tubes.

The MBHE can be designed with a separate sealpot or can be integrated directly with the sealpot. The solids are not fluidized but pass through the MBHE by gravity flow. This eliminates the need for high pressure fluidizing air, as the primary air fan has sufficient pressure head to transport the solids to the combustor. The solids flow is controlled by an ash control valve located after the MBHE. The use of spiral fin heat transfer tubes in the MBHE contributes towards its compact design. The MBHE also has other potential applications, including CFB retrofits and being an enabling technology for chemical looping and O₂-fired CFBs.

Figure 2-2: FBHE vs. MBHE Comparison

2.3. Phase II Project Objectives and Workslope

The objective of the CMB Proof of Concept project was to identify the technical, design, and performance challenges that need to be met to make a commercial CMB system. The project objectives were initially achieved in Phase I through a series of experiments that were conducted in ALSTOM Power's Multi-use Test Facility. A number of technical challenges were identified that were the basis for continuing this work in Phase II.

The specific Phase II project objectives were set by ALSTOM Power, in concert with the U.S. DOE, and are summarized below:

- Improve the understanding of gas-to-solids heat transfer and then apply it to scale-up predictions for commercial sized units
- Develop recommendations for the tubing in the hottest parts of the MBHE
- Evaluate alternate mechanical designs and develop designs for critical components to be included in the commercial conceptual design
- Evaluate alternative techniques to enhance process performance
- Conduct MTF pilot plant tests to improve overall CMB performance with several fuels and to gather scaled heat transfer data for CFD validation
- Evaluate gas-to-solids heat transfer in a scaled MTF test and gather heat transfer validation data for the CFD process model. Apply the model to make scaleup predictions of CMB performance in a commercial scale combustor.
- Evaluate solids-to-tube heat transfer and solids distribution in the moving bed heat exchanger
- Evaluate factors to control ash agglomeration in the bubbling bed
- Evaluate CMB process performance, including carbon burnout, sulfur capture in the furnace and in the backend FDA system, and other gaseous emissions, including CO, NO_x, and N₂O
- Develop a commercial CMB boiler concept and cost evaluation
- Conduct extended duration MBHE field tests to evaluate long-term heat transfer performance and MBHE operational reliability

3. STATEMENT OF WORK

The following outline is adapted from the Statement of Work that was accepted by the DOE for Cooperative Agreement DE-FC26-01NT41223. The outline, along with the major task objectives, provides an overview of how the project was structured and executed.

The work included in Phase II of this project included tasks to further study of gas-to-solids heat transfer, finned tube material design for the CMB process, and mechanical and process design improvements. These tasks were a continuation of the work begun in Phase I of this project and included items that needed to be done in parallel with the MTF test program as data was developed.

Three tests were conducted in the MTF Test Facility of approximately 2 weeks duration each. The first two tests were focused on CMB process improvements. The third test was focused on gathering the detailed gas-to-solids heat transfer data necessary for CFD validation in a scaled MTF. A fourth MTF test was actually part of an independent DOE funded project for oxy-CFB firing, but this test provided an additional opportunity to gather MBHE heat transfer data using CFB ash rather than bauxite as the heat transfer media.

The commercialization task included a study of a commercial CMB boiler design and a cost comparison with a conventional CFB design. Most of this task though was focused on accelerating the commercialization of the MBHE through an extended field-test of the MBHE. The Grant Town CFB was selected as the host site and the MBHE was modified into a slice configuration to allow an evaluation of solids flow distribution through the MBHE. The MBHE was operated for 121 days to gather detailed heat transfer data and an assessment of long-term reliability.

The following list summarizes the Work Breakdown Structure for the CMB Proof of Concept project.

1.0 Gas-to-Solids Heat Transfer

The objective of this task was to improve the understanding of gas-to-solids heat transfer and then apply it to scale-up predictions for commercial sized units. This task included reviewing CFD models to ensure that the underlying physics were properly modelled. A scale-up methodology was developed for predicting heat transfer behaviour in larger units. Specialized instrumentation needed for CFD validation was identified and tested prior to the MTF scaled heat transfer test. The scaled MTF design and test plan was also developed in this task.

1.1 Test Design and Planning

1.2 Test Loop Design and Construction

1.3 Heat Transfer Testing and Solids Distribution.

1.4 Data Analysis

1.5 Scale up

1.6 Demolition

2.0 CMB Extended Surface Tubing

The objective of this task was to develop recommendations for the tubing in the hottest parts of the MBHE. The task included a preliminary evaluation of candidate materials, selection of the most likely candidates, and conducting low temperature heat transfer tests. An analytical stress model that provided the design criteria for the maximum allowable fin tip to tube temperature differential was developed and validated.

2.1 Preliminary Evaluation.

2.2 High Temperature Evaluation

3.0 Mechanical Design and Process Control

The objective of this task was to identify and address the mechanical design and process issues that needed to be resolved in order to develop a commercial concept for the CMB process. MBHE design concepts were evaluated and critical components were tested in physical flow models. Some of the component designs needed for the Multi-Use Test Facility and the process information needed for scaled operation of the test loop were also developed under this task. A MTF test campaign with two fuels was conducted to complete the performance testing of the Proof of Concept program.

3.1 Mechanical Design

3.2 Process Alternatives

3.3 MTF Test

4.0 CMB Commercialization

The objective of this task was to develop a commercial concept of the CMB plant, including plant arrangement and costs. A further objective was to accelerate the commercialization of the MBHE by installing and then testing a MBHE slipstream in a commercial power plant for extended testing.

4.1 Commercial Plant Concept

4.2 MBHE Commercialization

4.0 Program Management

The objective of this task was to plan the work for the various tasks and sub tasks, control expenses, and track the progress of the work done versus the schedule completion dates.

4. GAS-TO-SOLID HEAT TRANSFER CFD MODELING

The objective of this task was to review the computational fluid dynamics (CFD) models of gas-to-solid heat transfer previously developed during Phase I to ensure that the underlying physics were properly modelled. The proper grid size and number of solid injections were assessed to ensure reasonable predictions without excessive computational times. Scale-up procedures were developed to predict gas-to-solid heat transfer performance in larger units. Data measurement requirements were identified for CFD model validation along with the validation of specialized instrumentation. The originally proposed large-scale heat transfer test rig was re-evaluated and a decision was made to instead conduct the heat transfer validation tests in a scaled MTF. Validation tests were run in the MTF and then modelled by CFD to first confirm that the MTF would provide the necessary results for model validation and then that the models could be used for scale-up. The actual scaled MTF heat transfer tests and results are reported separately in Sections 9.4, 10.1, and 10.2.

4.1. Summary of Phase I Modeling Results

Figure 4-1 summarizes the technical approach that was taken during Phase I to evaluate gas-to-solids heat transfer. Heat transfer tests were conducted in two separate cold flow models to develop basic process data and understanding. A CFD model was then developed to model the cold flow test rigs and to predict heat transfer in the MTF. Hot gas-to-solids heat transfer tests were then conducted in the MTF. Those results were used to further refine and validate the CFD models, which was then used to make scale-up predictions for a large scale CMB combustor.

Figure 4-1: Roadmap for Scale-Up of CMB Gas-To-Solids Heat Transfer

Phase I Cold Flow and CFD Modelling

The first gas-to-solids heat transfer tests were conducted in a 9-inch test column at temperatures between ambient and about 250°F. A second test column with a 15-inch diameter cross-section was built and operated at the same temperatures. These data were then used to validate Computational Fluid Dynamics (CFD) modelling efforts.¹

A CFD model was developed for predicting heat transfer in the falling solids region and to predict the effect of aerodynamics on the solids distribution. The solids distribution affects the contact between the gas and solids streams and determines the heat transfer coefficients. The accuracy of the model's predictions depends on an accurate description of the underlying physics and on sufficient grid resolution to capture the gas-particle interactions. Of particular concern is the importance of particle-to-particle interactions, as this was not handled well by disperse-phase computational codes (e.g. – Fluent) at this time.

¹ *Circulating Moving Bed Combustion Proof of Concept Final Report – Phase I*, Glen Jukkola, DOE Contract DE-FC26-01NT41223, April 2003

Results from these test columns were analysed with Computational Fluid Dynamics (CFD) models. The initial CFD models were run with relatively coarse grids to keep computational times reasonable. The models were calibrated by adjusting the heat loss from the walls, the drag of the bauxite particles, and other calculation variables until selected measurable test data were matched. These results were then checked against other measured test data to determine if the model was predicting correctly. The models were not intended to predict every component of the process. The primary prediction needed was the outlet temperatures of the solids. With most of the test data, the CFD models did a good job at predicting the measured temperature of the solids.

Phase I MTF Heat Transfer Test and CFD Modelling of Results

The objective of the first MTF test campaign (Phase I – Sub Task 4.1) was to evaluate gas-to-solid heat transfer at a larger scale than the previous models and at temperatures representative of commercial CMB plants. The CFD model was again used to predict gas-to-solids heat transfer performance in the MTF tests as part of the test planning. The model predicted that the MTF would operate in a flow regime where there would be optimum conditions for a good solids distribution using a scaled up version of the cold tests' solids distributor. The model also predicted that there should be good gas-to-solids heat transfer performance. The MTF tests showed that the CFD model trends were correct but that details of the predictions needed improvement. The CFD model under predicted the actual heat transfer by 5 to 20%. Figure 4-2 shows one CFD model prediction vs. MTF test results. This implies that the method for modelling was generally correct but that the grid resolution and calibration needed refinement.

Figure 4-2: MTF Data vs. CFD Predictions of Temperature

Scale-up Predictions

The disperse-phase CFD model was also used to predict the performance of a commercial sized plant. This effort was made to determine if there were significant fluid dynamics scale-up issues with a commercial CMB process design. Using data from the existing test campaigns, the CFD model was calibrated and run to predict commercial scale performance.

The results of the commercial scale modelling pointed out some issues in using CFD for scale up. Using computational variables that predicted the MTF performance, the commercial model gave a wide range of predictions on gas and solid flow distributions and heat transfer performance. The predictions shown in Figure 4-3 illustrate the varying results as the grid resolution and the number of particle injections was changed. The predicted solids distributions range from all of the solids falling to one side of the combustor, to solids falling to the combustor outside perimeter, or to solids with a very good distribution throughout the combustor. Heat transfer predictions accordingly varied from bad to good among these flow conditions.

Figure 4-3: CFD Predictions of Solids Distribution in 300Mw CMB

Since the disperse-phase CFD models did not handle particle-particle interactions well, an attempt was also made to model the CMB with a Eulerian-Eulerian Fluent model. This model simulation did not provide any reasonable results. Clearly, the results with both disperse-phase and Eulerian-Eulerian CFD models required a better representation of the physical phenomena occurring in the combustor, particularly the particle-to-particle interactions. The grid dynamics needed to be better understood also. The grid resolution must be fine enough that predictions will not change as grid sizes are further reduced. However, large grids and a high number of particle injections result in excessive computational run times. CFD models can likely be used as a reliable tool for scale up predictions once the proper selection for grid size and the number of solids injections has been determined. The CFD models can then be calibrated and used for scaling up to larger combustor sizes with more confidence.

Recommendations for Large Scale Heat Transfer Test Facility

A large-scale heat transfer test was originally proposed for Phase II as a logical extension of the CMB development program. This facility would be used to develop the necessary heat transfer data to properly refine and then calibrate the CFD model so it could be used for scale up. Small-scale cold flow heat transfer tests in a 9-inch and a 15-inch reactor showed good gas to solids heat transfer. Hot tests up to 2000°F in the 40-inch diameter MTF furnace also showed good results. CFD model predictions of these tests gave encouragement for scale-up to demonstration and commercial sizes. However, the demonstration plant size is a significant jump in size from the MTF scale and CFD modelling still had questions that

needed to be addressed with respect to particle-particle interactions and grid dynamics. The models also needed the validation data from a larger scale unit before they could be used with confidence. It was therefore proposed to develop an intermediate size test to determine scale up procedures before going into the costly process of building a demonstration plant. A further advantage of this size was that several critical components could also be tested at significant scale, including the solids distributor.

An 8-foot diameter heat transfer test loop was proposed as a scale that would help to characterize the aerodynamic response at an intermediate size between a pilot or demonstration plant. The effective heat transfer height would be 50 feet. This size was anticipated to be cost effective for a test program since it could be built using existing structures and infrastructure at ALSTOM Power's Windsor site. This size would also enable a reasonable extrapolation from the 8 foot diameter to a commercial diameter of 30-50 feet.

Figure 4-4 shows a schematic of the proposed Large Scale Heat Transfer Test Loop (LSHTL). Hot gas would be generated in an existing combustion test facility and conveyed to the LSHTL's inlet plenum and overfire air ducts. The hot gas would be cooled as it rises through the heat transfer column by the falling solids and then exhausted to an existing stack. Bauxite would be fed from the solids separator through a solids distributor and then into the heat transfer column. The solids would fall through the rising gas and pass through the open column bottom. They would be collected in a seal pot and then transferred to a direct contact cooler, where water would be sprayed directly on the hot bauxite particles. The resulting steam would be vented off into an existing stack. The cooled solids would then be pneumatically transported back to the top of the facility to repeat the process.

Figure 4-4: Proposed Large Scale Heat Transfer Facility

The MTF has a 3.33-ft diameter, making the proposed test loop approximately a 5.8 to 1 scale up in cross-sectional area. A 30Mw_e demonstration plant would have a cross section scale-up of about 4.6 to 1 from the proposed test loop. Without this proposed test loop, the scale-up from the MTF to a demonstration plant would be around 26 to 1. The proposed large test facility would put the larger scale up on a relatively inexpensive test loop instead of the first demonstration plant.

The test loop would be built to operate at a lower temperature than the MTF. The temperature would have a limit of 1000°F to allow the unit and auxiliaries to use less costly materials. This temperature would still give the test loop a scale up factor that was a nice intermediate between the MTF and a theoretical 30Mw_e demonstration plant. Table 4-1 lists the scale up progression through Phase I in the CMB test program.

Table 4-1: CMB Test Data Scale-up Progression

Item	Size			Area Scale up
	cross section	area - sq.ft.	height	Factor
First Test Rig	9x4 inches	0.25	15 ft	-
Second Test Rig	15 inch dia.	1.23	15 ft	4.92
MTF Test	40 inch dia.	8.73	45 ft	7.10
Proposed Large Scale Heat Transfer Test	8 ft. dia	50.2	50 ft.	5.75
30 Mw Demo (concept)	17.1 ft. dia.	230	100 ft	4.58
300 Mw Commercial (concept)	54 ft Dia	2300	100 ft.	10.00

Several effects of scale-up would be investigated. The gas flow distribution or mal-distribution caused by the falling solids over a larger cross-sectional area would be investigated. The larger area allows the solids to move farther laterally from the distribution point, demonstrating the falling solids tendency to affect gas flow velocity variations. These velocity variations affect heat transfer performance. Measurements of the velocity and solids distributions would be used to calibrate the CFD model at a larger scale.

4.2. Phase II Modeling Approach

4.2.1. Original Proposed Phase II Modeling Approach

Figure 4-5 shows the original roadmap for scaling up gas-to-solids heat transfer that was proposed for Phase II. An initial CFD modelling effort would be undertaken to refine the CFD model by evaluating both grid independence and particle tracking. The model would then be used to make scale-up predictions for the proposed large-scale heat transfer test facility. In parallel, design options would be evaluated for the large-scale test facility, followed by detailed facility design and construction. The facility would be operated to generate the heat transfer data necessary for CFD model validation. The model would then be used to scale-up the design to large-scale commercial sized plants.

Figure 4-5: Original Phase II Roadmap for Scale-Up of CMB Gas-To-Solids Heat Transfer

4.2.2. Revised Phase II Modeling Approach

At the Phase II project kick-off meeting, the project team was challenged to consider whether a scaled heat transfer test combined with improved CFD modelling might be a viable alternative to the proposed large-scale heat transfer facility. Figure 4-6 shows a revised Phase II roadmap for scale-up of CMB gas-to-solids heat transfer. A new parallel path was added to identify the CFD validation data needs and to identify qualified instrumentation vendors that could provide the instruments needed to collect the detailed validation data. A scaled MTF test was added to the various heat transfer test facilities that were being evaluation. As part of this evaluation, the MTF was operated under conditions of both good solids distribution and intentionally poor solids distribution. Subsequent CFD modelling showed that the CFD model could adequately model the MTF behaviour. After demonstrating grid independence and identifying the proper number of particle tracking, further CFD scaling predictions no longer showed large solid/gas imbalances in large-scale facilities as previously thought at the conclusion of Phase I. Thus, there was less of a need for the large-scale test facility. The Phase II program was therefore revised with the DOE's agreement to conduct the detailed heat transfer tests in a scaled MTF facility rather than in the proposed large-scale test facility.

The scaled MTF test facility had a number of advantages over the large-scale test facility. CFD validation data could be gathered much easier in the scaled MTF. The facility was well instrumented, flexible, had a smaller diameter to probe, more control over temperatures, plus optic measurement techniques could be more readily applied at this scale. The scaled MTF facility could be operated around the clock more cost-effectively (gas costs are very high for larger facilities).

Figure 4-6: Roadmap for Scale-Up of CMB Gas-To-Solids Heat Transfer

5. CFD MODELING VALIDATION

The CMB roadmap for scale-up relied on a combination of CFD models and experimental validation test data for predicting performance of commercial sized CMB combustors. The CFD models first needed to be further refined in regards to grid independence, number of particle trajectories, material properties, and assumptions. The models then needed to be calibrated with existing test data from smaller scale test facilities and with detailed validation data from a scaled MTF test. The CFD simulations could then provide insight into flow patterns and heat transfer performance for large units and be used for scale-up predictions. The following sections describe the tasks related to scale-up methodology, CFD model refinement (grid independence, particle tracking), scaling predictions, validation data requirements and techniques, and MTF validation tests. The actual MTF scaled tests will be reported later in Sections 9.4, 10.1, and 10.2.

5.1. Scale-up Methodology

A meeting was held with Professor Leon Glicksman of MIT in June 2003 to discuss scaling issues with the CFD modelling. The primary issues were aerodynamics and how this affects heat transfer. The meeting focused on the status of the CFD modelling to date and on scaling techniques.

The main goal of the CFD modelling work was to develop a practical model for calculating the performance of a CMB demonstration plant. The models under development included a one dimensional performance model and a CFD model to predict the aerodynamics and heat transfer. These models were needed for scaling the existing data from the test rigs to a larger size. The scaling work required validation that the models could predict performance in larger units. A necessary step was to determine what test data was required to calibrate the models.

The problems with the CFD modelling and scale-up concern the preliminary predictions made during Phase I that there might be flow distribution problems with the falling solids. Most of the Phase I CFD modelling was done just as a visualization tool. Early CFD models were calibrated to match the heat transfer from the low temperature test rigs. Using these calibrations, a model of the MTF predicted that the MTF would have very good aerodynamics and therefore reasonably good heat transfer. This was the case with the MTF tests. The heat transfer predictions of the CFD model were reasonable also.

In predicting the performance of the larger units, the CFD modelling was more ambiguous. As the height of the units modelled increased, there was more of a tendency for the model to predict a flow distribution problem. Varying some of the model settings could change the model results.

In order to calibrate the CFD model, Professor Glicksman was asked to comment on scaling techniques. Professor Glicksman previously evaluated procedures for scaling results of the MTF pilot plant to a commercial sized unit. He concluded that no proper scaling was possible at a temperature of 2000°F while using the proper sized bauxite particles. If the MTF was run at CMB operating temperatures and proper sized bauxite particles were used, then the full set of scaling relationships indicated that the scale model must be the same size as the commercial model, i.e. no linear scaling could be accomplished. Furthermore,

scaling the complete process was not practical since there are many processes in the CMB concept, including heat transfer, combustion, particle trajectories, and many others.

For this situation it can be shown that the linear dimension scales with the kinematic viscosity of the gas raised to the 3/2 power. For the full set of scaling variables, the MTF bed must be operated at a lower temperature so that the kinematic viscosity will be reduced. In this situation, the cold MTF will then simulate a larger hot riser. This would also require the use of solid particles that are denser than the bauxite.

Glicksman had also developed a simplified set of scaling relationships², which had promising application here. The full set of scaling parameters can be relaxed for most cases. The simplified technique required matching the Froude number, the solids to gas ratio, the ratio of gas superficial velocity to particle terminal velocity, the ratio of linear bed dimensions, the dimensionless solid flow rate, the particle sphericity, and the dimensionless particle size distribution. The simplified set requires that the following parameters be matched between the test and the larger units:

$$\frac{U_0^2}{gL}, \frac{\rho_s}{\rho_g}, \frac{U_0}{U_T}, \frac{L_1}{L_2}, \frac{G_s}{\rho_s U_0}, \phi, \text{particlesizedistribution}$$

The terms in this relationship are the Froude number based on the bed dimensions, the solids to gas density ratio, the ratio of gas superficial velocity to particle terminal velocity, the ratio of linear bed dimensions, the dimensionless solids flow rate, the particle sphericity, and the dimensionless particle size distribution, respectively. In the original derivation, the third term was the ratio of gas velocity to the minimum fluidization velocity. For the counterflow riser, the terminal velocity is the more physically meaningful velocity. This allows the dimensionless drag coefficient on a single particle to be matched.

To use this relationship, the linear scaling ratio between the test model and the commercial bed is first defined. From the Froude number, the test superficial velocity U_0 is calculated so that the test and the commercial bed have the same value of the Froude number. The terminal velocity is then calculated from the third term so the velocity ratio is the same for the test and commercial bed. The required mean particle diameter for the test bed is then found from the terminal velocity.

Table 5-1 shows an example of using this simplified scaling relationship for a linear dimension scaling of 4:1, with both beds at 2000°F and using appropriately sized bauxite particles.

² *Simplified Scaling Procedures*, Leon Glicksman et. al., Powder Technology, 1993, p 177-199

Table 5-1: Example of Simplified Scaling Relationship

<i>Parameter</i>	<i>MTF(test)</i>	<i>Commercial</i>
Diameter	40 inches	13.3 feet
Height	25 ft	100 ft
U _o	10 ft/s	20 ft/s
Particle diameter	0.48 mm	0.9 mm
Terminal velocity U _T	5.25 m/s	10.5 m/s
Minimum Fluidization velocity U _{mf}	0.305 m/s	1 m/s
U _T /U _o	1.72	1.72
U _T /U _{mf}	17.2	10.5
Cluster diameter (assumed)	2.5 mm	10 mm
U _T (cluster) / U _o	8.5	10

Note that the MTF must be modified so that the height of particle travel is only 25 feet to obtain the correct scaling. The terminal velocity scales by a factor of 2. For a 0.9 mm particle in the commercial bed, the test bed must have a particle size of 0.48 mm. The minimum fluidization velocity does not scale properly. It has a 3.2:1 ratio rather than a 2:1 ratio. However, in the riser the proper scaling of the forces for a bed of dense packed particles does not appear to be relevant. Assume that the cluster sizes scale with the bed linear dimensions. If the commercial bed has spherical clusters of 10 mm diameter and the MTF clusters of 2.5 mm diameter, then the terminal velocity of the clusters in the commercial bed and the MTF test bed scale reasonably closely with the terminal velocity of individual particles.

Carrying out the same calculations assuming that the particles in the commercial bed are 0.7 mm in diameter gives similar results. Again using a 4:1 linear scaling ratio, the scaled particles in the MTF must be 0.395 mm to yield a 2:1 scaling of the terminal velocity. The particle diameter ratio of the two beds is 0.56/1 whereas using 0.9 mm particles the particle diameter ratio is 0.53/1. This indicates that close scaling will be obtained for all particle terminal velocities by scaling all particle diameters by the ratio of 0.53/1.

Another consideration is the scaling of the heat transfer results. The transit time of the solids from the top to bottom of the riser must be compared to their thermal time constant. In this case the thermal time constant of single particles will be considered. The transit time for the MTF is given as:

$$\tau_{transit(test)} = \frac{U_T - U_0}{L} \sim \frac{\sqrt{L}}{L} \sim \frac{1}{2} \tau_{transit(commercial)}$$

For heat transfer to a single spherical particle, the thermal time constant is,

$$\tau_{thermal(test)} = \frac{\rho_s c_s \pi d_p^3 / 6}{\pi d_p^2 h} = \frac{\rho_s c_s d_p}{6h} \sim \frac{d_p^2}{Nu}$$

Using a correlation for the Nusselt number for a sphere of 0.9mm diameter from Rohsenow and Choi, the Nusselt number is 5.88 for the commercial bed and 3.51 for the MTF. The

ratio of the thermal time constants between the MTF test bed and the commercial bed is 0.48, almost exactly the same as that for the transit time constants when the effective MTF height is reduced to 25 feet. This suggests that if the heat transfer is primarily from the gas to dilute particles, then the MTF will accurately simulate the heat transfer as well as the bed dynamics.

This analysis reinforces the possibility of using the MTF to model a small demonstration plant. Using the simplified scaling technique, it should be possible to design a test program at the MTF, which is 40 inches in diameter, to model a roughly 25Mw_e commercial plant at 192 inches in diameter. The scale would be 4.8 in linear dimension. A 100-foot high 25Mw_e plant would be modelled by about 20.8 feet of the MTF height, which is feasible. The bauxite size would need to be smaller, depending on the availability of these sizes.

The meeting with Professor Glicksman generated the following action plan. It was decided to investigate running a scaled heat transfer test of a 25Mw_e CMB reactor at the MTF. This would provide a method to calibrate the CFD model. In order to do this, the CFD model first needed to be validated and checked for several concerns. The CFD model needed to be run to evaluate grid independence (see Section 5.2). This required a CFD model to be run with increasing amounts of cells or control volumes, until the solution didn't change. The model also needed to be checked for increasing the number of particle track calculations in a similar manner (see Section 5.3). Finally, the scale technique needed to be modelled for the MTF scale model and for the theoretical 25Mw_e CMB reactors (see Section 5.4).

5.2. Grid Independence

The grid independence study used two models, one of which was a conceptual 25Mw_e CMB, and the other was a scaled down version of the 25Mw_e unit that could be run in the MTF. A grid comparison was done for the 25Mw_e case at 310,000, 1,400,000, and 2,800,000 cells; and for the MTF case at 116,000, and 900,000 nodes. Both models showed essentially the same particle concentration profiles at all grid levels, indicating that the models were 'grid independent'. Figure 5-1 compares the solids distribution in the preliminary grid size comparison for the 25Mw_e CMB case. Figure 5-2 compares the solids distribution in the grid size comparison for the 40-inch diameter scaled MTF case.

Figure 5-1: Grid Size Comparison for 25Mw_e CMB

Figure 5-2: Grid Size Comparison for Scaled MTF

An issue with the grid independence test was that the larger models required a significant amount of computer time. The larger models took weeks to get results, even with the use of a computer cluster using ten, state of the art CPU's. It was also not clearly defined as to when the cases were adequately converged. The grid independence runs were continued until it was clear that the general results were not going to change.

The preliminary 25Mw_e grid independence cases showed that there were differences between the cases with different cell counts. However the differences were minor in comparison to the cases that previously had shown major flow unbalances. All three cases shown had very good solids distribution, even if the distribution was not perfect. This indicated that the solution was grid size independent.

Another feature of the model that was not shown was that the model results changed as the number of iterations increased. As the models were reviewed at various numbers of iterations it could be seen that the same flow pattern and solids distribution rotated around the vessel. This illustrated the problem with calculating particle tracks. The results could mean that the calculation was not mathematically stable or that the flow field and the particle field were not stable. Even though one would expect that the reactor flow field might behave this way in reality, it was not clear that this is what the model is calculating.

5.3. CFD Particle Tracking

Determining the number of particle tracks that were needed was another CFD question. The number of particle tracks that the model uses does not determine how much solids flow there is. The particle tracks simply determine the position in the flow field that the particles exchange momentum and energy with the gas flow. By using a larger number of particles, the number of places where the particles exchange momentum and energy increase. Since the solids flow is composed of millions of particles, it would seem that the larger number of particle tracks would provide a more representative solution of the model. The larger number of particles tracks takes more computer resources, however. A practical limit had to be found.

The 25Mw_e and MTF models were also run with varying numbers of particle track calculations. 240, 2400, and 240,000 particle track calculations were tried for each model. The results for each model were essentially the same regardless of the number of particle tracks. In all cases, the model results showed that the particle distribution was fairly well dispersed. However, the lower number of particles showed more of a tendency toward grouping of the particles, which may or may not reflect the real case. Figure 5-3 compares the solids distribution results of the particle track simulations for both the 25Mw_e and MTF cases.

Figure 5-3: Particle Track Comparison for 25Mw_e and MTF Models

5.4. Scaleup Prediction

A comparison of the 25Mw_e and the MTF scaled down models using Glicksman's simplified scaling criteria is shown in Figure 5-4. The 25Mw_e case used 310,000 grids and the MTF case used 116,000 grids. The solids distribution was very similar between the two cases, which supports the use of CFD models to capture scale-up similarities. The simplified scaling criteria required that the temperature of the two cases must be the same. This was not really the case for these models. The criteria required a different sized particle for the models according to the scaling formula. This results in a different relative particle surface area and in a different heat transfer coefficient. Using the same inlet temperature for both models resulted in a different outlet temperature. The temperature difference, however, didn't result in enough of a change in the gas density to totally invalidate the results. While the models were not scaled exactly they were very close.

Figure 5-4: Scale-up Comparisons

One of the issues that needed further investigation was why the models in this study were not showing the tendency to predict an unbalanced gas and solids distribution. It was decided that the best way to verify the model was to run an MTF test where the flow distribution was intentionally forced to be unbalanced and then to model that condition. The model would then be calibrated and more confidence could be placed on it. This special test condition is described further in Section 5.7.

The following conclusions were drawn from this CFD study:

- The CFD results predicted much better solids distribution than previously reported in Phase I of this program. The Phase I modelling results that showed severe particle and gas separation probably was due to incorrect model settings coupled with too low a number of particle track calculations.
- This study identified the number of cells necessary for grid independence and the minimum number of particle tracks necessary for a reasonable solution. Using these model guidelines, the CFD models predicted that the solids distribution would be relatively uniform in a 25Mw_e combustor and that the 25Mw_e plant could be simulated in a shortened MTF using simplified scaling criteria.
- While a larger number of cells and particle track calculations were desirable, the time and computer resources required to converge them was a serious issue. The largest model took 10 state-of-the-art processors and several weeks to run.
- For heat transfer calculations, it was sufficient to use about 240,000 particle track calculations. The grid size should be as high as is practical, which means whatever can run on four processors.
- For just flow visualization, half a million cells and about 2400 particle tracks would be acceptable and would run on one processor.
- The flow characteristics of the 25Mw_e conceptual plant could be modelled in the MTF.
- The flow characteristics of the 25Mw_e plant should have a relatively even solids distribution.

5.5. CFD Validation Measurements

Given the decision to use CFD modelling for scale-up predictions, it was clear that additional measurements were needed to validate the models. Existing measurements in the physical flow models and the MTF were not capturing certain features of the flow field. The model calibration required additional data points across a wider range of operating conditions. In addition, more detailed data was required to fully define the flow field variations. Fortunately, the use of the scaled MTF allowed greater temperature control than a larger test facility and made it easier to obtain more detailed data profiles.

A review of measurement methods on the MTF identified the data that could be easily measured:

- Gas and solids mass flow rates
- Gas and solids inlet and outlet temperatures
- Gas temperature profiles

The following additional data was needed to define flow field variations. This data was more challenging to acquire:

- Gas velocity profiles at several elevations
- Solids density in flight

- Solids temperature profiles

An effort was launched to identify suppliers of “developed” technology to locally measure velocity, concentration, and temperature of the solids in the CMB process. Vendors were contacted and asked if they had instrumentation that could take the required measurements given the operating conditions in Figure 5-5. Figure 5-6 shows the list of organizations and vendors that were contacted for either specialized instrumentation or guidance.

Maximum Gas Temperature	2000°F
Maximum Gas Velocity	10 ft/sec
Average Solids Concentration	0.01 lb/ft ³
Maximum Solids Concentration	0.3 lb/ft ³
Average Particle Size	400 μm
Average Particle Density	225 lb/ft ³

Figure 5-5: CMB Operating Conditions

REI
 Sandia
 Lawrence Livermore
 Brigham Young University
 University of British Columbia
 Micro-Epsilon
 Chalmers University
 University of Hamburg-Harburg
 Dantec Dynamics
 MSE Meili

Figure 5-6: Organizations Contacted About Specialized Instrumentation

The vendor search was narrowed down to three potential contractors with “developed” technology. Dantec Dynamics offered laser-measuring techniques, including a 2-D LDA and laser sheeting system, MSE Meili had a backscatter probe, and Chalmers University offered LDA, Capacitive, and Momentum Probes. In all cases, it was concluded that the semi-opaque CMB environment required intrusive measurements and that the temperature of the sampling environment was a major issue for all of the measurement technologies. After discussions with each vendor, it was decided to work with MSE Meili and to conduct some cold flow validations tests with their optic probe to confirm that it would provide the necessary data.

5.6. Optic Probe Cold Flow Validation

Test measurements with the MSE Meili Labasys® 100 measuring system were conducted in August 2003 to check its feasibility for solids concentration and solids velocity measurements of bauxite particles in the MTF hot gas scaling test. Cold flow measurements were conducted in a 14-inch diameter by 28 ft high cold flow model. Figure 5-7 shows a schematic of the optic reflectance probe on the left side, while the right photo shows the Labasys probe inserted into the test column.

Figure 5-7: MSE Meili Probe Schematic and Test Installation

A series of 12 tests were conducted in the PPL ambient column facility to evaluate the optic reflectance probe supplied by Dr. Reto Meili. The probe was traversed across the column to measure the local concentration and velocity of the particles. The measured values could then be integrated and compared with average values obtained from PPL instrumentation. Two sizes of particles were used for the evaluation; 700 microns for conventional test operation and 300 microns for future laboratory scale-up test operation. Solid flow rates were varied from 1,000 to 10,000 lb/hr and superficial gas velocities ranged from 3.5 to 15 ft/sec.

Figure 5-8 on the right shows the measured solids concentration across the width of the test column during one of the tests. These results were taken with 2000 lb/hr of the 300-micron particles being introduced into the top of the test column. The air rise velocity in the column during this test was 6 ft/sec. The sharp increase in solids concentration at 0.35m indicated a heavy loading of solids raining down that wall.

Figure 5-8: Solids Concentration Across Column Width

Eight detailed profiles were measured during this experiment. Each profile included 10 different radial measurement positions resulting in 20 or more data sets, since two measurements were taken at each position. For each data set 500,000 points per measuring channel were sampled at a sampling rate of 12 kHz, resulting in over 400 MB of raw data.

The results showed that the solids concentration profiles were quite flat; with higher values observed towards the walls. In most cases, the concentration in the center of the tube was low, higher near the wall, and much higher with increased solid flow rates at the wall. Certain runs with the finer bauxite showed that the velocity profiles had a maximum in the column center with positive values and minima towards the walls with negative values. Depending on the solids loading and the gas velocity, the profiles were more or less pronounced. As the solids flow increased, the concentration data was enhanced as well, while the velocity was reduced.

Figure 5-9 shows two sets of solids velocities and concentrations measured with the different bauxite sizes. The solids concentration in both figures is shown by the red and blue data points near the bottom of the figures, while the upper trend lines represents the velocity data. The left figure shows that the solids velocity (blue & green lines) with the 700-micron particles was always negative (or downward) near the walls and near zero in the middle of the column. The right figure with the 300-micron particles also showed negative velocities near the walls, but the positive solid velocities in the center of the column indicated that the particles were actually rising. Closer analysis showed that some of the particles in the 300-micron batch were considerably finer than 300 microns. The local gas velocity at the column center was greater than the terminal velocity of the finest particles, which accounted for the rising particles.

Figure 5-9: Solids Velocity and Concentration Measurements

The two samples of bauxite were screened in the laboratory after the cold flow tests were completed to determine the actual bauxite particle size distribution. The size distribution of the small particle size was lower than anticipated, 250-micron vs. 320-micron. The smaller size distribution caused an increase in particle concentration and a decrease in migration velocity at the center of the test column. The net result was that the instrument gave quantitative measurements in more detail than the average methods used previously by PPL. There was reasonable correlation between the Meili instrument and the PPL average methods when the conditions did not produce abnormally high centerline concentrations.

The following conclusions were drawn from these test results:

- Most of the solids concentration values were in a low concentration range of 0.02-0.5 vol.-%.
- Besides a few exceptions where dense 'particle strands' moved downwards quickly, the solids showed flat concentration profiles with a tendency to rise towards the walls.
- The measured mean velocities were mainly in the range of -3 m/s close to the wall and rising in the center of the tube to maximum values between 0 - 2 m/s depending on the operating conditions, i.e. gas velocity, solids loading, and particle size.
- The measured data was in reasonable agreement with the solids holdup calculated from pressure difference measurements and with the overall mass flow rate. Analysis of these test results indicated that the MSE Meili Labasys[®] 100 measuring system was a feasible tool for characterizing gas/solid flows during the CMB hot gas scaling test campaign.

Based on these results, plans were finalized to use the laser/fiber particle concentration/velocity measurement device from MSE Meili during the scaled MTF test program. The probe was modified to include a local gas temperature measurement in addition to particle concentration and velocity.

5.7. CFD Model of MTF with Deflector Plate

Although confidence was being gained in the use of CFD models and in the measurement techniques, there was still concern because the model was not predicting any flow distribution problems at all in the model predictions. It was therefore decided to conduct a brief test during the November 2003 MTF test (prior to the scaled MTF heat transfer tests)

where the solids flow distribution was intentionally disrupted and then see if the CFD model could match the resulting flow distribution.

A CFD model was initially created to model the flow pattern in the MTF combustor with a deflector plate under the solids injector. The reason for the deflector plate was to artificially induce a poor solids distribution to confirm that the CFD model could predict it. The model was configured in advance of the MTF test to ensure that the proposed deflector had a big enough effect on solids distribution to measure. Figure 5-10 shows the CFD model geometry that was configured for the MTF cases with the solids being deflected and undeflected, respectively.

Figure 5-10: CFD Model Geometry

Figure 5-11 shows the model's predictions for the gas temperature profile and solids distribution along the height of the MTF combustor with the solids deflector plate in position. The results showed that the proposed deflector plate would do its job very well. In fact, the deflector would probably deflect so much of the solids that heat transfer to the solids would be very low and that the associated temperature measurements across the MTF diameter would be too uniform except at the walls. The prediction suggested that it might be necessary to adjust the deflector plate angle during the actual MTF test to reduce the amount of solids deflection.

Figure 5-11: MTF Temperature and Solids Distribution Predictions While Using A Deflector Plate

The CFD calibration strategy was to run the MTF under two or more separate falling solids distribution conditions. To establish these conditions, an adjustable diverter plate was installed under the solids distributor. This plate could be moved to direct the solids over to one side of the combustor. This caused a significantly different solids and gas flow pattern at each plate angle.

The CFD model would be used to simulate the measured test conditions. Several critical constants would be changed by trial and error to get the best match with the data. The calibrated model would then be used to predict other test data. Several of the more fundamental variables to be calibrated included: particle coefficient of restitution after hitting a wall, particle effective aerodynamic sphericity, radiation adsorption coefficient, index of refraction, and other mathematical terms needed in the CFD code.

The solids deflection test was conducted as part of the November 2003 MTF test campaign. Figure 5-12 shows some of the MTF test data that was collected during the solids deflected and undeflected tests for use in the CFD model.

		deflected		undeflected
	Date	11/3/2003		11/4/2004
	Time	9:56 PM		1:33 AM
	MFMBHX	1038.01		2268.68
	MFTotalTair	8549.31		8656.95
	TE133	988.06		806.80
	TE134	1021.93		971.74
	TE135	855.05		819.53
	TE136	930.63		891.48
	MFFlueGas	9223.39		9332.35
	Qpan4	23242.00		22831.00
	Qtop	-695.00		70144.43
	Qpan2	0.21		0.00
	QCombLoss	150961.00		159366.61
elev 22	TE159	1854.21		1607.27
	TE160	1714.00		1717.00
	TE161	1749.00		1730.00
ave elev 22		1772.40		1684.76
Cyclone Rake Top	TE391	1012.53		961.41
	TE392	1012.78		960.22
	TE393	1007.00		958.65
Cyclone Rake bot	TE394	974.70		928.56
ave cyclone rake		1001.75		952.21
	TE128	80.00		78.90
solids T	TE129	375.92		379.85
	MFFBHE	4867.71		5048.37
	MFMBHXFlow	10906.08		11006.89
delta t		770.65		732.55

Figure 5-12: MTF Test Data for CFD Model

Figure 5-13 shows the CFD model temperature prediction for the 52-foot elevation for the November 6th MTF test condition. The CFD model used a simple particle injection model and did not account for gas/solids interactions near the solids distributor. Not too surprisingly, the initial model results did not match the actual test data very closely.

Figure 5-13: CFD vs. Data Using Simple Particle Injection (11/6/03 data)

After accounting for all of the MTF heat losses, the predicted temperature for the deflected case was less than 30°F lower than the recorded measurement. Overall, the total heat transfer prediction of the CFD model was less than 16% different from the data and in one case less than 6%. The heat transfer in the model was slightly over predicted in all cases run so far.

However, the results at the three furnace elevations where temperature traverses were made showed more variation than expected. One problem was that the traverse data showed a reverse trend from the combustor gas outlet temperatures. The tests with the deflected solids cases showed lower temperatures in the temperature traverses than the non-deflected cases. The CFD model predicted the reverse trend, which was expected.

The CFD model also predicted a larger temperature gradient than the traverse data. The traverse data represented average temperatures over a finite time period. The variation in the data at any given location was as high as 32°F, with an overall standard deviation of 5.2°F. The CFD results showed a steeper gradient than would be expected from the traverse data. This was actually a conservative result from a design standpoint since it predicted a slightly worse condition than measured.

In the model, the flow gradient at the traversing levels also showed another gradient in the direction perpendicular to the traverse direction. This gradient was larger than the one that followed the traverse and would not be detected by the traverse. If the flow field were constantly changing, this gradient would shift and cause the measured traverse data to vary, as was seen.

Based on the initial calibration effort to match the overall heat transfer performance, it was concluded that the CFD model correctly predicted that the deflected solids case should have a higher outlet temperature than the non-deflected case. However, temperature traverses taken at several locations in the combustor showed that the deflected case had lower temperatures in the reactor vessel than the non-deflected case.

This apparent discrepancy implied that there was more heat transfer at the solids injection point with the non-deflected case than with the deflected case. The model had originally modelled the solids injection point in a very simple manner because it was assumed that the injector was the starting point for the heat transfer. The solids inlet pipe and the shroud around the inlet pipe were not modelled in detail. In order to analyse the apparent discrepancy in the model's predictions, the model was rebuilt to show the details of the solids injection system.

A new CFD model was made that gave better solids distribution and predicted the proper temperature profile trends with both normal and non-uniform solids distribution. The overall heat transfer, however, was too high in the CFD model. The Nusselt number calculation in Fluent needed to be modified to reduce the solids to gas heat transfer coefficient.

The focus was on adjusting the calculated Nusselt number to match the November MTF testing. The calibration showed that a Nusselt number about half the theoretical value for a single particle gave the best heat transfer results. This was consistent with experimental heat transfer results obtained in the MTF also.

Figure 5-14 shows the predicted particle trajectories for the CFD model of the deflected case. The model predictions for the gas velocity and gas temperature for this case are shown in Figure 5-15 and Figure 5-16.

Figure 5-14: Particle Trajectories – Deflected Case

Figure 5-15: Contours of Gas Velocity - deflected

Figure 5-16: Contours of Gas Temperature - deflected

Figure 5-17 shows the predicted particle trajectories for the CFD model of the undeflected case. The model predictions for the gas velocity and gas temperature for this case are shown in Figure 5-18 and Figure 5-19.

Figure 5-17: Particle Trajectories - Undeflected Case

Figure 5-18: Contours of Gas Velocity - Undeflected

Figure 5-19: Contours of Gas Temperature - Undeflected

The modified CFD model with the detailed solids injector now predicted the main trends from the MTF correctly. The temperature trends predicted by the model across the combustor cross-section were close to the measured traverses for side-to-side variation. The model did a good job matching the results from the angled deflector test. The results with the vertical deflector position did not match as well. This could result from problems with the measured test data, which appeared to be inconsistent when compared with other tests. Figure 5-20 shows the comparison of the MTF data and model predictions for gas temperature for the undeflected and deflected cases.

Figure 5-20: CFD Results Using Detailed Injector

Based on the calibration results of the MTF test, it was concluded that the CFD correctly modelled the overall trends observed in the MTF for the deflected solids case with a reduced Nusselt Number. The Nusselt Number was about 50% of the calculation for single particle heat transfer. Both the CFD model and data analysis confirmed the Nusselt Number magnitude. The CFD model also successfully modelled the scale-up methods developed by Dr. Glicksman. The plans to move forward with the scaled heat transfer tests in the MTF were therefore finalized. The MTF test and the CFD modelling the results are described later in Sections 9.4 and 10.2, respectively.

6. EXTENDED TUBE EVALUATION

The objective of this task was to develop recommendations for the tubing in the hottest parts of the MBHE. The task included a preliminary evaluation of candidate materials, selection of the most likely candidates, and conducting low temperature heat transfer tests. An analytical stress model was also developed and validated that provides the design criteria for the maximum allowable fin tip to tube temperature differential.

6.1. Stress Analysis Model Development

The objective of this task was to establish MBHE process conditions and heat exchanger requirements by reviewing the fluid and solids temperatures and pressure conditions for both subcritical and advanced Rankine cycle designs. Materials and fin designs to meet these conditions were selected. Heat transfer tests were conducted in a low temperature test loop. An analytical stress model that provides the design criteria for the maximum allowable fin tip to tube temperature differential was also developed.

Stress Modelling

A stress model was developed to accurately determine the design limits of spiral-finned tubing, specifically limitations imposed on fin tip to fin base temperature difference. Previous finite element modelling work was reviewed and used as a basis for a new approach to develop an analytical, parametric model to allow design trade-offs and key variables to be evaluated more rapidly. Analytical models were developed to study the heat transfer and thermal stress in the fin and to couple these models to the tube and appropriately account for key variables such as different fin and tube materials. A limited amount of finite element modelling was performed to verify the analytical models.

The analytical model of the fin-on-tube was used to update guidelines for fin tip to fin base temperature difference limits as a function of key variables. The model allows heat transfer designers to understand the stress and life implications of surface selection. Some detailed non-linear finite element modelling was also performed to verify the performance of particular fin-on-tube designs in arduous cycling environments where creep-fatigue interaction may limit the useful life and performance (fin cracking away from the tube resulting in a significant degradation in heat transfer performance).

These analyses examined a high temperature extreme in which the tube and fin were constructed from Inconel 625. The model was also used to study a lower temperature case with ferritic steel fins and tube, but still with a significant fin tip-to-base temperature difference. The analyses for both high and low temperature cases demonstrated that the stresses in the tube and fin can be predicted, including cyclic response, but the lack of information about the strength of the fin-to-tube weld hampered quantitative life estimates. As a result, a test rig was conceived to experimentally determine the strength of the fin-to-tube weld under realistic loading. This rig would provide valuable calibration data that would be applied to estimate the permissible fin tip to base temperature differences in actual heat exchanger applications.

The analytical stress model was used to support development of the rig to test finned tubes under different fin tip to base temperature differences. This included some bounding calculations using the tools developed previously and some simple finite element models. The model was then used to develop a detailed finite element model to simulate the test.

6.2. Stress Test Rig

The objective of this task was to build and operate a finned tube thermal cycling test rig to provide validation data for the engineering criteria on the maximum allowable fin tip to tube temperature differential. This temperature differential results in a cyclic stress that acts to crack the fin to tube weld. Finite element modelling, which was used to establish final test rig specifications, required calibration with experimental results. Test cases with two different tube/fin alloy combinations were subjected to rapid thermal cycling to determine the fatigue limit under three selected temperature differentials. Additional tests were run that included a hold time at maximum temperature to investigate the influence of creep relaxation.

Testing was run using an electrically heated furnace equipped with a motion system to move the test segment into and out of the heated zone on an automatic basis. The test segment was initially air-cooled to temperatures representative of superheat steam tubes in a moving bed heat exchanger to establish the specified fin tip to tube temperature differential. Post exposure metallurgical examination was performed to detect the presence of corrosion, erosion, or oxidation induced degradation.

Figure 6-1 shows an overall view of the stress cold flow test facility at the shakedown stage, while Figure 6-2 shows a close-up view of the test section. Figure 6-3 shows the test section just before it was raised into the oven.

Figure 6-1: Thermal Cycling Rig

Figure 6-2: Thermal Cycling Test Section

Figure 6-3: Thermal Cycling Test Section Entering the Oven

Shakedown Operation

Shakedown operations focused on establishing parameters for automatic cycling of the test assembly. The cycle time was controlled by the heat up rate of the sample once it was raised into the furnace while the cooling was very rapid. Ten-minute cycles were confirmed with further room for optimization. The first test section was instrumented and installed into the apparatus. On heatup, this section initially exhibited improved heat transfer compared with the tube segment used in shakedown and was not able to reach the desired fin tip to tube delta temperature. Heat transfer performance was reviewed to establish an understanding of the difference between test sections and to determine corrective actions.

The shakedown difficulties encountered with the first run in the thermal fatigue test apparatus test run were resolved by elimination of cooling air leakage at the test section gaskets. Data logging software was also updated to permit better monitoring of test status. Typical thermal data obtained during the first test run with air-cooling is shown in Figure 6-4.

Figure 6-4: Typical Thermal Cycling Test Data

Difficulties experienced with the first test section were resolved by upgrading the Type K thermocouples to Type R to overcome burnout and by upgrading the cooling system from air to water to increase the achievable fin tip to tube temperature difference. Additional difficulties resolved included electrical shorting of TC connections that produced a false indication of fin separation. The water-cooling upgrade increased the fin tip to tube temperature differential by almost 100°C more than with air-cooling and exceeded the target specification of 225°C. Typical thermal data with water-cooling is shown in Figure 6-5. The test plan was adjusted to reflect that the fin had not separated from the tube under very severe conditions, reducing the need for some of the planned tests.

Figure 6-5: Typical Thermal Data With Water-Cooling

Testing on the current specimen resulted in accumulation of more than 2,000 thermal cycles, with about 700 cycles with air-cooling and another 1,300 cycles with water-cooling. The results did not show any indication that heat transfer was impaired by the cycling. Analysis of test conditions achieved indicated that fin failure was not expected, in accordance with the observed results. Testing on this specimen was discontinued and a second specimen was installed and cycled to confirm results of this first specimen.

The first test specimen was destructively analysed. Examination revealed no damage to the fin tip and only minor separations and defects in the tube-to-fin weld zone, which were likely artifacts of the original welding process. Figure 6-6 illustrates the conditions found in the first test specimen.

Conditions at fin base (1st location)

Conditions at fin base (2nd location)

Figure 6-6: Condition of the First Test Specimen

Testing of the second finned specimen was completed after the sample accumulated 3,345 thermal cycles with no apparent loss in heat transfer capability. This specimen was also destructively examined and again confirmed the lack of cyclic cracking.

Cyclic thermal exposure testing and finite element analysis began for a test case of a stainless steel fin on carbon steel tube. Exposure testing ended after 3,600 cycles with a fin tip to tube temperature differential of 360°F. While some minor degradation of thermal performance was noted after about 3,000 cycles, the fin did not separate from the tube. Microstructural examination of the original sample (see left photo in Figure 6-7) showed that a notch was typically present at the edge of the fin weld. Some indication was found after exposure that this notch might have lengthened somewhat, as shown in the right photo in Figure 6-7.

Figure 6-7: Condition of Stainless Steel Fin on Carbon Steel Tube

Stress Analysis Model Validation and Results

A finite element analysis model was made of the stainless steel fin on carbon steel tube shown in the left drawing of Figure 6-8. The finned tube was modelled as an axisymmetric body, which was verified with the 3D model. The model was validated with test results from the thermal cycling test rig while using water-cooling of the test sample. The graph in the right side of Figure 6-8 shows that the model temperature predictions closely matched the measured temperatures for both the fin tip and the tube. Points A and B represent the maximum measured temperatures of the fin tip and tube at the end of the heating cycle. The maximum fin-to-base temperature differential was 468°F in this test.

Finned Tube Geometry:

Fin: 304ss

Tube: CS

Figure 6-8: Comparison of Measured and Computed Temperature

Figure 6-9 shows the results of the transient thermal and stress analysis modelling of finned tubes with large tip-to-base temperature differences. Finite element modelling of the test case was able to reproduce the experimental temperature profiles. The modelling showed little cause for concern of failure at the tested temperature cycles.

Figure 6-9: Transient thermal and stress analysis modelling

Finite element analyses of the finned tubes with large tip-to-base temperature differences revealed:

- Low stress occurs at the base of the fin (weld to tube). Yield stress is not exceeded, so plasticity and hence fatigue damage is not expected.
- Although a triaxial stress state exists at the fin base, the normal stress across the fin-to-tube weld is half the yield strength (at a temperature difference of 350°F).
- The local stress singularity of the fin-to-tube corners may result in some localized deformation or cracking (such as that seen in metallurgical sections). The low stress across the majority of the section should ensure integrity of the weld.

- High compressive stresses occur toward the fin tip. This will cause yielding (and cyclic plasticity) at the fin tip but no detrimental effect on the integrity of the fin or tube (perhaps some buckling or waviness may occur in the fin).
- If internal pressure is also considered, there is little effect on the stress distributions (pressure stress is small compared to thermal stress). However, further work is needed to bound creep-fatigue interaction.
- The effect of the spiral fin is negligible. Predicted stresses from 3D and axisymmetric models are similar.

The overall conclusions relative to the question of fin integrity can be summarized as follows:

- The previously defined tip-to-base temperature difference limit of 250°F can be exceeded without concern for fatigue failure (e.g. separation of the fin from the tube).
- Tip-to-base temperature differences of up to 400°F are predicted to be permissible for cyclic operation where long-term creep failure is not a concern (i.e. the tube does not operate in the creep regime).
- Further work is needed to establish a temperature difference bound to ensure that creep-fatigue interaction does not result in premature failure of the fin or the tube (preliminary work indicates that the previously defined bound of 250°F is overly conservative).

6.3. Cold Flow Heat Transfer Tests

This section describes results of tests on a cold flow slice model of the moving bed ash cooler (MBHE) for application to CFB boiler heat exchanger. A series of tests were conducted to determine the heat transfer coefficient of horizontal spiral finned tubing with a bed of ash or particles moving slowly past the tubing. The velocity of the particles ranged from 15-300 fph. This low velocity minimizes erosion of the tubes. This concept can be applied to a moving bed heat exchanger (MBHE) for application to a circulating fluidized bed boiler (CFB).

Background

CFB fluidized bed boilers can use external fluidized bed heat exchangers to extract heat for superheater/reheater steam generation and to control the combustion characteristics. The external heat exchanger has the advantage of flexibility to fuel changes compared to in-combustor surface. The in-combustor heat transfer surface is less expensive, but less adaptable to changes of operating conditions. Also, if the size of the CFB is 300 MW or greater, the in-combustor heat transfer surface may not be made large enough and the external heat exchanger is needed.

A moving bed heat exchanger has been proposed as an alternative to the fluidized bed heat exchanger for the following reasons:

- There is a potential capital cost savings for the MBHE compared to the FBHE
- There is a potential cost savings in operating cost for the MBHE because no fluidizing air with subsequent blowers are required
- There is a potential service cost saving for the MBHE because the low ash velocities will reduce erosion of the tubes and reduce mechanical vibrations

- For certain applications, the elimination of fluidizing air in the MBHE permits optimum chemical process applications
- The MBHE provides more flexibility than the in-combustor heat transfer surface

Objective

The cold flow heat transfer tests had four objectives:

- Determine the effect of tube bundle depth on MBHE heat transfer performance
- Determine the effect of ash or particle composition on performance
- Determine the effect of transverse tube spacing on performance
- Extend the ash velocity range beyond that achieved in previous PPL MBHE tests

Facility description

The test section consisted of a slice model at ambient temperature conditions. A 3 x 10 tube bundle was installed in the slice to simulate the effect of depth on heat transfer performance. Several tube bundle arrangements were evaluated; 6" X 6" staggered, 8" X 6" staggered, 6" X 6" staggered with 8 degree offset, and 6" X 6" staggered with 8 degree inclined. The slice was nominally 4" X 16" in plan area and 8' in height. Inserts were added to adjust the plan area for different tube bundle arrangements. The arrangement of the tube bundles and test section is shown in Figure 6-10.

Figure 6-10: MBHE Cold Flow Tube Arrangements

The circumferentially finned tubes were electrically heated by cartridge heaters in the core of the tube. Three of the tubes had thermocouples attached to the fin tip and base at four circumferential locations as shown in Figure 6-11. These thermocouples were used to

measure heat flow through the fin to the ash. Additional thermocouples were installed in the ash near each instrumented tube to measure ash temperatures at the top, side, and bottom of the instrumented tube.

Figure 6-11: MBHE Cold Flow Tube Temperature Measurement Locations

Two tube/fin arrangements were tested. The first had 1.27" OD T22 tubing with A209 circumferential fins of 0.852" height and 0.102" thickness. The second had 2.00" OD T22 Tubing with A302 circumferential fins of 0.500" height and 0.096" thickness. Both arrangements had circumferential spiral fins on $\frac{1}{2}$ " spacing.

Ash or solids were supplied to the test section from a 1600-2000lb hopper on load cells to provide a measured solids flow rate. The flow rate was controlled by a distribution orifice under the tube bundle. Solids draining from the test section were collected in a lower hopper. At the end of the test, solids were pneumatically transported from the lower hopper to the upper hopper. The solids supply system is shown in Figure 6-12.

Figure 6-12: MBHE Cold Flow Heat Transfer Test Ash Supply System

Measurement System

The basic measurements were made by thermocouples spot-welded to the tip of the fin and to the tube at the base of the fin. Type K thermocouples of teflon insulated #36 gauge wire were spot welded to the fins at the locations shown in Figure 6-11. The thermocouples were then cemented to the fin and lead out of the test section through holes or slots in the tube-fin base. The thermocouples were run in a radial pattern from the spot-weld for a short distance in order to minimize error due to thermal gradients at the hot junction.

In addition to the tube/fin thermocouples, two to four thermocouples were placed in the ash flow stream at each side, above, and below the instrumented tubes to measure ash temperatures. These were complemented by a traversing thermocouple that was positioned

in steps across the test section to determine a transverse temperature profile at the bottom of the test section. Thermocouple leads were connected to a Fluke 2208 data acquisition system. Data from the Fluke was recorded on an Acer notebook PC hard drive for later data reduction and analysis.

Two load cells were connected to the supply hopper to weigh the solids left as the hopper was discharging into the test section. Weight vs. time measurements provided a solids flow through the test section. Test data was recorded by a PC Labview data acquisition system that displayed trends of the data and recorded results to a hard disk for later review.

Results

Tube fin-ash heat transfer coefficients for a 6"x6" staggered tube array are shown in Figure 6-13. There was little difference with bed depth at velocities below 100 fph and a small decrease in heat transfer at the bottom of the tube bundle at velocities of 100-300 fph. There were two fin geometries at the top of the tube bundle and different fin alloys, CF01 and ST01. The data shows that fin geometry had little effect on heat transfer.³

Figure 6-13: MBHE Fin Heat Transfer Test Results for a 6" x 6" Staggered Tube Array

However, the classic circumferential fin heat transfer analysis was done on the data to reduce the results to the surface heat transfer coefficient in each case. It is important to note that solids flowing straight down through the heat exchanger would be touched by a fin or a tube. The following equations fit the data for the top tube row.

³ MBHE Ambient Ash Heat Transfer, Bard Teigen, ALSTOM Power Internal Report PPL-04-CT-10, February 2009

$$h = C_o * (Vel_{max})^{0.3694} \quad (\text{Btu/h/ft}^2/\text{F})$$

Where:

h = the surface heat transfer coefficient on the circumferential fins (Btu/h/ft²/F)

C_o = constant

Vel_{max} = Maximum bulk velocity of ash flowing past the tubes (ft/h)

This equation can be applied to all of the test positions and conditions with the following corrections from Table 6-1.

Table 6-1: Comparison of Measured/Calculated Ratio of Fin Heat Transfer

6X6 array		Avg	Std dev
CFB Ash	Top	1.00	0.04
CFB Ash	Mid	1.01	0.10
CFB Ash	Bot	0.92	0.13
Bauxite 700	Top	1.09	0.06
Bauxite 701	Mid	0.92	0.07
Bauxite 702	Bot	1.37	0.19
Bauxite 500	Top	1.15	0.11
	Mid		
Bauxite 500	Bot	1.38	0.16
8X6 Array			
CFB Ash	Top 8	0.63	0.02
CFB Ash	Mid 8	0.32	0.01
CFB Ash	Bot 8	0.43	0.02

Heat transfer coefficients for a 8"x6" staggered tube array is shown in Figure 6-14, superimposed on the 6"x6" results of Figure 6-13. The 8"x6" results were significantly lower than for the 6"x6" array. This was because some of the ash flowed down between adjacent tubes without being touched by a heat transfer surface. In effect there was little or no mixing or diffusion of the ash particles. In addition, the bed depth showed more deterioration with the 8"x6" array.

Figure 6-14: MBHE Fin Heat Transfer Test Results for an 8" x 6" Staggered Tube Array Comparing Results with a 6" x 6" Array

Figure 6-15 shows the impact of particle size on the heat transfer component, comparing 900 μ bauxite vs 150 μ CFB ash. The bauxite was relatively uniform in particle size and was essentially sintered alumina. The CFB ash was composed of calcium, ash, and silica with a wider size distribution. There was little difference between bauxite and CFB ash except at the bottom of the heat exchanger. This probably was a measurement artifact.

Figure 6-15: MBHE Fin Heat Transfer Test Results for a 6" x 6" Staggered Tube Array Comparing Results with CFB Ash and Bauxite

Measurement of the heat transfer rate depends on the measurement of the ash temperature near the instrumented tube. Thermocouples were installed in the ash stream above, below, and each side of the tube. The top and bottom of the tube always had hotter values than at the side. This is highlighted in Figure 6-16 and Figure 6-17 where the ash temperature was measured by a traversing thermocouple. The temperature profile for the 6x6 array was relatively uniform, while the profile for the 8x8 profile shows significant decreased between tubes. The 8x8 results confirm there was little transverse mixing of the ash.

Figure 6-16: MBHE Fin Heat Transfer Results of Temperature Traverse at Bottom of the 6" x 6" Tube Array Test Section

Figure 6-17: MBHE Fin Heat Transfer Results of Temperature Traverse at Bottom of the 8" x 6" Tube Array Test Section

Results of the 6"x6" tests were compared to hot tests from the PPL lab for CMB applications with bauxite and CFB applications for enriched O₂ firing. The bauxite results in Figure 6-18 showed a more pronounced effect of bed depth that for the cold flow results. This was because there was a flow distribution problem with that particular test series. Subsequent ash control valve/flow distribution tests showed that the solids velocity profile could be improved with proper orificing. This was applied to the CFB results shown in Figure 6-19. In

this case, the top and bottom tube bundle of the moving bed heat exchanger performed close to the expectations from the 6"x6" cold flow tests.

Figure 6-18: MBHE Hot MTF Test Results Compared with Predictions Using Bauxite

Figure 6-19: MBHE Hot MTF Test Results Compared with Predictions using CFB Ash with Varying Fuels, Limestone, and O₂ Levels

Summary

The heat transfer performance of the MBHE was correlated for CFB ash from 15 to 300 fps with minimal deviation from a correlation of:

$$h = C_o * (Vel_{max})^{0.3694} \quad (\text{Btu/h/ft}^2/\text{F})$$

Where:

h = the surface heat transfer coefficient on the circumferential fins (Btu/h/ft²/F)

C_o = Constant

Vel_{max} = Maximum bulk velocity of ash flowing past the tubes (ft/h)

A comparison between this equation and the test data for various tube arrays and solids was given in Table 6-1. The average ratio of measured divided by calculated heat transfer was given for each condition. Also, the standard deviation of the ratio was given. This provided an indication of the effect on performance by each variation.

Moving bed heat exchangers should be designed with tube transverse spacing such that ash flowing downward will touch a fin or tube surface. Fin characteristics should be adjusted to match ash temperature – fin material limits. That is the fin height/thickness/alloy can be

adjusted to keep the fin temperature limits within ALSTOM Power guidelines without affecting the surface heat transfer coefficient between the fin and ash. The moving bed heat exchanger can be applied to a variety of ash or solids. Both bauxite and CFB circulated ash were tested with little difference in heat transfer performance.

7. MECHANICAL EVALUATION

The objective of this task was to identify and address the mechanical design and process issues that need to be resolved in order to develop a commercial concept for the CMB process. Another objective was to also develop some of the component designs needed for the Multi-Use Test Facility and the process information needed for scaled operation of the test loop. A MTF test campaign with two fuels would then be conducted to complete the performance testing of the Proof of Concept program. MBHE design concepts would be evaluated, critical components tested, and the MBHE then installed and tested in a slipstream in a commercial power plant for extended testing.

Alternate mechanical designs and plant arrangements would be evaluated. A new solids control device would be developed that would control solids flow from the MBHE, seal the MBHE from the solids transport air, and split the solids flow into a number of streams for the transport lines. The device would be tested at a bench scale size and then at a larger scale in the Multi-Use Test Facility.

Designs would also be developed for critical components that would be included in the commercial conceptual design. These components would include MBHE tube bundle supports, MBHE support structure, bauxite standpipes, and the solids distributors.

Overview

CFB fluidized bed recirculated ash heat exchangers can be replaced with moving bed heat exchangers. The principal advantage of the MBHE is that it requires no fluidizing air, blowers, or piping/nozzles. A second advantage is that capital cost of the pressure parts will be less expensive than for a fluidized bed heat exchanger. The final advantage is that the MBHE will have little or no vibration of the tube bundle, eliminating some of the bundle clamp problems that have been observed.

The MBHE is proposed as a replacement for the fluidized bed heat exchangers for CFB boilers. The MBHE has potential capital cost savings and operating savings, compared to a fluidized bed heat exchanger. Good MBHE performance requires that the solids flow rate is controllable and maintains a uniform ash flow distribution of the ash flow through the tube bundle. To meet these requirements, a series of tests were conducted in a cold flow slice model of a moving bed heat exchanger (MBHE) to provide design information for ash flow distribution and control. The primary objective was to quantify air flow requirements and physical design requirements of the ash flow control valve (ACV). Other tests were conducted to demonstrate the ability of the system to control and pass rubble, develop an active grid distribution system, and to develop a series of orifice plates to distribute the flow. The results from these tests are described in Sections 7.1 – 7.4. Additional work on the MBHE hydrostatic loading and vent system is reported in Sections 7.5-7.6. Finally, a summary of the MBHE commercial design evaluation is reported in Section 7.7.

7.1. MBHE Ash Flow Control Development

The objective of the Ash Control Valve (ACV) development program was to determine the operating characteristics of a pneumatic ash flow control valve with no moving parts in the ash stream and using low pressure control air from the primary or secondary air fan. A slice model of a commercial MBHE was fabricated to permit the evaluation of pneumatic ash control valves and flow distribution plates. The uniform, controllable flow of ash through a

MBHE tube bundle is essential for maximum performance. In addition, it is possible that some amount of refractory rubble, or ash agglomerates will be present in the recirculated ash, and the MBHE and ACV must accommodate this.

A commercial MBHE sized to replace the fluidized heat exchanger for superheat or reheat service is approximately 16 ft x 16ft in plan area. In addition, the height of the MBHE should be minimized to permit design optimization. The slice model dimensions selected were 8 ft wide by 8 ft high and 4-inch deep. This modelled a commercial MBHE having a center ash inlet and center ash outlet. The 4-inch depth was dictated by the ash flow capacity of the PPL cold flow ash transport facility.

The slice model shown in Figure 7-1 was constructed with Plexiglas walls to permit visual observation of the ash flow. A 2000 lb supply hopper on load cells was located above the slice model to provide a time vs. weight flow rate to the model. A 2000 lb receiving hopper on a digital scale was located below the model to receive the discharged ash from the model. A pneumatic transport system moved ash from the receiving hopper to the supply hopper after a test. The MBHE Slice Model included four 2" Ash Control Valves discharging into a 2" Transport Pipe. It also included perforated distributor plates above the ACV that provided uniform ash flow distribution (they will be further discussed in section 7.4)

Figure 7-1: MBHE Slice Model

Three methods of monitoring the ash flow through the slice model were used during this test:

- Visual observation of the flow patterns in the model
- A hot velocity probe to measure actual ash velocities in the model, giving ash velocity profiles.
- A video imaging system that monitored ash particles at the transparent wall of the model and could be used to give velocity profiles at the wall

A series of nozzles were located under the slice model to connect with ash control valves. The ACV shown in Figure 7-2, was pneumatically actuated by a low-pressure air system. Each ACV was supplied with fluidizing air through a sintered mesh at the bottom or sole, and at the bottom of the inlet dip leg or heel. A house compressed air supply passed through a pressure-reducing valve to provide an air flow at 60" H₂O gage pressure. This air

flow was measured by an ASME corner tap orifice meter. The discharge of the orifice meter was controlled by a parallel series of flow control valves. Several ACVs could be fed by the air supply. The four ACVs with two admission ports each were connected to the air supply. Each port had a flow control valve and an isolation ball valve to control airflow to the port. This permitted the flow to a single port to be set to a desired value and then turned on or off by the isolation ball valve without disturbing the flow setting.

Figure 7-2: Ash Control Valve

Ash Control Valve Description

The primary criteria for developing the ash control valve (ACV) was to eliminate moving parts in the ash flow stream and provide consistent flow with a low head pneumatic air source.

Two air admission ports were needed to provide maximum ash flow rates; a bottom or sole air port, and a dipleg or heel air port. Initially several single openings of various orientation were tried for the air port. However, pressure fluctuations in the discharge pipe caused momentary interruptions of the air flow and backsifting of the ash into the air ports. The back

sifting of the ash then plugged the port and reduced the ACV ash flow rate. The air ports were then replaced with a pad of sintered mesh having minimal pressure drop. The pads were recessed in the bottom and as such did not see the stream of ash flowing through the ACV. In addition, the pad was cooled by transport air and was not exposed to the full temperature of the ash. After installing the mesh pad air ports, backsifting was never a problem during this test campaign.

The ACV discharged into a 2" ID transport pipe and was pneumatically transported to the combustor or in this case a receiving hopper. The receiving hopper was at ambient pressure. However the pneumatic transport provided a backpressure. The backpressure increased as the ash flow increased. The ACV characteristics were such that ash flowed through at a very high rate. This high rate created high backpressure. The high backpressure then reduced the ash flow through the ACV. The system was self-limiting and controlled by the ash bulk density and dipleg height above the ACV inlet. The ash flow and backpressure was not uniform at the very high ash flow rates.

Ash flow control by the ACV was basically on or off. Minor ash flow control could be obtained by adjusting the air flows. However in order to pass rubble through the ACV, the air flows needed to be maintained at high flow settings. Low ash flows were obtained by a digital on-off operation of the valves. If the on-off ratio was adjusted, the average ash flow would correspondingly change. The MBHE tube bundle heat transfer was not affected by this intermittent flow of ash.

The discharge of the ACV dropped into an inclined tube for transmission back to a central chute and to the combustor. Several angles of inclination were tested ranging from -20° downward to 15° upward. Increased downward angles of inclination reduced the required transport pressure while upward angles required increased pressure. The height of the dipleg above the ACV determined how much backpressure the ash could flow against. The height of the diplegs was increased by 10" with a 1.88" ID tube extension into the slice model to provide increased backpressure flow capability.

Multiple ACV were tested, all discharging into a common return transport tube as shown in Figure 7-1. A 3" ACV valve shown in Figure 7-31 was also tested in addition to the 2" multiple ACV for scaleup purposes. The 3" ACV discharged into the 2" transport pipe for expediency.

Measurement System

The basic measurement system is shown in Figure 7-3. The ash flow through the slice model was measured by load cells on the supply hopper. Hopper weight was recorded every second and a weight vs. time calculation gave the flow from the hopper to the slice model. Solids flow from the slice model to the bottom-receiving hopper was weighed by a Cardinal scale and also recorded with time by the data acquisition system. This gave a dual ash flow measurement.

Figure 7-3: MBHE PID Showing Solids Measurement System

Control air for the ACV and transport duct is shown in the PID in Figure 7-4. Low range pressure regulating valves were connected to the house compressed air supply. This provided an adjustable supply pressure for the ash control valves and for the transport duct. The pressure regulators were set for ranges from 60 in H₂O to 120 in H₂O depending on the test requirements. Air flow to the ACV was measured by a corner tap orifice meter having a 0.500 in tube diameter and an orifice diameter of 0.1234 in. Air flow to the transport tube was measured by a diameter tap orifice meter having a 1.94 in pipe diameter and a 1.137 in orifice diameter.

Figure 7-4: MBHE PID Showing ACV Airflow Control and Measurement System

Air flowing from the measurement orifice to the ACV sole and heel ports was controlled by a $\frac{1}{4}$ in pipe needle valve and a $\frac{1}{4}$ in ball valve. The major pressure drop in the air supply system was in the needle valve. The needle valve for each ACV was set with the others isolated by their ball valves so that the orifice meter measured only the flow to that particular ACV. The needle valves were successively set for each port on each ACV and unchanged during a test.

Results

Test data was recorded by a Labview data acquisition program to a PC hard drive. Two systems shown in Figure 7-3 and Figure 7-4 were used; one for the hopper load cells and hot probe and one for DP cells used to measure air flow rates and pressure drops of the ACV control air and discharge duct transport air. Data from the two systems were combined into one common text data file.

Trend plots of the solids flow rate from the supply and discharge hopper load cells were used to calculate load vs. time for solids flow rates in and out of the facility as shown in Figure 7-5. Static and differential pressures of the two air flow orifices were used with ASME

calculation procedures to determine air flowrates for the ACV and transport duct as shown in Figure 7-6 and Figure 7-7. Reviews of these charts show that the pressures were not steady, but had considerable fluctuations. Also, conditions were changed during a test resulting in changes in ash flow rates. The data was averaged over a period of time during which the test conditions did not change. This might represent one or many average data points for a test run. These average values were copied to an average summary data file. Results of the summary data file were then analysed to determine the operating characteristics of the ACV.

Figure 7-5: Typical Solids Supply Measurements and Calculated Solids Flow Rates

Figure 7-6: Typical ACV and Transport Pipe Pressure Measurements

Figure 7-7: Typical ACV and Transport Pipe Pressure Measurements Showing Occasional Pressure Spikes

A pressure drop trace of the transport pipe or static pressure at the ACV shown in Figure 7-7 had a high degree of noise. Data for these tests was recorded every 1-2 seconds. Observation of the ash flowing through the transparent Plexiglas dip leg showed oscillations in flow of this order of frequency. Each test condition was determined by averaging the pressure drop over an extended time period in order to calculate performance criteria. The

averaged values of pressure drop vs. ash flow rate could then be evaluated as shown later in Figure 7-11 through Figure 7-15.

ACV Characteristic Tests

The primary characteristic of the ACV was the control of the air flow rates needed for activation, the subsequent solids flow rate, the influence of the discharge duct inclination, and the effect of rubble or oversize particles on performance. Ash flow rate control by an ACV was primarily an on-off process. Secondary control could be attained by varying the control air flow rates. The primary control method for MBHE application was to pulse the ACV by a digital air flow control, resulting in pulsed solids flow control. This pulsed flow control did not have a significant effect on heat transfer. Results of the digital control method are shown in Figure 7-8 through Figure 7-10. The ACV control air was on continually at the beginning of the test, then was digitally pulsed 10 seconds on 20 seconds off, then was pulsed 5 seconds on 25 seconds off. The solids flows for these conditions were 6500, 2300, and 1100 lb/hr as shown in Figure 7-8.

Figure 7-8: ACV Solids Flow Control Using Digital Pulse Method

Figure 7-9: ACV Control Air Flow using Digital Pulse Method

Figure 7-10: ACV Pressures using Digital Pulse Method

The ACV discharged into an inclined transport duct to a central discharge location. Increasing the downward angle of inclination reduced the transport pressure drop and increased the flow capacity of the ACV/transport pipe. Several angles of inclination from -30° to $+15^{\circ}$ were evaluated. Two sizes of ACV were tested; 2 inch diameter dipleg, and 3-inch diameter dipleg. Both sizes discharged into a 2-inch ID transport pipe. The results of the inclination angle and size are shown in Figure 7-11 through Figure 7-15.

Figure 7-11: Average ACV Transport Duct Pressure Drop vs. Ash Flowrate for Various ACV Sizes and Transport Duct Inclination

Figure 7-12: Average ACV Transport Duct Pressure Drop vs. Ash Flowrate for Various ACV Sizes and Transport Duct Inclination

Figure 7-13: Average ACV Transport Duct Pressure Drop vs. Ash Flowrate for Various ACV Sizes and Transport Duct Inclination

Figure 7-14: Average ACV Transport Duct Pressure Drop vs. Ash Flowrate for Various ACV Sizes and Transport Duct Inclination

Figure 7-15: ACV Ash Flow vs. ACV Control Air Flow

The transport pipe pressure drop as a function of solids flow rate is shown in Figure 7-11. Increased solids flow resulted in an increased transport pressure drop. The transport pressure drop limited the solids flow rate. The dipleg height above the ACV determined the maximum backpressure against which the ACV could transport solids.

Scaleup of the ACV tested in this project to that required for a commercial application needed confirmation. This was accomplished by tests of a 3" ACV for comparison to a 2" ACV. This provided a 2.25:1 scaleup ratio, and generally, the comparison between the two sizes indicated that similarity existed as shown in Figure 7-12 and Figure 7-14 for downward inclined transport pipes. Zero or positive angles of inclination were not tested simultaneously so no comparison could be made for these angles. The solids flux rate of Figure 7-12 was normalized by dividing the solids flow rate by the dipleg cross sectional area. This gave values that could be applied to any size dipleg. An alternative comparison is given in Figure 7-14 by multiplying mass flow of solids by the transport air velocity to arrive at a momentum. All the data for downward inclined pipes fell on the same line for this case. ⁴⁵

The air flow requirements of the discharge duct can be viewed by the solids to air flow ratio as shown in Figure 7-13. This figure sets the transport air flow required to transport a given amount of solids.

⁴ *Effect of Pipe Orientation of Dense-Phase Transport in Inclined Upward and Downward Flows*, Hong & Zhu, Chem. Eng. Comm. 2002, v 189(4), pp489-509

⁵ *Particle size segregation in inclined chute flow of dry cohesionless granular solids*, Savage and Lund, J. Fluid Mech., 1988, v 189, pp. 311-335

The control air flow requirements for actuating the ACV is shown in Figure 7-15. Very little air was required for control. Typical solids to air ratio for the ACV itself was of the order of 400 to 20,000 lb solids/lb air. This translated to 20-150 lb/hr/ft² based on the dipleg area as shown in Figure 7-15. In effect, the solids would flow if enough air was supplied to fluidize the material in the dip leg.

Summary

A pneumatic ash flow control valve can effectively control the flow rate and meet the capacity requirements of a MBHE. The final configuration tested contained sintered mesh control air fluidizing pads. Air injection ports without the sintered mesh pads were subject to ash backsifting and loss of flow capacity. The sintered pads never experienced backsifting or loss of flow.

A solids dip leg above the ash control valve was necessary for a pressure boundary seal. The height requirement of the dip leg was dictated by the ash density and backpressure. Backpressure would be governed by the combustor-cyclone pressure differential and by the discharge duct orientation. Discharge duct downward inclinations of 15-30° provided minimal backpressure. A single transport air supply before the ACV was sufficient to transport the ash to the center duct of the MBHE.

Low pressure primary air could be used for control activation of the ACV, and for transport of the ash from the ACV to the combustor, provided that discharge are sloped downward at an angle of 15-30°. Ash flow control valves could transport up to 400,000 lb/h/ft² of ash with an ash to air ratio of 2000 or greater. Ash flow control rates were governed by a digital on/off time sequence. The ash control valves had no moving parts in the ash stream.

7.2. MBHE Oversized Material Management

Another series of tests was conducted to demonstrate a rubble removal and classification system. The objective of these tests was to demonstrate the ability of the ash control valves to pass rubble and to evaluate a rubble screen system to prevent rubble from collecting in the MBHE tube bundles.

ACV Design to Pass Rubble

The clearances of the ACV were selected to pass a rubble piece of the same size as the valve inlet as shown in Figure 7-16. This required that the ACV not have restrictions in the passageway. Good design practice required that ash flow passages have a minimum diameter of 5 times the maximum particle size. The ACV tested had a minimum diameter of 2.7 times the maximum particle size. The ACV size selected for the slice model had a 2 in inside diameter and 2 in minimum passages throughout the valve. This ACV design operated well and was unaffected by rubble used for the screen and sieve tests. Figure 7-27 shows an example of the oversized material that was passed through the ACV during the active grid tests (Section 7.3).

Figure 7-16: Ash Control Valve Showing The Ability To Pass Rubble

Rubble Classification System

The recirculated ash particle size was relatively small ($D_{50} = 130 \mu$) and flowed easily. The minimum passageway in the MBHE was of the order of 1 inch between adjacent fin tips. This spacing was set to allow oversize ash particles to pass through the tube bundle and exit the ACV at the bottom. However, operating CFB boilers experience a small amount of refractory rubble or ash agglomerates that pass through the cyclone, seal pot, and fluidized bed heat exchanger into the combustor. These can be as large as a brick or dinner plate. Though small in quantity, they could accumulate over time in a MBHE and block ash flow.

Several concepts were evaluated to prevent oversize rubble from entering the MBHE. The most favorable was a shrouded screen in a fluidized region of the seal pot. The kinetic sieve concept also provided a significant reduction in rubble, but did not eliminate it sufficiently. The steam cooled grizzly concept blinded when installed in a packed column and would pass flakes.

The first concept was originally tested during the Phase I CMB MTF test program, which used a perforated cylindrical screen with a surrounding shroud in a fluidized bed as shown

in Figure 7-17. This device worked well with no significant pluggage during the hot MTF tests. The screen was constructed of RA253 alloy and survived operation in a 2000°F fluidized bed.

Figure 7-17: MBHE Perforated Inlet Standpipe With Shroud

A second option was a well screen type grizzly fabricated from water-cooled omega tubes and used a lower alloy content. A device similar to this was installed in the bypass port of the column facility shown in Figure 7-19. The screen was oriented to have a negative slope from the vertical. A shelf above the screen would prevent the rubble from entering the screen from the top. Any rubble entering the screen would have to flow horizontally from the main flow to the bypass. This device kept rubble from the bypass, but was blinded after a short operating time by particles that were wedged into the openings between bars as shown in Figure 7-18.

Figure 7-18: MBHE Grizzly Screen with Trapped Rubble

A 4-inch ID bypass column with a 2-inch ACV (shown in Figure 7-19) was installed with a separate discharge tube to test several methods to prevent oversized material from entering the MBHE. A Johnson Well Screen type of grizzly was installed at the inlet of the bypass column. The purpose of the grizzly was to prevent rubble from entering the bypass column. Oversized gravel, passed through a $\frac{3}{4}$ in screen, was fed to the top of the column along with ash from the supply hopper. Tests were run to evaluate the ability of the grizzly to prevent oversize material from entering the bypass column. Each discharge from the column passed through a separate $\frac{1}{4}$ " screen to permit the measurement of oversize material flowing down each column.

Figure 7-19: Square Column Cold Flow Model

A kinetic sieve concept was also evaluated in several forms to prevent rubble from entering the slice model. The most favorable form is shown in Figure 7-20 and Figure 7-21. The kinetic sieve concept used the gravitational flow through a series of flow disruptions to separate oversize particles from mainstream particles. This concept took advantage of a hopper fill problem wherein oversize particles roll down the slope of the fill cone of a hopper to the outside. This concept had large openings, did not blind, and used gravity as the motive force.

The advantage of this device was that it could be fabricated of refractory and had large passageways. The kinetic sieve made use of gravity segregation of rubble from a flowing stream of fines passing through a torturous passage. This phenomena had been analysed extensively and worked well for particle size differences of the order of $D_{max}/D_{50} < 5$. The application to the MBHE indicated that for $D_{max}/D_{50} > 5$, the segregation would be reversed. Results indicated this to be true. The sieve was fed with a stream of 0.75-inch top size gravel as well as CFB recirculated ash. The discharge of the sieve was split into two streams; one being bypass material and one being ash that was passing through the MBHE. The split of ash flow was 50/50. The rubble split was 90/10. This was a significant reduction

in oversize, but the 10% split passing in to the MBHE stream would still lead to eventual pluggage.

Figure 7-20: Kinetic Sieve for Rubble Separation From Ash

Figure 7-21: Kinetic Sieve Installed on Slice Model

Summary

The performance of the ash control valves demonstrated little or no degradation due to rubble passing through the valves. A series of ash/rubble classification systems including grizzly's and kinetic sieves gave up to 90% rubble reject rates. Kinetic sieves can reduce rubble admission to the MBHE but a fluidized screen in the inlet pipe is required to completely eliminate the rubble. A shrouded screen should be used to control oversize ash agglomerates or refractory rubble from entering the MBHE.

7.3. MBHE Solids Distributor Development

Moving bed heat exchanger (MBHE) heat transfer performance is maximized when there is a uniform solids flow distribution through the heat exchanger. A passive solids distribution grid was originally developed in Phase I to provide a uniform flow distribution of solids through the tube bundle. The passive grid was then installed in the MTF CMB moving bed heat exchanger for testing. The grid operation was erratic and frequently plugged from rubble accumulation. The rubble consisted of refractory pieces and construction debris that was transported to the MBHE from the MTF. The openings of the grid's first stage were increased from 1-1/4" square to 2" square in order to prevent plugging. The enlarged openings only worsened the solids flow distribution.

An active grid flow distributor was subsequently developed to provide uniform solids distribution control and the ability to pass rubble without plugging. The active grid also acts as a flow control device, eliminating the need for a rotary air lock between the MBHE and pneumatic transport system.

Background

PPL has used J-valve systems to control solids flow rates for a long time. This technology was applied to the current problem of controlling solids flow through the CMB MBHE. In addition, digital solids flow control was used in the TVA 160Mw Demonstration Plant fluidized bed boiler ash coolers since 1989 to permit solids as large as 3" diameter to pass through the control valve while being able to control solids flow from zero flow to maximum throughput. The MBHE active grid combines these two concepts.

Active Grid Description

The active grid shown in Figure 7-22 and Figure 7-23 was developed to provide controlled solids flow distribution through the MBHE tube bundle. The MTF CMB MBHE grid consisted of 16 pyramidal hoppers in a 4x4 array with a flow control shoe under each hopper.⁶

Figure 7-22: Active Grid Flow Control Shoe

⁶ *CMB Moving Bed Heat Exchanger Active Grid Development*, Bard Teigen, ALSTOM Power Internal Report PPL-03-CT-20, December 2003

Figure 7-23: Active Grid Cold Flow Development Test Section

The development of the active grid was conducted in a flow model shown in Figure 7-24 and Figure 7-25. A two x two grid array of four pyramidal hoppers was installed in a 16.5" x 16.5" x 8' Plexiglas enclosure. A supply hopper on load cells was used to supply bauxite to the model and to provide a measured bauxite mass flow by a time vs. weight of the load cell.

Figure 7-24: Active Grid Cold Flow Model

Figure 7-25: Active Grid Cold Flow Model Facility

The initial development of the shoe was with only one of the four hoppers installed, while the others were blanked off. After the shoe was developed, all four hoppers were tested to confirm that they would operate independently of neighboring hoppers.

A 16.5" x 16.5" Plexiglas column 8' high was installed on top of the grid. This permitted solid heights of up to 80" above the grid in order to confirm that solids height had no effect on grid performance. Most development tests were conducted with approximately 16" of solids over the grid. Solids were supplied to the column from a storage tank above the column. A hose fed solids from the tank to the column. When the solids in the column rose to a height above the bottom of the hose, flow was controlled by the grid. Solids passing through the grid were collected in a lower receiving tank. After a test, solids were pneumatically transported from the lower to the upper tank.

The last test was conducted with a linear voltage transmitting device (LVDT) attached to the wall of the column to measure the deflection of the wall as solids were added to the column and during operation. This deflection could be used with the wall properties to calculate pressure stresses on the wall of a MBHE needed to size support structures.

Instruments were installed on the solids supply hopper to measure hopper weight, and by a weight vs. time calculation, the flow of solids through the active grid could be measured. Air flow supplied to the hopper shoes was measured by a corner tap orifice meter. Data from these instruments was recorded every 4 seconds by a Fluke 2204A data acquisition system. This data was pasted into an analysis spreadsheet for subsequent analysis.

Shoe Performance Test Results

Solids flow was controlled by actuating the flow control shoe air for 10-15 seconds and then stopping the air flow for 20-30 seconds. The air flow was altered among the shoes of the grid in a prescribed pattern so that each shoe in succession was actuated. The total flow was controlled by the dwell time between activation. Flow control by this method is shown in Figure 7-26 where the dwell time was initially set for a total flow of 4,000 lb/hr, the dwell time was decreased to set the flow to 6,000 lb/hr, and finally, the dwell time was set to zero for a flow of 10,000 lb/hr. Flow control variation between shoes was by varying the on time for air flow activation of particular shoes. A programmable logic controller would be used to set shoe air flow sequences for MTF and commercial applications.

Figure 7-26: Active Grid Performance (6/24/03 data)

The flow control shoe shown in Figure 7-22 had a flow capacity of 10,000 lb/hr of bauxite for an air inlet pressure of approximately 30 psig. Solids flow through the shoe was controlled

by a jet of air introduced at the heel of the shoe. Increasing or decreasing the air flow would increase or decrease the solids flow. This provided an alternative flow control means. Reducing the air flow to the shoe would reduce the ability of the shoe to transport oversize rubble. Therefore, higher air flows to pass rubble, and digital on/off control sequences to control bauxite flow was required.

The shoe dimensions of Figure 7-22 were selected to pass a spherical particle of up to 2" diameter. The shoe was successful in passing rubble as shown in Figure 7-27. The rubble shown was collected below the shoes on a #4 mesh grid. Among the rubble was a short screwdriver shown on the far side of the mesh and some $\frac{3}{4}$ " nuts.

Figure 7-27: Example of Rubble Passed by Active Grid

Test Results – Wall Pressure Due to Bauxite Static Loads

A secondary concern arose during the grid development regarding the effect of the grid on mass flow and wall pressure at the bottom of the heat exchanger. An extensometer was installed on the model wall as shown in Figure 7-24. This measured centerline wall deflection 10.5" above the bottom of the wall. The wall was constructed of $\frac{1}{2}$ " thick Plexiglas. The walls were attached to each other at the corners. The bottom and top of the walls were not restrained and used a taped slip fit to prevent solids leakage. The extensometer was calibrated and set to a position of 0.273" shown in Figure 7-28.

Figure 7-28: MBHE Wall Pressure Profile/Wall Deflection (7/3/03 data)

The model was then filled with 1750 lb of bauxite to a height of 80.5" above the bottom of the wall. This increased the position to 0.295" for a net deflection of 0.022". The control air to the shoes in the grid was actuated at 50,200 seconds and produced positional variations of 0.298-0.305". These variations were due to shoe air pressure influences. After some time period, the wall deflection decreased until the model was empty at 51,600 seconds. The wall position did not return to the original value because bauxite was trapped between the wall and 4" grid elevation above the wall bottom.

Active Grid MTF Testing

A 4 x 4 array of 16 active grid devices was installed in the PPL MTF CMB moving bed heat exchanger (MBHE) as shown in Figure 7-29. A loop seal was installed below the grid to prevent solids transport air from flowing up through the shoes.

Figure 7-29: Active Grid Installation in PPL MTF CMB Moving Bed Heat Exchanger

After several cold flow tests with the MBHE access doors open, it was determined that a vent was required for the actuation air and transport air leaking upward from the transport discharge pipe. A shielded 1-1/2" vent pipe was installed in the lower MBHE hopper drain. The discharge of this vent was directed to the top of the MBHE and then to the discharge pipe. This vent was undersized and the design was prone to plugging. Cycling of the shoe transport air could be detected by ash puffs from the seams of the MBHE during operation. Also after a period of operation, the upward permeation of transport air reacted with carbon particles in the bauxite, causing them to sinter. The sintering caused channeling of the solids flow and degradation of the heat transfer performance.

The vent pipe was reconfigured to a double 2" pipe rake across the top of the shoes, and a double connecting standpipe from under the shoes to the rakes as shown in Figure 7-30. This vent system bled off all shoe activation air and lower seal pot transport air. No dust leakage was evident during operation, no sintering was found after operation, and the heat transfer performance of the MBHE did not deteriorate during operation.

Figure 7-30: Improved Active Grid Vent System

Thermocouples were installed over each shoe of the MBHE active grid to indicate whether each had uniform flow. Low flows would give low solids outlet temperatures, and high flows would give high temperatures. The programmable control logic supplying solenoid valve controlled air pulses to each shoe could be programmed to change the temperature distribution of the solids. Also, the dwell time between air pulses could be used to control the solids flow rate.

The MTF test results confirmed that the active grid could control the flow of 700 μ bauxite solids and pass rubble of the order of 1-1/2". Flow could be controlled by a digital air pulse method whereby each shoe was activated for 10 seconds, and a delay time between activation pulses was varied to provide a desired average flow. The active grid should be used as a flow control and distribution device on moving bed heat exchangers.

7.4. MBHE Distribution Orifice Plate Development

Preliminary layouts of the MBHE indicate that ash flow distribution and control are critical to the design. The original MBHE designs used mass flow hoppers with 70° angles to ensure uniform solids flow throughout the MBHE. This approach required a very tall MBHE or an excessive number of hoppers at the MBHE bottom. A series of orifice plates was therefore developed to reduce the height requirements. The multiple orifice plate design was simple and required a 35-40° slope with less elevation requirement, resulting in a 70-75% height reduction of the distributor. The orifice plates could be made of either high alloy steel or refractory tiles.

A series of orifice plates was an effective method of distributing the ash flow uniformly throughout the MBHE. There would be a selected number of ACV in the bottom of the MBHE. Each ACV could be considered the final series of the orifice plates with openings equal to the number of ACV. The next orifice plate would have 4 times the number of orifices. The next orifice plate would have 16 times the number of orifices, etc. Vertical spacing between orifice plates was governed by the angle of repose of the ash and the spacing between orifices on the plate above. Height requirements governed by the angle of repose were far less than those governed by the angle of friction. The angle of friction would govern the flow distribution without the orifice plates.⁷

Operation of the slice model without orifice distribution plates showed that ash flow plumes extended from the top of the ash column down to the ACV with little spreading of the plume. Installing two orifice plates above the ACV provided good distribution of ash flow throughout the slice model. The openings of the orifice plates should be 2.7 times the maximum particle size.

Two orificed distribution plates were installed above the ACV inlets to provide a uniform ash flow distribution through the tube bundle, as shown in Figure 7-31. The initial design criteria called for keeping the ash flow cross sectional area constant for each of the plates. In addition, the number of holes in successive plates increased with elevation. Four holes in a plate fed one hole in the plate below. The two dimensional slice model criteria called for two holes feeding one hole in the plate below. The four slice model ACV constituted four orifices at the bottom, the first orifice plate had eight openings and the top had 16 openings.⁸

⁷ *Fluidization and Fluid-Particle Systems*, Zenz & Othmer, Reinhold, NY, 1960, pp 154

⁸ *MBHE Slice Model Tests – Solids Flow Distribution*, Bard Teigen, ALSTOM Power Internal Report PPL-04-CT-10, January 2006

Figure 7-31: Slice MBHE Model with Orificed Distribution Plates

The concept of the orificed distribution plates was then transferred to a 3 dimensional 16.5 inch x 16.5 inch column with two distribution stages as shown in Figure 7-32. This column had the same ash flow cross sectional area as the slice model but wall effects were minimized. Traverse ports were installed in the model to permit traverses of the ash velocity profile at the center, orifice position, and wall position as well as from front to back in reduced increments. The 3 in ACV scaleup valve was installed in the center of the column for ash flow control.

Figure 7-32: MBHE Cascade Orifice Plate Ash Flow Distributor

Four ACVs were installed in the slice model with the hope that the ash would distribute itself uniformly at some level above the inlet of the ACV representing an angle of friction of 70° . The 70° angle of friction exists for a short distance above the ACV inlet, then a solids plume extends upward to the top as shown in the photo of Figure 7-33. A dead volume of ash exists between the plumes.

Figure 7-33: MBHE Slice Model Without Solids Distribution Plates

A limited number of ACVs will be used in a commercial MBHE. This requires that the ash flow above the ACV and through the tube bundle be uniform. A series of orifice plates, one above the other will accomplish this. Holes in a plate are fed by 4 holes from the plate above. The angle of repose governs solids flow between the holes in vertical adjacent plates. This angle is less than half of that required for mass flow in hopper bottoms and provides the minimum height requirement of several concepts considered. This concept is shown in Figure 7-32 for the one-dimensional slice model.

Ideally, the distributor holes would increase in size as the solids travel downward through the plates, with the open area of each plate less than that of the plate above. However, good design practice with ash particle flow requires a flow path minimum diameter of 5 times the maximum particle size. This lead to small openings in the first plate when several layers are used. The small openings plugged when oversize rubble was fed with the ash. The holes were enlarged to have the same size for all plates and still provided relatively uniform flow as shown in Figure 7-34.

Figure 7-34: MBHE Slice Model With Solids Distributor Plates

Performance of the distributor plate was quantified by a video traverse technique that measured the velocity of individual ash particles as they slid down the surface of the Plexiglas wall of the slice model. Figure 7-35 shows the results of such a test when oversize rubble was being fed simultaneously with the ash.

Figure 7-35: Ash Flow Distribution Above Orificed Distributing Plate in Slice Model

Validation of the distributor plate was extended to a two-dimensional application in a 16.5 in x 16.5 in column with a single ACV in the center as shown previously in Figure 7-19. A single four-orifice distributor plate above the ACV was used to distribute the flow. A hot ash velocity probe was used to determine the velocity distribution in the column. Two elevations of traverse ports were installed with ports located at the center, over the orifice holes, and at the half distance between the orifices and the wall. Results of the velocity distribution are given in Figure 7-36. Traverse elevations at 7 in above the orifice plate showed high velocity spikes above the orifices. Traverse elevations at 18 in above the orifice plate showed relatively uniform flow at all regions of the model.

Figure 7-36: Ash Flow Distribution Above Orificed Distributing Plate in Column Model

Summary

A series of multiple orifices provided a relatively uniform ash flow with a height requirement dictated by the ash angle of repose. Multiple orifices with openings of 5 x the maximum rubble particle size should be used to distribute flow in the MBHE. A cascade series of orificed plates having successively fewer holes should be used to provide uniform ash flow through the MBHE. The ash cross sectional flow area in each plate does not have to be uniform. The orifice diameters of the top plates can be opened to permit rubble to pass.

7.5. MBHE Support System – Tube Support Hanger Loading

Background

The moving bed heat exchanger (MBHE) consists of a series of tube bundles with ash flowing through to heat steam or water flowing in the tubes of the bundle. The tubes must be supported to carry their weight plus that of the ash surrounding the tubes. The support is complicated because some of the ash weight is transferred through the tube bundle to the heat exchanger floor below the tube bundle. The design of the supports depends on the load imposed on the tube bundle elements by the ash as well as the dead weight of the tube bundle. A series of tests were therefore conducted in the PPL MBHE cold flow column model to evaluate the load imposed on a simulated tube bundle by ash. The test results were used to develop an equation describing a MBHE tube bundle support load as a function of the bundle configuration and ash loading. Key effects evaluated during the testing included the impact of ash flow on tube support loads, the impact of ash depth effect above the tube bundle, and the impact of the number of tubes in the tube bundle.

Facility description

The test facility shown in Figure 7-37 was used to determine the ash load on a suspended tube bundle. Ash was supplied to the test column from a supply hopper above the column through a 4" hose that ended a fixed distance above the tube bundle. An ash control valve at the bottom of the column controlled the flow of ash through the column and discharged it to a collecting hopper below the column. Load cells at the top and bottom hopper provided a weight vs. time record of ash flowing through the column. A simulated tube bundle was mounted on two load cells to measure the load on the tube bundle by ash flowing down the column. Successive layers of tube bundle could be removed to determine the effect of bundle depth on the load. Introduction of ash at different elevations determined the effect of ash level above the bottom of the tube bundle on hanger load.

Figure 7-37: MBHE Column Model

The basic measurement system was connected to a Labview based PC system. In addition to the hopper load cells, solids flow from the column model to the bottom receiving hopper was weighed by a Cardinal scale and also recorded with time by the data acquisition system. This gave a dual ash flow measurement. A single 3" pneumatic ash control valve

(ACV) set the ash flow through the model. Control air for the ACV permitted the ash flow to be varied to a desired set of conditions.

The simulated tube bundle was hung from two load cells in the column model. The tube bundle was configured as a staggered array of four parallel 2" OD followed by a second array of 3 parallel 2" OD tubes. A second set of tubes could be arranged to provide either 14, 7, or 4 tubes in total. The tubes were 14" in length with a 1" wide angle on each end to tie to the tubes in place. The complete tube bundle arrangement is shown in Figure 7-37. The ash supply hose from the top ash hopper could be varied in length to change the ash height above the tube bundle.

Results

Three variables were measured for the ash loading on the tube bundle.

- Load vs. Number of Tubes
- Load vs. Ash height above the tube bundle
- Load vs. Ash Velocity

The tube bundle initially consisted of four levels of tubes as shown in Figure 7-37. The bottom two levels of tubes were removed after testing the four level configurations. Next the third level was removed leaving a single layer of four tubes. This provided a tube bundle depth effect on ash loading.

A series of ash depths and velocities were tested for each bundle configuration. The ash loadings for all of the tests are plotted in Figure 7-38 and Figure 7-39.⁹ The ash loading can be described by the equation:

$$W_h = C_1 * (W_{ash})^{0.333} * (\text{No Tubes})^{0.275} \pm 2.1\%$$

Where:

W_h = Ash loading on the tube bundle (lb/ft of tubing)

W_{ash} = Weight of ash above the bottom of the tube bundle (lb/ft²)

No Tubes = Total number of tubes in the tube bundle at all elevations

C_1 = Ash loading constant

The effect of ash velocity flowing through the tube bundle was negligible as shown in Figure 7-39.

⁹ MBHE Tube Support Hanger Load, ALSTOM Power Internal Report PPL-10-CT-11, January 2010

Figure 7-38: MBHE Tube Bundle Ash Loading on Support - Hanger Load vs. Ash Height Above the Bottom of the Tube Bundle

Figure 7-39: MBHE Tube Bundle Ash Loading on Support - Effect of Ash Velocity

Conclusions

The MBHE tube support loading can be described by a simple equation with ash weight above the tube bundle, bundle weight, and number of tubes in the bundle. The results showed that the ash velocity does not affect the hanger load, the ash depth has a low exponential power effect on hanger load, and the number of vertical tubes in the bundle has a low exponential power effect on hanger load. The hanger load can be expressed mathematically by the ash depth and number of tubes to within +/- 2%.

7.6. MBHE Hydrostatic Loading

The support structure for portions of the CMB must be designed to carry the load of the bauxite solids. To determine the effective bauxite solids loading, a simple finite element model was created and analysed to simulate the Plexiglas model test described in Section 7.5. The predicted and actual deflection at a point on the wall of the model was compared to estimate the effective hydrostatic loading of the bauxite solids. The results indicate that the effective hydrostatic pressure was approximately 0.13 times the theoretical hydrostatic head.

Introduction

In the construction of the circulating moving bed (CMB) heat exchanger, it is necessary to ensure that the walls and associated structure can contain the load imposed by the bauxite solids. To provide some data on the load imposed by the bauxite solids, a Plexiglas model test was performed in which the deflection of the wall was recorded for a particular bauxite solids loading. Figure 7-40 shows a schematic diagram of the active grid cold flow model with the detail view showing the support for the Plexiglas enclosure and location of the extensometer. This section presents an analysis of this model test to determine the effective hydrostatic pressure (head) of the bauxite solids. This value can then be used for future designs of walls and supports.

Figure 7-40: Schematic Diagram of Active Grid Cold Flow Model

Analysis Methodology and Description

The basic physical model enclosure for the 700 μ bauxite solids is 8ft high with 16.525in x 16.525in (internal dimension) walls of 0.5in thick Plexiglas (Figure 7-40 and Figure 7-41). For the test to determine the wall deflection, the enclosure was filled with 1750lb of bauxite solids to a height of 80.5in. Hence, the volume of bauxite solids was $16.525 \times 16.525 \times 80.5 = 21982.6 \text{ in}^3$. The effective density of the bauxite solids was, therefore, 0.0796 lb/in^3 . The peak (theoretical) hydrostatic pressure at the near base of the wall (elevation $\sim 6.4 \text{ in}$, where the bauxite solids hoppers shield the wall) was 5.93 lb/in^2 (Figure 3).¹⁰

¹⁰ CMB Moving Bed Heat Exchanger Active Grid Development, Bard Teigen, ALSTOM Power Internal Report PPL-03-CT-20, December 2003

Figure 7-41: Isometric view of Plexiglas Enclosure With Key Dimensions

Due to the inherent symmetry, only a quadrant of the enclosure was modelled (Figure 7-42) with shell elements. Symmetry boundary conditions were applied to the vertical (cut) edges of the wall. The base of the wall was analysed with two different boundary conditions but in each case it was supported vertically. For Case A, the base of the wall was assumed to be free to slide and rotate, thereby representing support by a frictionless, rigid, surface. For Case B, the base of the wall was assumed to “stick” and no lateral motion was permitted (rotation was allowed), simulating a very high friction coefficient.

Figure 7-42: Schematic Diagram Showing Surface (Shell) Model, Pressure Load Distribution on the Walls and Boundary Conditions

The Plexiglas was assumed to have an elastic modulus of 450,000psi based on typical data obtained from www.matweb.com.

A simple linear elastic analysis was performed to determine the deflection of the wall, subject to the hydrostatic pressure. The ABAQUS general-purpose finite element code (version 6.4-1) was used for the analysis.

Results

The calculated deflection of the walls of the model is shown in Figure 7-43. The results from Case A (frictionless) and Case B (high friction) were very similar when displacement was compared at the extensometer location; both models predicted a deflection of 0.17in. This indicated that the location measurement point and inherent flexibility of the model were sufficient to ensure that the precise nature of the boundary condition at the base did not have a significant effect on the results. The computed displacement of 0.17in at the

extensometer location was approximately 7.7 times the measured deflection of 0.022in. Therefore, the effective hydrostatic pressure of the bauxite solids was approximately 0.13 times the theoretical hydrostatic loading.

Figure 7-43: Color Contour Plot Showing Displacement Magnitude (in)

For comparison, an analysis was also run in which the hydrostatic pressure of the bauxite solids was assumed to act from the base of the Plexiglas model to the full solids height (80.5in). This simulated a poor seal between the Plexiglas wall and the hoppers. The predicted wall deflection at the extensometer location was 0.22in. This was slightly larger than for the other analyses due to the higher solids loading.

Conclusion

The calculated deflection of the wall of the Plexiglas model for the theoretical hydrostatic head of the bauxite solids was approximately 7.7 times the measured deflection. Therefore, the effective hydrostatic pressure of the bauxite solids was approximately 0.13 times the theoretical value.

Recommendations

The theoretical hydrostatic pressure should be multiplied by a factor of 0.13 to estimate the bauxite solids loading on support structures.

Additional physical testing, and modelling, should be performed to confirm that 0.13 times the theoretical hydrostatic pressure is appropriate for estimates of bauxite solids loading. Such testing should evaluate alternate geometries (e.g. hoppers) and measurements of structural deflections should be taken at multiple locations.

7.7. MBHE Vent System

A vent system was installed in the MTF CMB Moving Bed Heat Exchanger. The vent was necessary to direct transport air from the active solids distribution grid to the vent at the top of the heat exchanger. The vent consisted of 2 - 2" standpipes connecting the plenum below the active grid to two rakes above the grid. The rakes above the grid consisted of 2 - 2" pipes with a 1/2" wide slit on the bottom to withdraw air from above the grid. These rakes also had a shrouded opening to receive the end of the 2 standpipes. The rakes were inclined 10° from horizontal to minimize solids pickup. The rakes exited the door of the heat exchanger, turned upward and re-entered the heat exchanger at the top fill port. The vent system is shown in Figure 7-44.

Figure 7-44: Moving Bed Heat Exchanger Active Grid Vent System

A shroud was also constructed and installed around the CMB Combustor bauxite overflow standpipe to prevent coal particles floating on the top of the bauxite bed from entering the grizzly on the standpipe. The shroud was open at the bottom to let bauxite flow upward to the top of the standpipe. The shroud was open at the top to vent fluidizing air carrying bauxite to the overflow pipe. A cover plate prevented bauxite and coal particles from falling into the vent pipe from the top of the combustor. The shroud was sufficiently high to extend beyond the top of the bauxite bed in the combustor. The shroud is shown in Figure 7-45.

Figure 7-45: CMB Combustor Standpipe Shroud

8. PROCESS EVALUATION

The objective of this task was to evaluate alternate techniques to enhance process performance. Sulfur capture methodologies and NO_x control options would be evaluated. Combustion optimization, alternate fuel firing, and mitigation of bauxite coating growth would also be examined. Alternate heat transfer solids other than bauxite would be investigated and promising candidates would be tested in the laboratory.

8.1. Agglomeration Potential in a MBHE

Summary

The MBHE is being developed as a low cost replacement of a CFB FBHE. It also has application as an enabling technology for CFBs with ultrasupercritical steam conditions, chemical looping, and O₂-fired CFB. Although the MBHE performance has been demonstrated through various cold flow facilities and two hot pilot scale MBHEs, agglomeration and scalability (impacting solids flow distribution) have been identified as a potential issues that could impede MBHE commercialization. This document addresses the potential for agglomeration issues in the MBHE. Scalability issues are currently being addressed in the cold flow MBHE slice model.

Early operation in PPL's two pilot scale MBHEs was hampered by solids flow disruptions resulting from several causes. Most blockages were attributed to rubble built up at the MBHE discharge from refractory spalling, rock accumulation, or wet ash buildup from improperly dried refractory. The MBHE design was successfully modified to prevent these blockages in future operation. The MBHE operation during the recent O₂-fired CFB MTF test in July 2005 had smooth uninterrupted solids flow throughout the duration of the test campaign. The following items summarize some of the conclusions from this report.¹¹

1. Rubble accumulation at the MBHE discharge. Rubble sources included refractory that spalled off the MBHE refractory walls and MBHE cyclone target zone, and rocks and debris that were transported to the CMB MBHE from the combustor. The CFB MBHE also had some accumulation of refractory that had spalled off of the refractory walls. This problem was addressed by refractory improvements and modifying the MBHE discharge to handle larger size material. A commercial MBHE design will need a classification device at the MBHE inlet to prevent oversize particles from entering the bundle. The MBHE cold flow model modified for a classification test at the MBHE solids inlet (see Section 7.2).
2. Wet ash buildup from improper refractory dryout. The refractory in the original CFB MBHE was not dried out properly. This resulted in an accumulation of wet ash clumps in the MBHE that had to be cleared out. This problem was eliminated by properly curing the refractory.
3. Ash agglomeration from overheating. There was one instance of ash agglomeration due to overheating in the CMB MBHE. This resulted from a plugged vent underneath the MBHE that allowed grid activation air to pass through the MBHE. The problem was exacerbated by the pneumatic transport of ash and burning coal particles from

¹¹ *Agglomeration Potential in a MBHE*, Glen Jukkola, ALSTOM Power Internal Report, January 2006

the combustor to the top of the MBHE. The MBHE vent was modified for subsequent operation and this problem was eliminated. Note that the vent was required for the CMB MBHE because of the seal underneath the MBHE going to the pneumatic transport line that carried solids back to the top of the combustor. This seal is not necessary on the CFB MBHE design and a vent is not required there. The TVA bed ash cooler design also used a moving bed principle as solids flow through axial-multi-tube coolers. These coolers have operated over 20 years without any significant instances of agglomeration or sintering.

4. Agglomeration experience in BFB boilers. A concern was raised about the agglomeration that was observed in early bubbling fluidized bed boilers. ALSTOM Power has experience with bubbling bed boilers from the Great Lakes Demonstration Plant (50,000 lb/hr steam boiler), the TVA 20Mw BFB Pilot Plant, and the TVA 160Mw Demonstration Plant. All of these boilers had multiple bed compartments and operated with a bed slumping strategy for turndown. None of these BFB boilers experienced any bed agglomeration due to air leakage through the bed. Early BFB operation did have some operational issues with the top of bed in the slumped compartments crusting over. This was a result of burning coal particles being thrown from the active bubbling bed onto the slumped bed. The burning particles created local hot spots that resulted in minor agglomerates forming on the bed surface. This problem was eliminated by occasionally “fluffing” the slumped zone to break up the surface crusting.
5. Carbon burnout predictions. A simple carbon burnout model was developed to predict the solids temperature rise that would occur in the MBHE if all of the air permeating through the MBHE were consumed by the available carbon. The model showed that the lack of available air permeating through the MBHE limits the potential combustion and that local solids temperature should not increase significantly due to carbon combustion in the MBHE.

The following sections provide further details about the CFB and CMB MBHE designs and operating experience, TVA’s bed ash cooler experience, and more details about the potential impact of carbon burnout in the MBHE.

CFB MBHE Description and Operating Experience

A moving bed heat exchanger (MBHE) was placed in the overflow line of the cyclone to combustor return line of the PPL MTF. This permitted the evaluation of the MBHE concept for application to a CFB or to a CMB boiler. Material flowing through the MBHE was CFB recirculated ash rather than bauxite proposed for CMB operation. The MBHE is shown in Figure 8-1.

Figure 8-1: CFB MBHE Design

Ash entered the top port of the MBHE, passed through the tube bundles, through a distribution grid, into a return duct, through a rotary valve, and into the bottom of the combustor. Initial operation was satisfactory, although an accumulation of wet ash at the

end of the tube bundle occurred from moisture released from incompletely cured refractory. Subsequent operation was improved once the wet ash clumps were removed. Sufficient operating data was gathered to develop a heat transfer model for MBHE design.

Early slice flow models of a moving bed heat exchanger indicated that a pyramidal grid would provide a uniform flow distribution through the tube bundle. This was used on subsequent MBHE test facilities. However, the openings in the grid must be sized to the solids flow. Initial opening sizes in the CFB MBHE were for more solids flows than could be achieved in the MTF. This gave less than desired flow distributions. Also, reduced loads further aggravated the distribution. The current MBHE design avoids this problem by operating the solids control valves for a design flow rate. The flow rate through the valve is maintained constant while the valve is in operation. Changes in solids flow rate are accomplished by cycling the control valves on and off. This principal has been demonstrated on the MTF MBHE and works well. This principal has also been proved with the TVA bed drain coolers, as discussed below.

Evaluation of the MBHE heat transfer performance showed that stratification or platinization of the solids between adjacent vertical tubes shaded part of the tube-fin surface. A shading factor was determined that matched predicted and measured heat transfer. This factor was used for the MBHE commercial design that showed significant cost savings compared to a conventional FBHE. Subsequent cold flow heat transfer tests with an electrically heated tube bundle showed that spacing the tubes more closely could reduce the platen effect.

The MBHE installed in the MTF ash return line was rebuilt with a tightened tube bundle as suggested by the cold flow tests. Operation of the compact MBHE was initially erratic because improperly cured refractory spalled off and plugged the distributor grid openings. Finally, rework of the refractory permitted operation of the MBHE without significant problems. The MBHE was again operated in June 2005 during the O₂-fired CFB tests after the previous refractory issues had been addressed. The MBHE operated smoothly and reliably during this test campaign without any flow disruptions. Visual inspection inside the MBHE after the campaign did not show any evidence of refractory spalling, wet ash buildup, nor any agglomeration. Heat transfer was improved over the original tube spacing.

CMB MBHE Description and Operating Experience

A second moving bed heat exchanger (MBHE) was installed in the PPL MTF to evaluate heat transfer performance during a CMB mode of operation. Bauxite was heated in the combustor of the MTF and pneumatically transported to the top of the moving bed heat exchanger located to the rear of the MTF as shown in Figure 8-2. This MBHE operated at hotter temperatures than the CFB MBHE as the ash originated from the 2000°F CMB bubbling bed. The solids inlet temperature after pneumatic transport was about 1800°F for the MBHE compared to about 1600°F for the CFB MBHE.

Figure 8-2: MBHE Installed in the PPL MTF for CMB Operation

A cyclone at the top of the moving bed heat exchanger separated the bauxite from the transport air. The bauxite dropped on top of the tube bundles and the transport air was vented to the atmosphere. The solids level was maintained 2-3 feet above the tube bundles. The solids level was monitored by a viewport located on the top of the MBHE having two inclined lasers focused on the top of the solids surface, and a video camera to record the distance between the laser intersections of the surface. The video camera also provided a view of solids at the top of the MBHE. This video camera showed that large burning coal particles (still with volatile burning plumes) were frequently transported to the top of the MBHE, as show in Figure 8-3. Despite the introduction of burning coal particles combined with the presence of transport air in the region just above the MBHE, there was only one instance of agglomeration in the MBHE that resulted from a vent design issue. Modifications to the vent design eliminated this problem.

Figure 8-3: MBHE Laser Level Indicator

The MBHE included four water-cooled tube bundles consisting of spiral-finned tubes arranged in an inline-offset clearance. Measurements of the water flow to each bundle, inlet-outlet temperature difference determined the heat flow from the solids to the bundle. Measurement of the solids inlet temperature, below the top surface, provided a solids inlet temperature. Additional thermocouples immersed in the solids below the bottom tube bundle provided the solids outlet temperature. The 4x4 grid of 16 thermocouples at the bottom was also used as an indication of uniform distribution of the solids discharging from bottom of the MBHE. A heat balance between the cooling water heat pickup, and solids temperature difference was used to calculate the solids flow rate.

Control of the solids flow through the MBHE was by a device at the bottom of the heat exchanger, ensuring that it remained full of solids and operated in the packed bed mode as shown in Figure 8-4. Initially, a variable speed rotary valve was used to control the solids flow. A 4x4 passive pyramidal grid above the rotary valve was used to provide a uniform flow distribution at the bottom of the MBHE. The openings in the grid were sized to distribute the flow evenly. However, flow through the MBHE was impeded by an accumulation of refractory rubble and rocks from the coal fed to the combustor that plugged the grid openings. A perforated screen located above the grid was of little help in reducing the plugging problem.

Figure 8-4: CMB MBHE Installed in the PPL MTF

An active grid was therefore developed with large openings to prevent the plugging problem. Each opening of the 4x4 pyramidal grid had a pneumatic flow control valve under the opening. The openings in the grid were enlarged to a 2"x2" size and rubble of up to that size could pass through the grid. Air to actuate the flow control valve was connected to a solenoid bank. Air was sequentially supplied to each flow control valve for a short period, followed by a rest time. Varying the difference between operating and rest time controlled the flow through the valve. This timing was controlled by a computer solenoid control system and was manually controlled to provide uniform temperature profiles at the bottom of the MBHE. The intermittent flow through the grid did not significantly affect the MBHE heat transfer performance.

Solids exiting from the bottom of the MBHE passed through a pigtail loop seal, into a pneumatic transport pipe, back to the top of the combustor. Backpressure from the transport pipe was sealed by the loop seal. However, the grid activation air needed to be vented. Initially, an under grid vent was used, but was prone to pluggage by solids. Unreliable vent operation resulted in nonuniform solids flow distribution and in one case grid activation air permeating up through the bed caused sintering of coal particles transported in the MBHE.

A revised vent consisting of two rakes over the active grid, and two standpipes passing through the grid enabled the grid activation air to be vented to the top of the MBHE and out the transport air vent. This rake/standpipe modification eliminated upflow of air from the grid to the tube bundle. The MBHE was operated for 3 additional days with the combustor firing of Pittsburgh 8 coal and with Pet Coke with no deterioration in performance.

Description of TVA Bed Ash Coolers

The initial concept of the moving bed heat exchanger was implemented on the TVA 20Mw pilot unit and the TVA 160Mw Demonstration Plant boiler. The spent bed ash coolers consisted of axial-multi-tube coolers. The coolers were installed under the bed in a vertical orientation. Spent bed material entered the top of the cooler, flowed down 9x4" ID tubes, through a flow control valve, and into the pneumatic spent bed material transport pipe. The TVA Demonstration Plant boiler operated with Kentucky #9 coal, pet coke, and other low grades of fuel during the 20 years period of operation. The coolers have been in operation during all of this time. Figure 8-5 shows the bed drain cooler that was installed at the TVA 160Mw Demonstration Plant.

Figure 8-5: Bed Drain Coolers at TVA 160Mw Demonstration Plant

There were several relevant lessons from this application. First, the coolers have been in operation for 20 years with no major operational issues. There were some mechanical design issues as discussed below, but this had nothing to do with the moving bed aspects of this design. Second, performance analysis of the cooler heat balance indicated that there was carbon burnout in the cooler. Despite the evidence of carbon burnout, there have been

no reported instances of agglomeration or sintering in the TVA bed ash coolers over the 20-year operating period. Finally, control of the bed material through the cooler can be of cyclic mode without affecting heat transfer or cooler integrity.

These coolers have had several mechanical design problems. A low temperature weld was used in a high temperature region, leading to weld failure. The tubes were re-welded to the tube sheet, using a high temperature weld technique to eliminate this problem. The coolers experienced initial plugging that was caused by cooling of the tubes during shutdown, causing the bed material to be compacted to a non-flowing mass. This was alleviated by periodically letting a small amount of bed material flow through the cooler during shutdown.

Initially, a knife-gate valve at the bottom of the cooler adjusted to provide a desired flow control of the bed material. This valve was plugged at low flows by oversize bed material. A dump valve was then installed at the bottom of the cooler, and periodically completely opened for a short time, then closed. The flow control was achieved by varying the open/shut time of the valve. This mode of operation did not affect heat transfer and did not affect the heat exchanger integrity.

Carbon Burnout Calculations

A simple carbon burnout model was developed to predict the solids temperature rise that would occur in the MBHE if all of the air permeating through the MBHE were consumed by the available carbon. The model showed that the small quantity of available air permeating through the MBHE limited the potential combustion and that local solids temperature should not increase significantly due to carbon combustion in the MBHE.

The model assumed that the ash flowing through the MBHE contained 1% carbon by weight. The airflow rate through the MBHE was based on the Rose & Rizk correlation for fixed bed gas permeation pressure drop and permeability tests conducted at PPL. Figure 8-6 shows the results from a PPL permeability test that was conducted that closely simulated MBHE conditions. The test section pressure drop was measured across a 7 ft deep fixed bed of CFB ash as a range of airflow rates was passed through the bed. The intersection of the test data and the Rose & Risk correlation line indicates the predicted air permeability rate under these conditions. For a 7 ft bed of CFB ash, the air permeability rate was 0.04 ft/sec.

Figure 8-6: MBHE Permeability Test Results and Rose & Rizk Correlation

The model was then used to calculate the potential solids temperature rise if all of the available air reacted with the carbon passing through the MBHE. The model considered different air flow cases: air leakage based on the predicted permeability rate, all solids control valve air entering MBHE, and enough air flow to just fluidize the MBHE. Solids flow rates were varied to a low of 20 ft/hr (as used during the CMB MBHE tests) and considered as high as 350 ft/hr (maximum design velocity for a CFB MBHE).

The results in Figure 8-7 show that the solids temperature rise for the average bulk solids temperature under all air flow scenarios is very low when the solids velocity through the MBHE are within the CFB MBHE design range of 200 to 350 ft/hr. The predicted solids temperature rise remains low at very solids flow rates when the air leakage is limited to the permeability rate. The solids temperature rise at low solids flow rates does become significant (over 150°F) if all of the control valve air rises through the bed – as occurred when the vent line plugged on the MTF MBHE. Note however that the plugged vent would not have had a significant impact if the solids velocity were much higher, as designed for in a CFB MBHE.

Figure 8-7: Combustion Temperature Rise vs Solids Velocity

Conclusions and Recommendations

The MBHE pilot plant results have shown that a properly designed MBHE can be operated with reliable solids flow and with an absence of solids agglomeration. Most of the early MBHE operational problems have been addressed through proper refractory curing and vent design. However, no amount of analysis nor MTF testing will completely demonstrate that the MBHE can operate trouble free through all of the flow conditions and plant upsets that can occur during long-term operation. An extended duration MBHE test was therefore been proposed and accepted by the DOE as part of the approved CMB program modification. The long term MBHE testing is described further in Section 12.

8.2. Alternate Heat Transfer Media

The CMB pilot plant tests have shown that a layer of ash can build up on the sintered bauxite used as the solids circulating material. The buildup comes from softened ash from burning coal particles sticking to and wetting the bauxite surface. The ash buildup is very hard and tenacious. With time, the ash accumulation can impact the size and aerodynamic properties of the bauxite and the particles will need to be withdrawn from the system and replaced. Since the bauxite is quite expensive, an effort was made to identify alternative candidates for the CMB heat transfer media. Because of the large volume of required material, a primary requirement for the alternate media is that it is commonly available worldwide and relatively inexpensive. The physical and chemical properties desirable in a CMB heat transfer media are:

- Good heat transfer properties
- Minimal interaction with the coal ash and lime sorbent or at least
- The media can be separated from the ash with a minimum of effort
- Appropriate particle size and density for CMB design velocities
- Low mechanical attrition
- Low abrasion potential

A search was made of viable heat transfer materials. Table 8-1 shows the eleven alternative candidates that were selected for evaluation, along with the baseline sintered bauxite from Norton-Alcoa. Figure 8-9 shows a graphical matrix that describes some of the properties of the various heat transfer media. The x-axis shows increasing particle size while the y-axis shows increasing particle density. The location of the circle in the graph shows the relative particle size and density, while the size of the circle shows the relative cost of the particle. The letter in the first column of Table 8-1 corresponds to the letter next to each circle in Figure 8-8. Some of the particles would need further processing (screening or further size reduction) to have aerodynamic properties similar to the sintered bauxite particles.

Table 8-1: Alternate CMB Heat Transfer Media

	Company	Product	Material	Size [μm]	Density	Price [\$ /ton]	Price [cents/lb]
A	Norton-Alcoa	1708-920	Sintered Bauxite	600-800	220	540	27
B	Specialty Minerals	CCA	Dolomite coarse	pebble size	175	25-30	1.25-1.5
C	Specialty Minerals	DF 1000, crushed	Dolomite	149-1190	175	25-30	1.25-1.5
D	CE Minerals	Mulcoa 47, 4x8 mesh	Calcined Kaolin	2380-4760	163	250-300	12.5-15
E	CE Minerals	Mulcoa 47, 8 mesh	Calcined Kaolin	0-2360		111	5.6
F	CE Minerals	Mulgrain 47, 14x28 mesh	Calcined Kaolin	600-1410	163	250-300	12.5-15
G	CE Minerals	RK Bauxite	Calcined Bauxite	590-840	199	250-300	12.5-15
H	CE Minerals	Alpha Star	Calcined Bauxite	590-840	222	250-300	12.5-15
I	Eurolime	Slag from alloy industry	Low in iron*	?	?	Approx. 125	Approx. 6
J	LaFarge	Secar 51, 0-1 mm	Aluminate cement	0-1000	188	400-450	20-22.5
K	A/S Olivin	Mix 1800	Olivin sand	0-2000	206	120	6
L		Recycle ash from the CMB process	Lime/coal ash mix	600	approx. 1	0	0

Figure 8-8: Alternate Heat Transfer Media

Alternate Media Test Review

A laboratory-screening test was then developed to assess the potential of each of these candidate materials for use in a CMB environment and to reduce the number of candidates for future MTF testing.

Some of the key performance questions that needed to be answered include:

- Will it be able to operate at CMB temperatures without agglomeration?
- How will the particle size distribution, density, and shape impact gas-to-solid and solid-to-tube heat transfer?
- Will the particle quickly attrit and thus require a lot of makeup?
- Will the particle hardness and shape cause excessive erosion?

Tests that may be need on the heat transfer media itself:

- Chemical composition and particle size
- Melting/softening temperature
- Agglomeration at high temperatures
- Heat transfer tests
- Hardness
- Attrition
- Shape factor
- Particle density
- Quick and dirty heat capacity/conductivity
- Thermal stability
- Porosity

The following section describes tests on a blend of the heat transfer media mixed with ash (Pittsburgh #8), lime, and a synthetic mix of coal ash components.

The objective of the tests was to study the thermal stability in a simulated CMB environment (thermal, chemical) with respect to melting, softening, and tendency to agglomerate. The tests simulated the conditions in which the CMB heat transfer media would be exposed to coal ash and limes during combustion. The degree of interaction between the media and the coal ash and lime would then be assessed by the physical appearance and tenacity of the agglomeration.

Experimental

The following CMB Heat Transfer Media Samples were sized to a uniform 16 x30 mesh (1200 x 600 microns):

- Current CMB Bauxite
- Speciality Minerals Dolomite
- CE Minerals Mulcoa 47 Calcined Kaolin
- CE Minerals RK Bauxite
- LaFarge Alumina Cement
- A/S Olivine Olivine Sand
- MTF Thames Bed Ash

A Pittsburgh No. 8 coal / lime mixture equivalent to 32% calcined lime and 68% coal ash was prepared. This was done using the Pittsburgh #8 seam coal prepared to < 60 mesh (250 microns) and a pulverized, freshly calcined sample of North Adams quicklime.

The sized CMB media samples were blended with the coal / lime mixture in a proportion equivalent to 80% by weight media and 20% coal ash/lime. This blend was then ashed at 1380°F following the proscribed ASTM procedure for producing coal ash. The fully oxidized material was then cooled, transferred into a vial or bottle, and then stored in a desiccator.

An electrically heated horizontal tube furnace was preheated to approximately 1600°F with a furnace purge of air at a rate of approximately 1 change per minute. The blend of CMB media and coal ash/lime was transferred from its storage container to a ceramic combustion boat. The boat shown in Figure 8-9 was placed into the hot zone of the furnace that holds up to 3 boats per test. The furnace program was activated and the temperature was increased at 27°F per minute to the final temperature of 2100°F, 2200°F, or 2300°F. The boat containing the CMB media and ash/lime mixture was held at the final temperature for 1 hour. The boat was then removed from the furnace, cooled, marked, and placed into a desiccator for storage. These initial tests helped to identify the proper laboratory test temperature that would show an impact with the various materials. The resulting agglomeration then was examined visually for signs of material degradation, wetting, reaction, etc.¹²

Figure 8-9: Boat in Furnace

80% CMB Bauxite at 2100°F Observations: The coal ash/ lime mixture at 2100°F (see Figure 8-10) was mostly unchanged from the as-heated material. Some of the mineral particles exhibited signs of plasticity and coalescence. However, the surface of the bauxite grains remained unchanged. The bonding strength of this agglomerate was low, as there was little plastic material to effect particle to particle bonding. 2100°F appeared to be too low a temperature to visually reveal any real interactions between the ash/lime and the CMB ash.

¹² *Status Report on the Testing of Alternative CMB Heat Transfer Media*, Kurt Johnson, ALSTOM Power Internal Report, May 2004

Figure 8-10: 80% CMB Bauxite at 2100°F - 7.5X magnification

80% Bauxite at 2200°F Observations: The coal ash/ lime mixture at 2200°F (see Figure 8-11) had been mostly molten with only a few particles (sand) that appeared unchanged. The surface of the bauxite appeared to be mostly covered with a thin layer of glassy material. The grains of bauxite appeared to have been completely wetted by the molten ash. The bonding strength of this agglomerate was high as most of the ash was bonding the bauxite particles together. This appeared to be the ideal temperature to visually observe interactions between the coal ash/lime mixture and the CMB media.

Figure 8-11: 80% CMB Bauxite at 2200°F - 7.5X magnification

80% Bauxite at 2300°F Observations: There was complete melting of the coal ash/ lime mixture at 2300°F, as shown in Figure 8-12. The bauxite particles appeared to have been partially dissolved in the molten ash. With these results, it was decided that 2300°F is too

high a temperature to exhibit any differences between the media samples. All remaining tests were therefore conducted at 2200°F.

Figure 8-12: 80% CMB Bauxite at 2300°F - 7.5X magnification

Specialty Minerals Dolomite at 2200°F Observations: Even though there was complete melting of the coal ash/ lime mixture in Figure 8-13, its physical appearance suggested that the melt froze before cooling. This would indicate some interaction between the dolomite and the ash, perhaps the Mg migrating into the molten glass causing its melting temperature to increase. The dolomite surface did not appear to be wetted to any degree by the ash. Note the very angular particle shapes of the dolomite.

Figure 8-13: 80% Speciality Minerals Dolomite at 2200°F - 7.5X magnification

Mulcoa 47 Calcined Kaolin at 2200°F Observations: There did not appear to be any effect on this calcined kaolin media from the molten coal ash/ lime mixture as shown in Figure 8-14. Conversely, there did not appear to be any influence on the molten behaviour of the ash from the media. The surfaces of the media particles appeared to be partially wetted by the molten ash. The calcined kaolin particles were very angular and might be very hard.

Figure 8-14: 80% Mulcoa 47 Calcined Kaolin at 2200°F - 7.5X magnification

80% RK Bauxite at 2200°F Observations: This bauxite media appeared to be wetted and fully coated by the molten coal ash/ lime mixture, as shown in Figure 8-15. Note the smaller white particles adhering to the media. This appeared to be sand and glass particles from the

coal ash that had not dissolved into the melt. This bauxite contained a very high proportion of angular particles. Bauxite in general is known as a very hard mineral.

Figure 8-15: 80% RK Bauxite at 2200°F - 7.5X magnification

80% Alumina Cement at 2200°F Observations: The particles of alumina cement shown in Figure 8-16 were imbedded into the glass produced from the coal ash/ lime mixture. The glass appeared to have been formed at a lower viscosity than the previous samples. It was unknown whether the calcium in the alumina cement migrated into the melt and lowered its viscosity. The surface of the alumina cement did not appear to have been wetted at all by the molten glass. The particle shape of this material can properly be called shards.

Figure 8-16: 80% Alumina Cement at 2200°F - 7.5X magnification

80% Olivine Sand at 2200°F Observations: The particles of olivine sand were completely coated by the molten coal ash/ lime mixture. Its surface had been completely wetted by the ash. Beyond these physical interactions, there did not appear to have been any chemical exchange between the olivine sand and the molten ash. The olivine sand's particles appeared to be naturally formed and abraded.

Figure 8-17: 80% Olivine Sand at 2200°F - 7.5X magnification

80% MTF Thames Bed Ash at 2200°F Observations: This visual examination of the particles in Figure 8-18 indicated that there were differences between the particles making up this bed ash. All particles seemed to have been wetted by the molten material. However,

the ash on the surface of the large dark particles appeared to have frozen or perhaps penetrated inside, yet the smaller particles seemed to be surrounded by the glass phase. These differences might be attributable to the composition of the bed ash particles: calcium sulfate or coal ash minerals. None the less, there appeared to be a high degree of chemical interaction between the coal ash/lime and the bed ash.

Figure 8-18: 80% MTF Thames Bed Ash at 2200°F - 7.5X magnification

Analysis

The Pittsburgh No. 8 coal used for this experiment was a high sulfur fuel with high iron ash due to its iron pyrite content. It is well known that coals (and mixtures) with iron to calcium ratios of 0.5 to 2.0 and $\text{Fe}_2\text{O}_3 + \text{CaO}$ contents of 25% or more result in lowered ash fusibility temperatures and reduced slag viscosity. Although extreme, the conditions of this test (chemistry and temperature) did represent possible CMB combustion conditions.

The major differences indicated by this testing were whether or not the molten ash wets or penetrated the surface of the heat transfer media particle and if chemical interactions seemed apparent. At 2200°F, the molten ash completely wetted the surface of the current CMB bauxite, the RK bauxite, and the Olivine sand. Partial wetting occurred with the Mulcoa calcined kaolin and the Thames bed ash. No wetting was apparent with the dolomite and the alumina cement. There did appear to be penetration of the molten ash into some of the Thames bed ash particles. The two samples that indicated some chemical interaction were the dolomite and the alumina cement. If these reactions occurred, they appeared to be exactly opposite each other. The molten ash on the dolomite appeared to have frozen before the sample was cooled. Conversely, the molten ash on the alumina cement was glassy and appeared to have had a much lower viscosity at temperature than any other sample. If this was the case for either the dolomite or the alumina cement, then a transfer of magnesium and calcium respectively might have occurred.

The data shown in Table 8-2 and Figure 8-19 was obtained using a force meter connected to a probe 4 mm in diameter that was used to penetrate the media and ash/lime mixtures after oven sintering. This showed little correlation between the sample appearance and the force required to penetrate the sintered material. It did show the dolomite was easiest to penetrate at 2200°F and the Thames bed ash the most difficult. This was a very crude technique that was highly dependent on a number of variables beyond control in this test, but was done to observe if there were any discernible differences.

Table 8-2: Force Required to Penetrate CMB Media Boat Test Samples

Material	CMB Bauxite	Dolomite	Alumina Cement	Olivine Sand	Rk Bauxite	Thames Bed Ash	Mulcoa 47
% Material	80	80	80	80	80	80	80
Temperature	2100	2100	2100	2100	2100	2100	2100
Breaking Strength (lbs)	6	5	24	36	7	5	19
Material	CMB Bauxite	Dolomite	Alumina Cement	Olivine Sand	Rk Bauxite	Thames Bed Ash	Mulcoa 47
% Material	80	80	80	80	80	80	80
Temperature	2200	2200	2200	2200	2200	2200	2200
Breaking Strength (lbs)	64	28	64	55	40	82	82
Material	CMB Bauxite	All penetrations done with a 0.0707 cm ² round surface					
% Material	80						
Temperature	2300						
Breaking Strength (lbs)	119						
Material	CMB Bauxite						
% Material	50						
Temperature	2200						
Breaking Strength (lbs)	>130

Figure 8-19: Graphical Representation of Breaking Force

Pilot tests showed that the current bauxite media was adequate for this purpose even though it was completely wetted by the ash. This suggested that wetting by itself was not sufficient to disqualify a material. However, more extensive interactions such as that exhibited by the alumina cement and dolomite might be problematic, yet highly fuel specific. Certain combinations of fuel and media may result in complete agglomeration while another fuel with the same media may have very desirable ash separation properties. It may be that this behavior is secondary to more obvious mechanical properties such as attrition and abrasion. It is interesting to note that the current bauxite media had particles that were uniformly spherical, presenting minimal abrasion potential, yet presumably as hard and attrition resistant as any other calcined bauxite material. The bauxite, calcined kaolin, and olivine sand were in theory about as hard as the sintered bauxite, yet their shape was highly irregular and could be highly abrasive.

Conclusions

This experiment showed some differences in the interactions between the coal ash/lime mixture and the media. These interactions were both physical such as surface wetting and penetration by the molten ash or chemical such as the apparent dissolution of a component from the media into the molten ash. It was unclear whether surface wetting was relevant to the ash/media separation that must take place in CMB as the current bauxite was completely wetted by this ash. However, the apparent chemical effects of the dissolution of the calcium from the alumina cement into the ash that resulted in a glassy melt would have a profound impact on ash/media separation. Mostly these media candidates need to be judged on the basis of their cost, attrition rates, and abrasion potential.

Table 8-3 shows the analysis for the Pittsburgh #8 coal used for these tests. Table 8-4 summarizes the base physical properties of various samples tested. Additional tests of physical and thermal properties and hot pilot tests were not completed. Before any further analyses were undertaken, a decision was made (as discussed in Section 11) to redirect the project towards accelerating the commercialization of the MBHE.

Table 8-3: Pittsburgh No. 8 Coal Analysis

COMPANY : ALSTOM Power - USDOE PPL NUMBER: 4-0131-C
LOCATION : MTF GREENHOUSE GAS PROJECT ANAL DATE : 04/21/04
SAMPLE ID: PITTSBURGH NO. 8 SEAM COAL FROM PPL BUNKER 3/1/04
CONTRACT : PROJECT : 890673

	AS RECEIVED -----	MOISTURE FREE -----	MOISTURE & ASH FREE -----
AIR DRY LOSS, WT. PERCENT	1.1		
PROXIMATE ANAL., WT. PERCENT			
MOISTURE (TOTAL)	2.4		
VOLATILE MATTER	34.9	35.8	40.3
FIXED CARBON (DIFF.)	51.8	53.0	59.7
ASH	10.9	11.2	
TOTAL	100	100	100
HHV. BTU/LB	12804	13118	14765
LB ASH/MM BTU	8.5		
ULTIMATE ANAL., WT. PERCENT			
MOISTURE (TOTAL)	2.4		
HYDROGEN	4.9	5.0	5.6
CARBON	70.9	72.6	81.8
SULFUR	3.5	3.61	4.1
NITROGEN	1.2	1.24	1.4
OXYGEN (DIFF)	6.2	6.4	7.1
ASH	10.9	11.2	
TOTAL	100	100	100
ASH FUSIBILITY	REDUCING ATMOSPHERE		
I.T. DEGREES F	2155		
S.T.	2265		
H.T.	2318		
F.T.	2372		
TEMP. DIFF. (F.T.-I.T.)	217		
ASH COMPOSITION, WT. PERCENT			
SiO2	42.5	RATIOS	
Al2O3	19.8	BASE/ACID	0.44
Fe2O3	22.8	Fe2O3/CaO	9.62
CaO	2.4	SiO2/Al2O3	2.15
MgO	0.5		
Na2O	0.8		
K2O	1.5		
TiO2	0.9		
P2O5	0.3		
SO3	2.5		
MnO	3.3		
TOTAL	97.4		

CHLORINE CONTENT = 0.051% (DRY BASIS)

FLUORINE CONTENT = 0.049% (DRY BASIS)

BaO 0.1

SrO 0.1

MERCURY CONTENT = 0.04 PPM (DRY BASIS)

Table 8-4: Base Physical Properties of CMB Alternative Media Samples

PPL Sample No.	4-0017-M	4-0018-M	4-0019-M	4-0020-M	4-0021-M	4-0022-M	4-0023-M
Vendor	Specialty Minerals	Specialty Minerals	CE Minerals	CE Minerals	CE Minerals	CE Minerals	CE Minerals
Material I.D.	CCA coarse Dolomite	DF1000 Crushed Dolomite	Muloca 47 4x8 mesh Calined Kaolin	Muloca 47 < 8 mesh Calined Kaolin	Mulgrain 47 14x28 mesh Calined Kaolin	RK Bauxite - Calcined	Alpha Star Calcined Bauxite
Ash Fusbility - <100 mesh CMB Media							
I.T. deg F		>2718	>2718	>2718	>2718	>2718	>2718
S.T.		>2718	>2718	>2718	>2718	>2718	>2718
H.T.		>2718	>2718	>2718	>2718	>2718	>2718
F.T.		>2718	>2718	>2718	>2718	>2718	>2718
Ash Fusbility w/ 80% <100 mesh CMB Media & 20% Coal Ash/Lime Mixture							
I.T. deg F					2365	2380	
S.T.					>2730	>2730	
H.T.					>2730	>2730	
F.T.					>2730	>2730	
Ash Fusbility w/ 50% <100 mesh CMB Media & 50% Coal Ash/Lime Mixture							
I.T. deg F					1680	1698	
S.T.					2579	2562	
H.T.					2591	2570	
F.T.					2608	2585	
PPL Sample No.	4-0024-M	4-0025-M	4-0026-A	4-0027-A	4-0131-C	Ash/Lime	
Vendor	LaFarge	A/S Olivin	N/A	N/A	N/A	N/A	
Material I.D.	R50 Aluminate Cement	Mix 1800 Olivine Sand	MTF Thames Coal Bed Ash	MTF Pitt 8 Coal Bed Ash	MTF Pitt 8 Coal 01-Mar-04	Pitt 8 Coal Ash w/ 32% Calcined Lime	
Ash Fusbility - <100 mesh CMB Media							
I.T. deg F	2664	>2718		2361	2155	2158	
S.T.	2667	>2718	>2718	2372	2265	2176	
H.T.	2671	>2718	>2718	2378	2318	2185	
F.T.	2680	>2718	>2718	2390	2372	2200	
Ash Fusbility w/ 80% <100 mesh CMB Media & 20% Coal Ash/Lime Mixture							
I.T. deg F		2430					
S.T.		>2718					
H.T.		>2718					
F.T.		>2718					
Ash Fusbility w/ 50% <100 mesh CMB Media & 50% Coal Ash/Lime Mixture							
I.T. deg F		1730					
S.T.		2418					
H.T.		>2718					
F.T.		>2718					
% Total Ca				31.64			
% Total Mg				0.3			
% Total C				0.10			
% Total S				10.96			

8.3. CO Emissions Control

The original CMB concept was based upon concentrating all of the heat transfer surface in the MBHE, thus eliminating the need for any combustor or backpass surface. This approach required that the combustion gas be cooled to less than 700°F leaving the furnace. Our pilot plant test philosophy was to operate the MTF with the same gas residence time and temperature profile as we'd expect in a large commercial CMB. The MTF was thus operated with a combustor outlet temperature of 1300°F since that's what we predict at the 65 ft elevation in a large CMB boiler. However, it was observed that CO emissions increased as the outlet temperature dropped, which raised the concern with what happens in the upper regions of the combustor as the gas temperature continued to drop below 1000°F. A number of technical options were looked at for reducing CO emissions and are described below.

8.3.1. Minimizing Char Elutriation

Computational modelling was used to study the net rate of CO conversion in the CMB in two separate phases. Parametric runs were made using Chemkin V3.7 and FLUENT V6.1.22. The Phase I study showed that char particles continued to react and produce CO at temperatures as low as 800°F. However, the gas-phase CO conversion results showed that the CO reaction was essentially quenched as temperatures dropped below 1200°F.

The Phase II study was conducted to assess the net CO production assuming a means was provided to minimize carbon carryover from the combustor. The FLUENT model assumed that a series of U-beams or louvers were located halfway up the combustor that were used to "knock out" or prevent a fraction of the char particles from passing through the upper combustor. The longer char particle residence time in the higher temperature lower combustor would result in greater carbon burnout and lower char elutriation. The model showed that the outlet CO concentration was approximately linear with char particle flow. The CO outlet concentration could thus be reduced if an internal separator, such as a U-beam array, could be designed for the combustor. This design approach would require considerable flow modelling and testing to develop an internal separator that could retain char particles in the lower combustor and also survive in the high temperature environment.

Phase I Analysis. CFD results were configured as 2-D (flow between infinite parallel plates) with adiabatic walls. The k-epsilon turbulence model and the discrete ordinates radiation model were activated. The walls were assumed to have zero wall shear stress, so that the flow was essentially plug flow over the desired height of 100 ft. For a given overall mass flow rate and a prescribed inlet temperature, the computed flow velocity was very similar to the measured superficial gas velocity in the MTF.¹³

The particle model was run with one-way coupling only (i.e., the flow field impacted the particle, but the particle did not impact the flow field). The released particle was assumed to be a char particle consisting of pure carbon plus ash with various initial reactivities, densities, and diameters. Heterogeneous reaction of the particle occurred through the $C + \frac{1}{2}(O_2) \rightarrow CO$ reaction. A baseline char reactivity corresponding to Pittsburgh #8 was

¹³ *Summary of CMB CO Conversion Runs*, David Sloan, ALSTOM Power Internal Report, July 2005

assumed. Additional rates, calculated by multiplying the pre-exponential factor by various factors were also applied, in order to provide adequate coverage of all of the experimental rates in the ALSTOM Power database of drop-tube reaction data.

For the parametric runs, the initial fractional char conversion was arbitrarily fixed at 60%, which corresponded to an initial fixed-carbon composition in the char of 60.9% and an initial char particle density of 32.6 lbm/ft³. The parametric runs were conducted over a range of 4 rates x 7 temperatures x 8 diameters for a total of 224 runs. These runs amply demonstrated that even 80 to 100 micron particles continued to react and produce CO, even at temperatures as low as 800°F or 900°F. However, Chemkin runs (based on the GRI-Mech V3.0 mechanism) of gas-phase CO conversion demonstrated that the CO reaction was essentially quenched at temperatures higher than those values (e.g., 1000°F to 1100°F).

Phase II Analysis. In this analysis, the net CO production was analysed, in a more quantitative fashion, by importing the GRI-Mech mechanism into FLUENT itself and solving FLUENT with detailed (laminar) chemistry. A series of U-beams or louvers were presumed to exist at the 50 ft elevation, which were used to “knock out” or prevent a prescribed fraction of the particles from traversing the upper CMB furnace region (from the 50 ft elevation to the 100 ft elevation at the outlet). A single column of cells was used for the computational domain and a fixed, linear temperature profile of 1300°F at the 50 ft elevation to 650°F at the outlet was applied. An estimated flue gas composition and a carbon-in-ash value of 2.5% were presumed at the 50 ft elevation. The char kinetics was then adjusted until a 0.5% carbon-in-ash value was achieved at the outlet. At the outlet, a gas flow rate of 10,000 lb_m/hr was presumed with a solids loading of 1lb_m solid/lb_m gas.

Computations with both the coupled gas phase reaction and without the gas-phase reaction were completed. It was found that the rate of CO production, even with the lowest char reactivity, was significantly greater than the rate of CO destruction over the length of the reactor. Consequently, the CO mole fraction values without the coupled gas-phase reaction chemistry were only about 10% higher than the CO mole fraction values (about 4.7 % by volume, wet) with the coupled gas-phase reaction mechanism. For an 80-micron particle, it was found that changing the particle reaction rate did not impact the outlet CO a great deal (as long as the particle reaction rate was sufficiently high to cause complete particle burnout). It was also found that the outlet CO concentration was approximately linear with the particle flow rate. In other words, if a certain fraction of the particles could be knocked out by the U-beam array, then the CO outlet concentration would decrease commensurately.

8.3.2. Designing the CMB for a Higher Gas Outlet Temperature

CMB designs were developed that significantly reduced the combustor height, while correspondingly increasing the combustor gas outlet temperature. Although this approach would reduce CO emissions, the cost savings from the reduced combustor height and smaller MBHE costs were overwhelmed by the added cost of a high temperature cyclone, additional refractory, and a backpass. The cost study (reported in Section 11.1) showed that boiler island material costs for the higher temperature design increased by 7% or approximately \$2.5 million with erection for a 2x150Mw_e boiler. Further studies are necessary to determine if even higher temperature designs are required to ensure acceptable CO levels at all load conditions.

8.3.3. Adding In-Furnace Additives to Control CO

Steam injection was tested in the lower elevations of the MTF combustor as a way to increase the OH radicals necessary to catalyse the oxidation of CO. MTF tests and laboratory tests showed that this could reduce CO emissions, although at the expense of increased NO_x. As reported in Section 10.6, injecting 100 lb/hr of steam into the MTF reduced the CO by about 50 ppm. Although the magnitude of CO reduction was not large, it is conceivable that the steam performance could be improved with further optimization of steam injection location and volume. A comparable steam injection into a 2x150Mw_e boiler would have an evaluated cost of about \$2 million.

8.3.4. Using Backend Additives

A MTF test campaign for a different program included some mercury capture tests with halogenated activated carbon injected after the combustor. Figure 8-20 shows that the CO emissions decreased from about 10 ppm to 2 ppm while the additive was injected. It was uncertain if the additive had a small impact on CO (8 ppm) or if it was actually a much larger impact (80% reduction) that would have been more significant if the starting CO levels were higher. Assuming some improvements in sorbent distribution and an 80% CO reduction at the MTF sorbent rates, then the evaluated costs for a 2x150Mw_e boiler would be \$4 million.

Figure 8-20: CO Reduction With Activated Carbon Injection

8.3.5. Adding a Backend CO Oxidation Catalyst

Catalysts are available for CO oxidation and may be used to oxidize CO after the combustor. An extra bank of CO catalyst could be loaded into the front end of a

conventional SCR. The catalyst cost is typically 20 – 30% of the total SCR cost. Adding additional CO catalyst would certainly increase the total SCR catalyst cost, but it might only require modest cost increases for the non-catalytic portion of the SCR, which is the major portion of its cost. While expensive, the cost impact of this approach would be much less if future environmental legislation required more stringent NO_x emissions that could only be met with SCRs. For example, the consensus CURC/EPRI Roadmap for Advanced Combustion-based Steam Plants targeted NO_x emissions for bituminous coals at 0.05 lb NO_x/MMBtu in 2010 and 0.01 lb NO_x/MMBtu in 2020. Achieving these targets would likely require the use of SCRs for all combustion technologies.

One concern with backend catalysts that needs to be addressed is the potential for catalyst erosion and deactivation from sulfur poisoning. Several catalyst vendors were contacted for quotes, but only one vendor was willing to propose a system that would survive in this combustion environment. Even so, this vendor would only quote the catalyst if it was located after the NID system – i.e. with very low particulates and SO₂. This required additional costs and equipment to reheat the flue gas after the NID up to 650°F entering the catalyst and then to recover the sensible energy in the gas before it was exhausted in the stack. The evaluated costs for a 2x150Mw_e plant was \$3.5 million, plus the costs of the additional gas reheat and cooling equipment. These costs included annual catalyst replacement of a system that was capable of removing 90% of 1000ppm CO.

8.3.6. CO Mitigation Evaluated Costs

The following table summarizes the evaluated costs of the various CO mitigation options discussed in this section. The various options would likely have evaluated costs of \$3 - \$5 million for a 2x150Mw_e CMB boiler, although the cost impact might be reduced with better integration or in combination with future more stringent environmental controls. The performance of some options was unclear, although the backend CO catalyst was commercially available for other applications and could most likely be safely adopted for this situation. Additional flow modelling and testing is necessary to identify the most cost effective approach to mitigating CO emissions.

Approach	Evaluated Costs	Comments
Internal Solids Separator	Not assessed	Requires flow analysis and testing
High Combustor Outlet Temp	\$2.5M	May still be marginal or need higher temps
Steam Injection	\$2M	CO reduction may not be sufficient
Activated Carbon Injection	\$4M+	Unknown impact
CO Catalyst	\$3.5M + heat exchangers	90% CO reduction (design for 1000 ppm)

Figure 8-21: CMB CO Emission Control Options for a 2x150Mw_e Boiler

8.4. Alternative Approaches to Sulfur Capture

The CMB combustor meets SO₂ emission requirements through a combination of in-furnace sulfur capture and sulfur capture in the backend NID system. This approach is similar to the method used in many CFB boilers to achieve high levels of sulfur capture. However, the distribution of sulfur capture between the CMB combustor and the NID system can be quite different as compared to a CFB. Because of the high CMB combustor temperatures, in-furnace sulfur capture may be limited. Depending on the coal sulfur content, the limited in-furnace sulfur capture could also have a significant negative impact on the boiler efficiency and overall economics relative to a conventional CFB design. This makes the NID performance crucial for this technology. Development decisions were therefore made early in the CMB project to minimize sorbent consumption and the resulting boiler efficiency penalty by maximizing the sulfur capture in the NID system through operation at higher humidity levels than normal commercial practice. MTF tests confirmed that NID performance was greatly enhanced with no reliability issues while operating under these new conditions.

The CMB process was reassessed during a CMB Design Review Meeting in 2005 and a new strategy was recommended for mitigating SO₂ emissions. For risk mitigation reasons, it was decided that the NID system should be restricted to using commercially demonstrated humidity levels despite the positive MTF results. This approach would require a significantly higher sorbent feed rate than that observed during MTF tests using the higher humidity levels (see Section 10.7). It was therefore decided that an evaluation should be made of alternative approaches for CMB sulfur capture. The following section discusses in-furnace impacts on boiler efficiency, alternate capture technologies, and provides a cost comparison of the various technologies.

8.4.1. In-Furnace Impact on Boiler Efficiency

The sulfation reaction is very exothermic. In conventional CFB boilers, most of the sulfur removal takes place within the combustor, while the NID system or comparable system is used primarily for final polishing of the gas. The sulfation energy released in the furnace heats up the combustion gases. At these high temperatures, the energy can be readily recovered. This energy is treated as a boiler efficiency credit when it is released in the combustor. Energy is also released when the sulfur is captured in the NID system. However, the gas temperature at this location is so low that any gas temperature rise due to sulfation cannot be effectively recovered. Thus, a boiler efficiency credit only occurs when the sulfur is captured in the furnace. There is also a boiler efficiency loss associated with the endothermic calcination of the limestone in the furnace.

Figure 8-22 shows the effect of combustor sulfur removal on the boiler efficiency credit for two different coals at a constant sorbent flow, which in turn results in a fixed calcination heat loss. The high sulfur coal represented by the upper red line is for a 4% sulfur coal with 35% ash. The lower blue line is for a 2% sulfur coal with 12% ash. The maximum boiler efficiency credit for these coals occurs when all of the sulfur is captured in the furnace. The shaded red region shows the typical range of in-furnace sulfur capture in CFB boilers. CFBs typically capture 80% - 95% of the sulfur in-furnace and get almost the full sulfation credit at reasonable Ca/S mole ratios. This is because the sorbent has such a long residence time at the optimum furnace temperature for sulfur capture.

Figure 8-22: Effect of In-Furnace Sulfur Capture on Boiler Efficiency

The expected CMB in-furnace sulfur capture ranges from 20 – 70% for a given sorbent flow and is also shown in Figure 8-22 as a light blue shaded region. The CMB design requires that a significant portion of the sulfur capture take place in the NID system. The combustor can be operated to increase the in-furnace capture, although there may be limits to how much in-furnace capture can be achieved at reasonable Ca/S mole ratios. This figure shows that the CMB boiler efficiency credit for sulfation can be between 0.1 – 1.2% less than a CFB for a low sulfur coal and double that for a high sulfur coal.

It also shows that the CMB is sensitive to coal sulfur content. With 50% in-furnace sulfur capture, a CMB firing this low sulfur coal receives half of the maximum sulfation credit, or 0.8%. The sulfation credit increases to 1.6% with the high sulfur coal. For comparison, a CFB with 95% in-furnace sulfur capture receives 95% of the maximum sulfation credit or 1.5% with the low sulfur coal and 3.0% with the high sulfur coal. These results show the importance of considering the CMB operating conditions with respect to in-furnace sulfur capture and their impact on boiler efficiency.

Table 8-5 gives a further comparison on the impact of in-furnace sulfur capture on boiler efficiency. The table compares the projected performance for a CFB w/o NID™ system, CFB w NID system, and a CMB w NID system, all for two different coal sulfur levels (uncontrolled SO₂ of 5478 and 3000ppm) and for two different levels of overall sulfur capture (90% and 98%). The first grouping of three technologies (at 90% capture and 5478ppm SO₂) shows that the CMB only achieves 35% in-furnace sulfur capture, while the CFB w/o and w NID system get 90% and 84% in-furnace capture, respectively. The resulting CMB boiler efficiency is 1.8% points less than the CFB w NID system for this scenario. Increasing the coal sulfur content to 5478ppm SO₂ uncontrolled (see second grouping in the table) increases the in-furnace capture for all three technologies, with a significant increase for the CMB. The resulting CMB boiler efficiency penalty is now 1.6% points as compared to the CFB w NID system. The final grouping of technologies in this table shows the impact of reducing the coal sulfur content to 3000ppm SO₂. In this case, the CMB boiler efficiency penalty is reduced in half to 0.8% points. This table clearly shows the importance of maximizing the CMB in-furnace sulfur capture performance, especially when firing high sulfur coals.

Table 8-5: Impact of In-Furnace Sulfur Capture on Boiler Efficiency

Technology	CFB	CFB	CMB	CFB	CFB	CMB	CFB	CMB
Sulfur Capture	90%	90%	90%	98%	98%	98%	98%	98%
Uncontrolled SO ₂	5478	5478	5478	5478	5478	5478	3000	3000
Sorbent Size	Std	Std	Fine	Std	Std	Fine	Std	Fine
Combustor Performance	No NID	With NID	With NID	No NID	With NID	With NID	With NID	With NID
Ca/S	1.75	1.4	1.4	2.75	1.95	1.95	1.76	1.76
% Sulfur Capture	90.2%	84.1%	35.0%	97.5%	92.5%	50.0%	90.3%	50.0%
Unused Ca/S	0.85	0.56	1.05	1.77	1.02	1.45	0.86	1.26
Remaining SO ₂ (ppm)	538	873	3561	135	408	2739	291	1500
NID Performance								
Humidity		50%	70%		50%	70%	50%	70%
Remaining Ca/S	8.6	3.5	1.6	71.9	13.7	2.9	8.8	2.5
% Sulfur Capture	0.0%	34.4%	86%	0.0%	73.3%	96%	80.1%	95%
NID SO ₂ Capture		300	3053		299	2635	233	1424
Total Performance								
% Sulfur Capture	90.2%	89.5%	90.7%	97.5%	98.0%	98.1%	98.1%	97.5%
Boiler Efficiency %	82.5	82.9	81.1	81.2	82.3	80.7	82.3	81.5
Delta Efficiency %			1.8			1.6		0.8

8.4.2. Alternate Sulfur Capture Technologies

A wide range of alternate approaches to sulfur capture were considered, including both commercially ready technologies and some under development. The following section briefly describes the technologies that were considered.

Wet Scrubber

Nearly 100% sulfur capture is normally done in the wet scrubbers since they are very efficient contacting devices and can significantly reduce the required sorbent consumption. Since the sorbent is not calcined in the boiler, it also eliminates that boiler efficiency penalty. However, wet scrubbers are very capital intensive and have high power requirements. Wet scrubbers will only make economic sense for large capacity boilers and if there is a local market that the resulting ash can be sold to as gypsum. This study assumed that a single 300Mw wet scrubber could be used for a 2x150Mw_e boiler.

Dry Scrubber With Commercial Lime

This approach is similar to the wet scrubber and eliminates the boiler efficiency penalty from calcination and also reduces the sorbent requirement, but at the expense of a very expensive sorbent.

Dry Scrubber With Lime Prepared By The CMB

This approach uses the hot solids from the CMB to calcine the limestone for use by the dry scrubber. This method was not evaluated at this time and requires some integration to effectively transfer sensible heat from the hot circulating solids to the sorbent.

Increasing In-Furnace Sulfur Capture Followed By A Conventional NID System

One approach towards improving the furnace performance is to couple it with a process to enhance the sorbent reactivity, such a steam reactivation. The gas residence time available in the optimum temperature window for sulfur capture in the CMB furnace is much shorter

than in a conventional CFB. However, the CMB recycle solids are available at an ideal temperature for steam reactivation, which may offset the short residence time and significantly improve furnace performance.

Using Cyclone NID As An Alternate To The NID System

The Cyclone NID (patent pending) works similar to the NID™ system, except that the mixer and reactor duct are added upstream of a small cyclone and before the fabric filter or ESP. In this arrangement, the Cyclone NID (CycNID) is also able to operate with much higher humidity levels for additional performance without the plugging concerns in the downstream ESP or fabric filter. This is because the cyclone will be removing the bulk of the recirculating solids and the downstream particulate collector will thus only handle normal solids loading. A reheater may be needed downstream of the cyclone during high humidity operation to avoid acid condensation by warming the gas before it reaches the stack. The Cyclone NID can also be combined with in-furnace sorbent enhancement (steam reactivation) for further improvements.

The dry scrubber and wet scrubber are both commercial technologies. The NID system with high humidity has been tested with very good results in every CMB test campaign, but this level of humidity is beyond current commercial practice. Hence, a long-term test needs to be performed to evaluate the feasibility of this system. Steam reactivation has also been demonstrated in the MTF during CFB operation and works best with low solids temperatures. However, its performance is unknown with the short residence time at optimum temperatures in the CMB combustor. The Cyclone NID concept has not been tested, but should work in principal similar to the NID system at either conventional humidity levels or with high humidity levels.

Table 8-6 summarizes the required sorbent, in-furnace sulfur capture, and commercial readiness of each of these technologies.

Table 8-6: Status of CMB Sulfur Capture Options

Sulfur Capture Technology	Sorbent	In Furnace Sulfur Capture	Commercial Status
NID with High Humidity	CaCO ₃	up to 50%	High NID humidity not commercially demonstrated
Steam Reactivation + NID	CaCO ₃	up to 90%	Steam Reactivation effectiveness unproved in CMB design
Wet Scrubber	CaCO ₃	0%	Commercial
Dry Scrubber	CaO	0%	Commercial
Cyclone NID	CaO	0%	CycNID has not been tested
Cyclone NID with High Humidity	CaCO ₃	up to 50%	CycNID has not been tested
Steam Reactivation + CycNID	CaCO ₃	up to 90%	CycNID has not been tested Steam Reactivation effectiveness unproved in CMB design

Figure 8-23: CMB Sulfur Capture Options

8.4.3. Capital Costs of Sulfur Capture Options

The evaluation of sulfur capture options included an assessment of sorbent requirements, capital costs, and overall economics. The following section discusses the capital costs (in 2005 dollars) associated with each alternative for a 2x150Mw_e CMB plant. It also includes the comparable costs for a CFB both without and with a backend NID system. The economic comparison of these systems (including sorbent requirements) is included in the overall CMB plant economic evaluation discussed later in Section 11.1.

The capital costs for the environmental equipment were based upon designs using a high sulfur lignite fuel (5478 ppm SO₂ uncontrolled) and 98% sulfur capture. Quotes for material costs were obtained for most of the equipment from ALSTOM Power. Some costs (particularly sorbent handling systems and Cyclone NID) were obtained from other studies or from material take-offs. Table 8-7 summarizes the capital cost difference (relative to a CFB with NID system) for the environmental cleanup equipment for a 2x150Mw_e boiler for various CMB and CFB configurations. No costs were included for CO mitigation at this time. All of the designs using a NID system or dry scrubber included an added charge to the ID fan due to the higher-pressure drops through the backend system. The NID system or CycNID designs with high moisture also included costs to reheat the gas before exhausting it to the stack.

The CFB with no NIDTM system and the three designs using CycNID all had low environmental equipment capital costs. The CFB with NID system and CMB with steam reactivation/NID system had comparable costs. The wet scrubber was clearly the most expensive design.

Table 8-7: Sulfur Capture Capital Costs for 2x150Mw_e CMB (\$K)

	CFB No NID Baghouse	CFB With NID	CMB NID Hi H2O	CMB Stm React FDA conv	CMB Bagfilter Wet Scrub
Delta CFB w NID	(4,712,680)	-	750,000	105,000	22,392,144
	CMB Dry Scrub Lime	CMB Adv NID Hi H2O CaO, bagh	CMB Adv NID Hi H2O CaCO ₃ , bagh	CMB Stm React Adv NID low hum, CaCO ₃	
Delta CFB w NID	(1,950,570)	(5,055,308)	(3,350,000)	(3,995,000)	

- Assume High Sulfur Coal (5478 ppm Uncontrolled)
- Design for 98% Sulfur Capture
- Performance and Quotes by ECS
- Material Costs Only
- Assume Erection Costs Equal Material Costs
- Not Including Any Costs for CO Control

8.5. Fuel Flexibility

The CMB is capable of handling a wide range of fuels similar to conventional CFBs, although some limitations may result due to the higher temperatures in the lower furnace. The best fuels for CMB are high reactivity, low sulfur fuels, including many subbituminous, high volatile bituminous, lignites, brown coals, and biomass. These attributes aid carbon burnout, CO emissions, low sulfur capture equipment requirements, and are attractive for ultra high efficiency and potential CO₂ control schemes. Low reactivity, high sulfur content, very high ash, and very low melting ash fuels may require design compromises that increase costs. Changes in sulfur capture methodology may enable high sulfur fuels to be used cost effectively in the CMB. Petroleum coke, with its high sulfur content, will show better economics from a sulfur capture standpoint as discussed in Section 8.4.

Fuel Flexibility

The CMB is able to operate with a broad range of fuels, including most coals and petroleum cokes. This is because of the relatively low combustion temperatures (relative to PC firing), the lack of a backpass or low backpass temperatures, and various options for sulfur control. The CMB is also less sensitive to highly corrosive (Cl) and highly erosive (quartz) fuels because the high temperature heat transfer surfaces are located in the MBHE away from combustion gases.

Agglomeration

The CMB may be limited by potential bed and MBHE agglomeration associated with very high alkali content fuels. This includes high sodium coals, high potassium biomass, etc. with > 0.8 lbs alkali/MMBtu. The CMB may have potential problems with high ash content, low ash melting fuels if burning continues as the circulating solids enter the MBHE.

Sulfur Capture

The CMB as presently arranged with low in-furnace sulfur capture could have significant boiler efficiency losses for high sulfur fuels. Very high sulfur fuels (>5 lbs SO₂/MMBtu) require substantial post combustion SO₂ capture, which limits the use of conventional NID systems. These are both economic issues that may be addressed by alternate methods for

sulfur capture in a CMB system. Improving in-furnace sulfur capture will reduce concerns of using high sulfur fuels in a CMB system.

9. MTF TEST CAMPAIGNS

The objective of this task was to complete the performance testing of the CMB Proof of Concept program during a MTF test campaign with two different fuels. These tests would evaluate combustion, NO_x control, sulfur capture, heat transfer, and agglomeration. This task was subdivided into two separate campaigns. A further objective was to examine gas-to-solids heat transfer with optimum gas and solids flow distributions in a test campaign in the scaled MTF. This section describes the MTF facility, the facility modifications, and the test campaigns that were completed to evaluate the CMB concept.

9.1. Description of ALSTOM Power Multi-Use Combustion Test Facility

ALSTOM Power Inc.'s "Multi-Use Combustion Test Facility" was developed by its US Power Plant Laboratories (PPL) to support the Power Generation Business's strategic development needs. This facility provides the flexibility to perform pilot 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 at the US Power Plant Laboratories facilities in Windsor, CT. It has accrued over 8000 hours of operation on a wide variety of fuels since the initial startup in 1997.

The MTF can be operated under atmospheric conditions at firing capacities up to 3 MW_{th} (9.9 MBtu/hr). The combustor has an inside diameter of 40 inches and can be configured to have an overall height of more than 60 ft. It is equipped with comprehensive instrumentation and control systems and is housed in an enclosed building with supporting ancillary equipment.

The combustor design provides the flexibility to switch lower combustor sections, which permits operation under various firing regimes. When operating in pulverized coal firing modes, the facility's capabilities allow fundamental combustion and emission investigations, fuel characterizations, and assessment of fireside impacts on various aspects of boiler operation.

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 fluidized bed combustion 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 Fabric Filter Test Facilities are also incorporated into the Multi-Use Test Facility.

The MTF facility was initially modified in this project for CMB operation with natural gas or coal firing. The hot solids pick up and transport line was installed to transport bauxite from the bottom of the combustor to the top. The solids transport line was a water-cooled pipe that allowed testing of the hot gas-to-solids heat transfer before the moving bed heat exchanger was installed. The MTF baghouse was also converted into a NID system.

The final modification was adding a moving bed heat exchanger, which allowed the MTF to operate in a fully integrated CMB configuration. The moving bed heat exchanger was not located directly underneath the combustor as envisioned in a commercial unit, since this

was not cost effective. Instead, the moving bed heat exchanger was placed next to the lower combustor and a separate pneumatic transport system carried the hot solids up to it. Areas where the MTF modifications were made are shown in Figure 9-1.

Figure 9-1: MTF Modifications

9.2. MTF TEST 1 – September 2003

Objective

The objective of this task was to complete the performance testing of the CMB Proof of Concept program during a MTF test campaign with two different fuels (Pittsburgh #8 bituminous coal and a petroleum coke) and with different sizes of limestone. The test matrix developed for this campaign had four major objectives:

1. Demonstrate improved solids flow distribution and heat transfer in the MBHE. These tests evaluated the MBHE performance with the improved solids distributor in the lower outlet hopper.
2. Evaluate process improvements to combustor performance. Tests were conducted to evaluate the impact of eliminating ash recirculation from the cyclone on combustion performance and on emissions. NO_x reduction methods included variations in staging, excess air, overfire air elevation, and SNCR. CO reduction methods included multiple overfire air injection points and steam injection.
3. Evaluate sulfur capture in the combustor and NID system. Limestone size distribution were varied to assess the impact on calcium oxide carryover to the NID system and on NID performance. Operation without cyclone ash reinjection was also tested to assess the impact on the calcium carryover to the NID system. NID optimization included variations in humidity and NID ash recirculation rates.
4. Evaluate gas-to-solids heat transfer with poor solids distribution. The solids leaving the solids distributor at the top of the combustor were deflected towards one sidewall to assess the impact of poor solids distribution on heat transfer. Detailed temperature profile data were gathered with both uniform and biased solids distribution to validate the CFD model scaling technique that would be used during the MTF scaling heat transfer test campaign.

Modifications

MBHE

The MTF test program in Phase I demonstrated that high heat transfer rates could be achieved in the MBHE, provided that the solids flowed uniformly through the tube bundles. However, problems with the outlet solids distribution resulted in solids flow maldistribution throughout the MBHE and reduced heat transfer to the tube bundles.

A cold flow test was therefore conducted in the MBHE with the old solids distribution grid at the hopper outlet to give visual indications of potential solids flow maldistribution. The access door was removed from the MBHE and bauxite was loaded to a level of 12" above the grid. The rotary valve was then started and the flow of the bauxite was observed. The maximum solids flow occurred directly above the rotary valve, with significantly less flow around the perimeter. Figure 9-2 shows the bauxite distribution at the end of the test. The distribution rubble screen was exposed in the area directly over the rotary valve, while higher solids levels were observed around the perimeter. The exposed area showed that the solids were primarily flowing through only four out of the sixteen outlet hoppers. This test indicated that the solids maldistribution originated because the hopper outlet design did not adequately distribute the solids flow as it flowed into the rotary valve.

Figure 9-2: Solids Flow Pattern with Rotary Valve

A modified distributor design was developed that replaced the outlet distributor with sixteen individually controlled hoppers. The hoppers could be cycled on and off pneumatically to control the solids flow throughout the MBHE. A proof of concept test of the active grid design was completed on a single hopper and then on four hoppers arranged in a 2 x 2 array. The performance was measured to determine ability to start, stop, and control solids flow rate. Minimum air flow requirements were measured to ensure the ability to pass rubble through the grid. This test was described in Section 7.3.

The solids distributor grid system at the bottom of the MTF CMB heat exchanger was then converted to a 4x4 pneumatic shoe flow control configuration. A sixteen solenoid valve / programmable control system was designed and fabricated to control solids flow. This system permitted control of the average solids flow to meet demand and also to bias the flow from one side of the heat exchanger to another in order to eliminate temperature imbalances. A pneumatic loop seal was installed between the grid system and the pneumatic transport system to minimize the effect of transport air backpressure on the system. The 45° elbow in the transport system that had exhibited erosion wear was replaced with a 90° impact tee that is less susceptible to erosion.

Cold flow distribution tests were then run on the 4x4 pneumatic shoe flow control configuration that was installed in the bottom of the MBHE. Solids were loaded into the bottom portion of the MBHE with the east door removed for access. The pneumatic actuators were cycled through all sixteen hoppers while observing the resulting flow pattern.

Figure 9-3 shows the solids flow profile in the MBHE after cycling the hopper injectors once. Solids flow was initially poor along the south and east wall due to air leakage through the gap between the wall and the grid support. Air leakage caused excessive solids flow in this region, as evidenced by the areas showing the bare grid in this photo.

Figure 9-3: Excessive Flow Along MBHE South Wall From Leakage

The MBHE hoppers were modified to eliminate areas of air leakage. Figure 9-4 shows the improved solids distribution after the leakage was stopped and after the injectors were cycled once. This test also demonstrated the necessity of a vent line to relieve the air pressure in the hopper after the air injectors were pulsed. The solids distribution was further improved by adding a vent line from the hopper discharge to the top of the MBHE.

Figure 9-4: Improved solids distribution after leakage stopped

Refractory-Lined Transport Lines

The refractory-lined transport lines were inspected for wear. The refractory-lined elbows and cyclone above the MBHE had little wear. Ultrasonic measurements were taken along the stainless steel liner pipe inside the straight sections of the refractory-lined pipe. The pipe sections also showed little signs of wear. The only area with any significant wear was a 45° bend in the stainless steel pipe section between the MBHE bottom outlet and the vertical transport line to the top of the combustor. This section was removed and repiped to eliminate the 45° bend.

Cyclone

The solids return leg from the cyclone to the combustor was bricked off at the bed and covered with a plate in the seal pot below the cyclone. This modification ensured that cyclone ash will only be reinjected through the transport system below the FBHE. The ash recirculation flow could now be measured by heat balance.

The cyclone was also modified to provide the ability to shut off ash recirculation by blocking the cyclone bottom exit to increase the carryover of ash (and therefore calcium) to the backend scrubber. The cyclone water-cooled dipleg was replaced by an uncooled dipleg with an isolation valve. This change allowed the cyclone to effectively be bypassed by closing the isolation valve and thus simulate CMB operation without a cyclone.

Overfire Air

The overfire air system was modified to provide the ability for multiple overfire air locations with some steam addition to see the influence on carbon monoxide emissions.

Solids Diverter

A solids diverter was installed at the top of the furnace to bias the bauxite flow towards the sidewall. The diverter provided the ability to disrupt the bauxite distribution to see the effect on the bauxite-gas cooling profile. Figure 9-5 shows two views looking up at the solids diverter plate just below the solids discharge at the top of the furnace.

Figure 9-5: Diverter Plate Below Solids Discharge

Test Matrix

Figure 9-6 outlines the test matrix that was planned for the first test campaign. Many of the tests had multiple objectives, as the MBHE, combustor, and NID performance could all be tested simultaneously.

Only a portion of the test matrix was completed in the September test campaign because of some difficulties with the cooling water pond level and MBHE flow. The balance of the test matrix was completed in the November test campaign, as discussed in Section 9.3.

Figure 9-6: CMB Test Matrix for 1st Test Campaign

Operational Summary

The first MTF test campaign for CMB Phase II started on September 3rd and was completed on September 11th. Tests were conducted with Pittsburgh #8 coal, which had 4% sulfur, and two limestones of varying size distributions. The tests were suspended after a pond that receives cooling water discharge reached its maximum allowable level. The limit was reached prematurely because of the large volumes of rain that the region received in prior months.

The premature shutdown also allowed us to address two solid flow problems. The MTF bed pressure drop was gradually increasing during the test. Subsequent inspection showed that the grizzly on the ash discharge line from the bed was about 95% blocked. The MBHE performance had also deteriorated towards the end of the test. The inspection after shutdown showed that the top of the MBHE was full of agglomerates that were disrupting

solids flow through the MBHE. The balance of the test campaign was resumed in November after modifications were made to the facility, as described in Section 9.3.

The test campaign started out with operation on Pittsburgh #8 coal and Ecocal limestone. Initial solids distribution tests were made with the diverter plate during both gas warmup and coal operation. Tests were run with SNCR, multistaging, and steam injection to look at their impact on NO_x and CO emissions. The cyclone efficiency was artificially reduced and the switch was made to the finer Marblewhite limestone to look at sulfur capture improvements in the NID system.

Table 9-1 and Table 9-2 show the coal and ash analyses for the Pittsburgh #8 coal. Table 9-3 shows the ASTM Ash Fusibility Temperatures for this coal.

Table 9-1: Pittsburgh #8 Coal Analysis

Component	Wt % - Pittsburgh #8 Coal
C	70.0
H	4.9
O	6.2
N	1.2
S	3.7
Ash	11.6
Moisture	2.4
HHV (Btu/lb)	12,783

Table 9-2: Pittsburgh #8 Coal Ash Analysis

Component	Wt % - Pittsburgh #8 Coal
SiO ₂	43.5
Al ₂ O ₃	20.3
Fe ₂ O ₃	22.3
CaO	2.8
MgO	0.8
Na ₂ O	0.6
K ₂ O	1.6
TiO ₂	0.9
P ₂ O ₅	0.2
SO ₃	2.4
MnO	2.8
BaO	0.1
SrO	0.1

Table 9-3: ASTM Ash Fusibility for Pittsburgh #8 Coal

	Pittsburgh #8 Coal
I.T	2140 °F
S.T	2185 °F
H.T	2240 °F
F.T	2290 °F

Initial test results from the September test campaign showed that:

- The new MBHE solids distributor provided improved control over both solids flow rate and solids flow distribution. The solids flow rate could be controlled by varying either the on or off time for the pulse air to each solids distribution hopper and also by varying the number of hoppers receiving the pulse air. The maximum solids flow rate with four (out of sixteen) hoppers simultaneously receiving the pulse air was 16,000 lb/hr.
- Varying methods were tested for controlling NO_x and CO emissions. These techniques included varying primary air/secondary air ratio, single vs. multiple overfire air injection points, excess air level, steam injection, and SNCR. NO_x emissions were reduced to less than 90ppm without SNCR, while staging with two overfire air levels and SNCR injection further reduced the NO_x to 45ppm. No ammonia slip was detected during these tests.
- A test was also conducted to minimize CO emissions by injecting steam with the overfire air at two locations. The objective of this test was to stage the lower furnace for NO_x reduction and then use the steam to both enhance freeboard mixing and to generate OH radicals for enhanced CO burnout. There was a modest reduction in CO emissions (about 50 ppm) with the steam injection, but the overall CO levels were still about 250ppm during this test.
- All sulfur removal tests were conducted at a constant Ca/S mole ratio of 2. The sulfur removal in the NID system was again impacted by the limited amount of sorbent that was able to pass through the cyclone. Attempts were made to make the cyclone less efficient to bypass more of the sorbent over to the cyclone. The NID system showed an improvement during this period, but the furnace temperature profile changed significantly due to disruptions in the cyclone ash recycle rate. The tests were thus stopped before the NID system could stabilize. Sulfur removal with the coarser Eco-Cal limestone was limited to around 90%.
- Overall sulfur capture performance dramatically improved after we switched to the finer Marblewhite sorbent. This limestone was fine enough that most of it could pass through the cyclone on the first pass and reach the NID system. Sulfur removals of 98% were achieved with this sorbent at a Ca/S of 2, with about 1/3 of the capture in the combustor and the balance in the NID system.
- Tests were also conducted to determine if the MTF could be used for future gas-to-solids heat transfer scaling tests. The objective of the test was to intentionally disrupt the bauxite distribution entering at the top of the furnace and then to measure the resulting temperature. Furnace temperature probing verified that disrupting the bauxite distribution in the upper combustor would impact the gas temperature profile.

This confirmed predictions made with the CFD model of the MTF with solids maldistribution.

9.3. MTF TEST 2 – November 2003

Objective

The objective of this test series was to complete the test matrix begun in the September test campaign and to also address some issues arising from that campaign. There were several specific objectives for this campaign:

- Demonstrate MBHE heat transfer performance and MBHE solids flow control without the agglomeration issues from the prior campaign.
- Continue the NO_x/CO reduction tests with two new fuels
- Continue sulfur performance evaluation tests with a lower humidity ratio in the NID™ system
- Repeat the solids diverter tests

Modifications

The MTF tests in September 2003 were suspended prematurely after a pond that receives cooling water discharge reached its maximum allowable level. We also had some flow problems in the MBHE that required modifications. The modifications were completed in October. The completion of the test campaign was postponed until early November due to vendor delays in delivering the coal for the test.

MBHE Vent System

The photo on the left side of Figure 9-7 shows the agglomerates that formed at the top of the MBHE at the end of the September test campaign. These agglomerates were attributed to localized combustion at the top of the MBHE after the MBHE pulse air bypassed a plugged vent and passed up through the MBHE. The photo on the right side of Figure 9-7 shows the condition of the MBHE after cleaning just prior to the start of the November test campaign.

Figure 9-7: MBHE Before and After Cleaning

The first modification was improving the vent system in the CMB Moving Bed Heat Exchanger. The vent was necessary to direct transport air from the active solids distribution grid to the vent at the top of the heat exchanger. The original vent was undersized and prone to pluggage. The revised vent shown in Figure 9-8 consisted of 2 - 2" standpipes connecting the plenum below the active grid to two rakes above the grid. The rakes above the grid consisted of 2 - 2" pipes with a $\frac{1}{2}$ " wide slit on the bottom to withdraw air from above the grid. These rakes also had a shrouded opening to receive the end of the 2 standpipes. The rakes were inclined 10° from horizontal to minimize solids pickup. The rakes exited the door of the heat exchanger, turned upward and reentered the heat exchanger at the top fill port.

Figure 9-8: Revised MBHE Vent System

Bauxite Overflow Standpipe

A second modification was the addition of a shroud around the bauxite overflow standpipe in the combustor to prevent burning coal particles floating on the top of the bauxite bed from entering the grizzly on the standpipe and then being transported to the MBHE. The shroud was open at the bottom to let bauxite flow upward to the top of the standpipe. The shroud was also open at the top to vent fluidizing air carrying bauxite to the overflow pipe. A cover plate prevented bauxite and coal particles from falling into the vent pipe from the top of the combustor. The shroud was sufficiently high to extend beyond the top of the bauxite bed in the combustor. Figure 9-9 shows the original downcomer design, a schematic of the downcomer with the shroud, and the downcomer with the shroud installed.

Figure 9-9: Grizzly Before and After Modification

Test Matrix

Figure 9-10 outlines the test matrix that was planned for the November test campaign. Many of the tests had multiple objectives, as the MBHE, combustor, and NID performance could all be tested simultaneously.

Only a portion of the original test matrix was completed in the September test campaign because of difficulties with the cooling water pond level and MBHE flow. The balance of the test matrix was planned for the November test campaign.

Figure 9-10: CMB Test Matrix for 2nd Test Campaign

Operational Summary

Tests were completed with a West Virginia bituminous coal (3.4% sulfur) and a petroleum coke (5.1% sulfur). Two limestones were tested, Eco-Cal 5050 and Marble White 325. The two limestones came from the same quarry but had different particle sizes, with the Eco-Cal 5050 being the coarser one. The limestones were fed pneumatically with the injection point above the bed. This was a change from the September test, where the limestone was fed with the coal directly above the bed. Test variations included sorbent size distribution, SNCR injection, and steam injection. Highlights for the test campaign included:

1. The MBHE operated very smoothly during this campaign, with steady and reliable solids flow through the heat exchanger.
2. The MBHE heat transfer rates were sustained at a very high level throughout the test.
3. There were no signs of any agglomeration or blockage in the upper MBHE as experienced at the end of the September test campaign.

This MBHE improvement was attributed to two successful modifications.

1. The MBHE vent was modified to improve the venting of the pulse air from the individual hoppers in the MBHE. Before, the vent was frequently plugged and the pulse air passed up the MBHE. This may have caused localized combustion at the top of the heat exchanger.
2. A shroud was added in the combustor around the downcomer grizzly. There was a concern that burning coal particles floating on the bed surface were getting pulled down into the downcomer and were then transported still burning to the MBHE. The

shroud forced the particles to move down into the bed first and then up into the downcomer. The TV camera at the top of the MBHE showed a clear reduction in glowing (burning coal) particles entering the top of the MBHE.

Figure 9-11 shows the condition of the MBHE during the inspection after the conclusion of the November test campaign. The MBHE was very clean without any signs of any agglomerates.

Figure 9-11: MBHE After November Test Campaign

Three sets of in-furnace gas traverses were completed for O_2 , CO, NO_x , SO_2 , and N_2O . The traverses were completed at 2 different elevations - two sets without SNCR and one set with SNCR. The traverses showed that highly reducing conditions exist on the south side of the furnace where the fuel is being introduced. This was not surprising, but the large shroud around the downcomer pipe may have exacerbated the reducing conditions. The shroud blocked a large area of the bed and may have inhibited mixing of the fuel across the bed. The poor mixing in the lower furnace may also have made it more difficult to control the CO and NO_x emissions.

Six sets of in-furnace temperature traverses were completed at three different elevations. Each elevation was traversed with the bauxite diverter damper in the vertical position (no effect) and set at 18 degrees from the vertical (which diverted the bauxite towards the west wall). These results were used as validation data for CFD modelling of the combustor (see Section 10.2).

Combustion performance was very good with both fuels. The flyash samples with the West Virginia bituminous coal and the petroleum coke both visually looked low in carbon. The combustion behavior of the petroleum coke was initially surprising - it had completely different burning characteristics from the bituminous coals because it was so low in volatiles. This was because the petroleum coke was initially fed above the bed at a temperature that was too low for good combustion. The fuel burned great once the operating conditions were modified to increase the local temperature at the fuel feed point. The transition to petroleum coke firing would have been simplified if the fuel was fed directly into the bed.

Various combinations of staging, multi-staging, SNCR, and steam injection were tested to evaluate their impact on NO_x and CO emissions. NO_x was generally around 100-120ppm without SNCR and as low as 60ppm with SNCR for both fuels. CO was typically 140-

300ppm with the West Virginia coal and consistently around 120ppm with the petroleum coke. N_2O varied from 0 to 30 ppm depending upon conditions, but was generally less than 10 ppm.

The in-furnace and NID sulfur removal performance results are summarized in Table 9-4. The sulfur capture with the coarser Eco-Cal limestone was again impacted by difficulties in getting the sorbent through the cyclone and into the NID system. This significantly reduced the backend sulfur capture performance. NID performance was greatly improved with the Marblewhite sorbent, which was fine enough to pass through the cyclone and reach the NID system. Generally, the in-furnace sulfur capture was about 30% with the coarser sorbent and 40-48% with the finer sorbent. The best NID performance with the Eco-Cal was almost 90%, giving a total sulfur capture of 92% at a Ca/S mole ratio of 2. The Marblewhite performance was quite good. The NID system captured virtually all of the remaining SO_2 with the West Virginia coal, giving almost 100% sulfur capture at a Ca/S mole ratio of 2. The performance was also very good with the petroleum coke, with 98% total sulfur capture at a Ca/S mole ratio of 2.

Table 9-4: Summary of Sulfur Capture Performance

Fuel	WV Bituminous	WV Bituminous	Pet Coke
Sorbent	Eco-cal	Marblewhite	Marblewhite
Sulfur Removal			
In-Furnace	31.8%	47.6%	39.4%
FDA	87.7%	99.2%	96.8%
Overall	91.6%	99.6%	98.1%

Figure 9-12 through Figure 9-17 show the temperature traverses taken on November 3rd and 6th at three different elevations in the furnace as the solids diverter plate was moved to different positions.

Figure 9-12: Temperature Traverses on November 3rd at Level 34

Figure 9-13: Temperature Traverses on November 3rd at Level 42

Figure 9-14: Temperature Traverses on November 3rd at Level 52

Figure 9-15: Temperature Traverses on November 6th at Level 34

Figure 9-16: Temperature Traverses on November 6th at Level 42

Figure 9-17: Temperature Traverses on November 6th at Level 52

Table 9-5 and Table 9-6 show the coal and ash analyses for the West Virginia Waynesburg coal. Table 9-7 shows the ASTM Ash Fusibility Temperatures for this coal.

Table 9-5: Waynesburg Coal Analysis

Component	Wt % - Waynesburg Coal
C	64.7
H	4.5
O	5.6
N	1.3
S	2.2
Ash	17.1
Moisture	4.6
HHV (Btu/lb)	11,083

Table 9-6: Waynesburg Coal Ash Analysis

Component	Wt % - Waynesburg Coal
SiO ₂	54.8
Al ₂ O ₃	25.6
Fe ₂ O ₃	12.4
CaO	0.8
MgO	0.7
Na ₂ O	0.4
K ₂ O	2.2
TiO ₂	1.3
SO ₃	0.1
Other	1.1

Table 9-7: ASTM Ash Fusibility for Waynesburg Coal

	Waynesburg Coal
I.T	2410°F
S.T	2465°F
H.T	2552°F
F.T	2643°F

Table 9-8 and Table 9-9 show the coal and ash analyses for the Billings shot coke. Table 9-10 shows the ASTM Ash Fusibility Temperatures for this coal.

Table 9-8: Billings Shot Coke Analysis

Component	Wt % - Billings Shot Coke
C	85.5
H	3.6
O	0.0
N	1.5
S	5.4
Ash	0.7
Moisture	3.3
HHV (Btu/lb)	14,523

Table 9-9: Billings Shot Coke Ash Analysis

Component	Wt % - Billings Shot Coke
SiO ₂	47.7
Al ₂ O ₃	19.9
Fe ₂ O ₃	22.9
CaO	3.0
MgO	0.7
Na ₂ O	0.6
K ₂ O	2.1
TiO ₂	1.0
P ₂ O ₅	0.5
SO ₃	1.5
MnO	<0.1
BaO	<0.0
SrO	0.1

Table 9-10: ASTM Ash Fusibility for Billings Shot Coke

	Billings Shot Coke
I.T	2140°F
S.T	2185°F
H.T	2240°F
F.T	2290°F

9.4. MTF Test 3 (Scaled MTF Gas Heat Transfer Tests)

Objective

The primary objective of this test campaign was to determine the heat transfer coefficient of the falling solids to the rising gas. Previous test campaigns were generally conducted to demonstrate the CMB proof of concept rather than to collect heat transfer data. This test campaign was designed specifically to collect detailed gas-to-solids heat transfer data with optimum gas and solids flow distributions.

A second objective was to collect detailed validation data to calibrate a CFD model of the CMB process. The CFD model was being used as a scale up tool. Previous test data did not give enough information on the flow field to determine if the CFD model was predicting the correct flow and temperature pattern. Some of the results from the CFD models predicted undesirable flow fields, which could be a problem in larger units.

A third objective of the test was to demonstrate a new measuring device to determine the velocity and density of the flow field using a laser probe designed by MSE Meile. The probe had previously been tested at Power Plant Laboratories for ambient temperature tests of slower moving solids, as in a moving bed or down-comer in the chemical looping process. This test involved hotter solids in free fall conditions, which was a more challenging application.

Modifications

Multi Use Test Facility

Scaling procedures were used to modify the MTF to simulate a much larger commercial-sized CMB combustor and to design the test experiments. The optic probe selected for measuring solids concentration and velocity was limited to operating temperatures less than 700°F. The tests were therefore designed for hot air operation only and with no combustion taking place within the combustor.

The Multi Use Test Facility (MTF) combustor was modified to measure heat transfer from a rising hot gas to a cloud of falling scaled particles. The scaling procedure developed by Professor Glicksman required a reduced combustor height. 700-micron bauxite particles were supplied from a 5" inlet pipe to a basket distributor 6 feet below the bottom of the cyclone inlet duct. This provided a dispersed flow of bauxite to the rising combustor gas stream. The supply height was lowered in order to simulate a shorter combustor. A full range of test data including pressures, temperatures, and flows were logged into the data collection system. In addition, test probes were used to get detailed temperature traverse measurements at several locations in the MTF. The port openings were at elevations 34, 42, and 52 feet. Data was taken from two perpendicular directions, north and west, to get a better understanding of the flow field. Traverses were also made with the thermocouple probe and with the Meile laser probe to measure particle velocity and density. The thermocouple rakes and solids optic traverse planes were located at elevations that simulated a revised combustor bottom, effectively making the height of the scaled combustor equal to 20 ft.

The temperature of the gas rising in the combustor was measured by a series of thermocouple rakes. The rakes consisted of inverted angle shields that prevented falling bauxite from impacting the thermocouples. A series of thermocouples were mounted under the shields to measure the gas temperature at several radial locations. The position of each

rake is indicated in Figure 9-18. A vertical thermocouple rake located in the center of the cyclone duct measured gas temperatures exiting the combustor.

Figure 9-18: MTF CMB Scale-up Combustor Instrumentation

Heat flow from the hot gas to the bauxite particles was determined by the average difference of successive thermocouple rakes at different elevations in the combustor. The heat flow, bauxite inlet temperature, and bauxite specific heat was used to calculate a bauxite temperature profile in the combustor. A heat balance method was used to calculate the bauxite flow rate based on gas temperature increases and flow, and bauxite temperature changes in the combustor.

Average plug flow bauxite particle density in the combustor was calculated by the difference between calculated particle terminal velocity, gas rise velocity, bauxite mass flow rate, and the combustor flow cross sectional area. The density, mass flow, and particle diameter were used to calculate the particle surface area in the combustor. This surface area, the gas side heat flow, and the log mean temperature difference between gas and solids were used to calculate a particle surface heat transfer coefficient.

Meili Labasys 100 Probe

A Meili Labasys 100 probe was used for the simultaneous determination of local solids concentration and velocity. It consisted of a measuring probe, which integrated all optical and electronic components and protected them against environmental harms, the power supply and the personal computer based data evaluation unit. This device used a pair of laser beams shining into the flow field to determine the concentration of particles in the gas at specific locations. The comparison of several signals allowed the correlative calculation of the solids velocity also.

The original Meile probe was not suitable for high temperature measurements and could not be used as designed for the MTF test program. Meili subsequently modified the original probe to enable operation at temperatures up to 622°F. The existing probe tip was refitted with newly developed high temperature sensor heads, which had a confined measuring volume without distracting additional optics, which was especially problematic for the planned use at elevated temperatures. The sensor heads were designed to endure temperatures up to 622°F. To ensure the overall optical stability of the probe tip, long-term tests in a high temperature oven were conducted.

Figure 9-19 shows the revised Meili probe tip. It used a 3 channel probe tip that included 2 normal sensor heads for measuring solids velocity, a "NT" sensor heat for accurate determination of solids concentration, and a type K thermocouple shielded against falling solids. Protective caps (not shown) were screwed onto the heads to protect the sensitive sensor heads.

Figure 9-19: Meili Probe Tip and Sensors

Figure 9-20 shows the Meili probe and support rail as the instrument is being introduced into the MTF.

Figure 9-20: Meili Probe and Support Rail

Figure 9-21 shows a schematic of ALSTOM Power's Multi-Use test facility (MTF), which was operated in gas mode only – i.e. the unit was operated without any combustion. Instead the air was heated up in an external heater and then introduced to the bottom of the combustor. The air temperature at the inlet was limited to 600°F to protect the un-cooled Labasys probe tip. The probe measuring positions at the MTF unit are indicated in Figure 9-21. The probe was used on 4 different levels and from 2 sides, with 7 different positions measured at each location.

Figure 9-21: Meili Probe Measurement Locations

Table 9-11 shows the nominal MTF operating conditions for each of the test conditions where the Meili probe was used to measure solids concentration and velocity.

Table 9-11: Nominal Operating Conditions for Meili Probe Measurements

Test		'1'	'2'	'3'	'4'	'5'	'6'
Measuring Program							
Date:		Dec, 6/7th, 2004	Dec, 8th, 2004	Dec, 8/9th, 2004	Dec, 9th, 2004	Dec, 9th, 2004	Dec, 9th, 2004
Runs:		1-10	18-24	25-31	11-17	32-38	39-46
Measuring Positions:		7 different inlets on 4 levels: 11.4, 13.5, 15.1, 17 m; at each 7 radial posit.: 0.495 (5 cm from wall), 0.437, 0.369, 0.286, 0.165, 0.0 (at centre), -0.165 m					
Data Sampling:		500'000 pts at 12 kHz --> measuring period: 41.6 s					
Number of repetitive measurements:		2 for all positions --> 2x46x7=644 measur. à 500'000 pts (1.8 GB raw data)					
Operating Conditions							
Gas temperatures:		Inlet: 600 °F / 315 °C, Outlet: 300 °F / 149 °C, Average: 450 °F / 232 °C					
Average Gas Density:		0.0436 lb/ft ³ / 0.698 kg/m ³					
Superficial Gas Velocity:	[ft/s]	3.3	1.49	1.49	3.3	3.3	3.3
	[m/s]	1.01	0.45	0.45	1.01	1.01	1.01
Air feed rate:	[lb/h]	4520	2036	2036	4520	4520	4520
	[kg/h]	2050	924	924	2050	2050	2050
Solids feed rate:	[lb/h]	10000	10000	5000	5000	2500	10000
	[kg/h]	4536	4536	2268	2268	1134	4536
Solids loading:	[-]	2.21	4.91	2.46	1.11	0.55	2.21
Mean gas velocity / terminal velocity v_{mean} / v_t :	[-]	0.66	0.30	0.30	0.66	0.66	0.66

Test Matrix

Figure 9-22 shows the planned test matrix for the gas-to-solids scaled heat transfer tests. There were six main test conditions with detailed probing measurements at multiple elevations for temperature, solids concentration, and solids velocity.

Figure 9-22: MTF Heat Transfer Scaling Test Matrix

The test conditions were selected by a design of experiments based on the scaling criteria and the necessity of obtaining a range of data that would bracket future operating requirements. The detailed test matrix is given in Table 9-12.

Table 9-12: MTF CMB Test Matrix

CMB Scale Particle Test Matrix 11/19/04							
Test		1	2	3	4	5	6
Dc	in	40	40	40	40	40	40
Ac	ft ²	8.73	8.73	8.73	8.73	8.73	8.73
Dp min	μ	180	180	180	180	180	180
Vt600	fps	4.8	4.8	4.8	4.8	4.8	4.8
Vt300	fps	5.1	5.1	5.1	5.1	5.1	5.1
Test Ratio		0.666	0.3	0.3	0.666	0.666	0.666
Vtest	fps	3.30	1.49	1.49	3.30	3.30	3.30
Tgo	F	300	300	300	300	300	300
Tgi	F	600	600	600	600	600	600
Tavg	F	450	450	450	450	450	450
ρ_o	lb/ft ³	0.0523	0.0523	0.0523	0.0523	0.0523	0.0523
ρ_i	lb/ft ³	0.0375	0.0375	0.0375	0.0375	0.0375	0.0375
ρ_{avg}	lb/ft ³	0.0436	0.0436	0.0436	0.0436	0.0436	0.0436
Wg	lb/h	4,520	2,036	2,036	4,520	4,520	4,520
Vo	fps	2.75	1.24	1.24	2.75	2.75	2.75
Vi	fps	3.84	1.73	1.73	3.84	3.84	3.84
Vo/Vt		0.57	0.26	0.26	0.57	0.57	0.57
Vi/Vt		0.75	0.34	0.34	0.75	0.75	0.75
Ws	lb/h	10,000	10,000	5,000	5,000	2,500	10,000
Ws/Wg		2.21	4.91	2.46	1.11	0.55	2.21
Tk	K	506	506	506	506	506	506
μ_{ts}	lb/ft/hr	6.51E-02	6.51E-02	6.51E-02	6.51E-02	6.51E-02	6.51E-02
Re		2.65E+04	1.19E+04	1.19E+04	2.65E+04	2.65E+04	2.65E+04
Vcl ratio		1.25	1.28	1.28	1.25	1.25	1.25
Vclo/Vt		0.72	0.33	0.33	0.72	0.72	0.72
Vcli/Vt		0.94	0.43	0.43	0.94	0.94	0.94
Vmig avg c	fps	2.35	3.86	3.86	2.35	2.35	2.35
Vmig avg i	fps	0.96	3.07	3.07	0.96	0.96	0.96
Vmig cl o	fps	1.67	3.51	3.51	1.67	1.67	1.67
Vmig cl i	fps	0.01	2.59	2.59	0.01	0.01	0.01
ρ_{so}	lb/ft ³	0.14	0.08	0.04	0.07	0.03	0.14
ρ_{si}	lb/ft ³	0.33	0.10	0.05	0.17	0.08	0.33
$\rho_{scl o}$	lb/ft ³	0.19	0.09	0.05	0.10	0.05	0.19
$\rho_{scl i}$	lb/ft ³	31.81	0.12	0.06	15.91	7.95	31.81
Qin	Btu/h	6.22E+05	2.80E+05	2.80E+05	6.22E+05	6.22E+05	6.22E+05
Qbauxite	Btu/h	3.39E+05	1.53E+05	1.53E+05	3.39E+05	3.39E+05	3.39E+05

Operational Summary

Six series of test data were taken during this test campaign. Each of the six series was taken as close to constant conditions as possible. This was difficult because of the time needed to make all of the data traverses. The six test conditions were chosen to show a combination of gas and solids flow rates that are considered for the CMB process. The actual flow conditions are given in Table 9-13. They include solids to gas ratios from 0.5 to 4 #solid/#gas.

Table 9-13: CMB Heat Transfer Test Conditions

Test series	Solids Flow	Solids T	Gas Flow	Gas T	S/G
1	8905 #/hr	228F	4599 #/hr	483F	1.94
2	9330	287	2184	589	4.27
3	4542	210	2282	648	1.99
4	5438	184	4537	480	1.20
5	2428	110	4412	409	0.55
6	9354	245	4598	462	2.03

Each of the six test series included two parallel test sequences; heat transfer from the flue gas to the raining bauxite, and optic probe traverses at specified elevations in the combustor. The tests ran smoothly, although two problems were encountered with the optic probe. First, the probe was too short to completely traverse from wall to wall in the combustor. Second, the probe sequences were quite time consuming and it was difficult to maintain the MTF facility at steady state conditions during certain lengthy traverses.

Figure 9-23 shows the variations in the gas and bauxite flow rate during the test period. Figure 9-24 shows the variations in gas heat flow during the same time period.

Figure 9-23: Gas and Bauxite Flowrate vs. Operating Time

Figure 9-24: Gas Heat Flow vs. Operating Time

9.5. MTF Test 4 (CFB O₂-Fired)

Objective

The objective of this test campaign was to collect moving bed heat transfer data with the Multi Use Test Facility operating in CFB mode rather than CMB mode. This would allow the impact of solids media (CFB ash vs. bauxite) on heat transfer to be assessed. These tests were run in parallel with an oxy-fired CFB test campaign that was run independent of the CMB program, but which used a MBHE for combustor temperature control.

Modifications

Figure 9-25 shows sectional views of the MBHE that was used for this test. It was installed in the MTF in parallel with a fluidized bed heat exchanger used for conventional CFB firing. Solids were supplied to the top nozzle of the MBHE by a sideslip stream from the MTF cyclone seal pot. A rotary valve at the bottom controlled solids flow through the MBHE. The MBHE was previously fabricated and installed in the Multi Use Test Facility during the CMB Phase I program to evaluate moving bed heat transfer performance.

The MBHE consisted of two tube bundles shown in Figure 9-25, the upper having 4 tubes in depth and the lower having 6 tubes in depth. Each bundle was seven tubes wide. The bundles consisted of 1.5" OD tubes with 0.5" high by 0.06" thick circumferential fins on 0.5" spacing.

Instruments were installed on the MBHE to measure its performance. The primary instruments were type K thermocouples to measure solids inlet and outlet temperatures and the tube bundle coolant inlet and outlet temperatures. In addition, two turbine meters measured cooling water flow in each bundle. All instruments were read by the MTF Labview data acquisition system and stored on the PPL server network for later analysis.

Figure 9-25: MBHE Sectional Views

The tube bundles were cooled by MDC water of high purity. A precision turbine meter measured the water flow rate. The MBHE heat transfer rate was determined for each bundle, using the cooling water as a heat flow medium. The inlet temperature of the coolant

was measured by two inlet header thermocouples. The outlet temperature of each tube pass of a tube bundle was measured by a thermocouple inserted axially through the header, into the outlet of the tube to a depth of 8". A boundary layer trip ring was installed in each tube outlet, upstream from the thermocouple to provide a mixed fluid temperature. The average heat flow to a bundle was calculated from the average inlet-outlet temperature difference, coolant flow rate, and coolant specific heat.

The solids flow rate through the heat exchanger was calculated from the solids average inlet-outlet temperature difference, MBHE heat flow, and solids specific heat. The rotary valve could also have been used as a solids flow meter, but it was uncalibrated. Also, rotary valves are volumetric devices and the pockets in the valve may not be full under some circumstances. The heat balance method of calculating solids flow rates was preferred because of this.

An intermediate solids temperature between tube bundles was calculated from the average solids inlet temperature, top tube bundle heat transfer, calculated solids flow rate, and solids specific heat. This solids temperature was calculated because the spacing between top and bottom bundles was too low for an accurate temperature measurement by thermocouples.

Operating Results

The average inlet (T_{si} avg) and outlet solids temperatures (T_{so} avg) for this test campaign are shown in Figure 9-26. The calculated solids temperature between the upper and lower tube bundles (T_{smu} calc) is also shown. Measured test data is indicated by lines on these figures. Predicted conditions are indicated by the symbols.

Figure 9-26: MBHE Average Solids Temperature

The heat flow to the coolant in the upper (Qu avg) and lower tube bundles (Ql avg) is shown in Figure 9-27. Also shown in this figure are bars indicating operating conditions in the MTF. Finally, the calculated solids flow (Ws avg) through the MBHE is shown in Figure 9-28. Also shown in this figure is the rotary valve speed (RVSPDMBHE) that controlled the solids flow rate through the MBHE.

Figure 9-27: MBHE Tube Bundle Heat Flows

Figure 9-28: MBHE Solids Flow Rate

The difference between measured and predicted heat flow does not significantly vary from test to test. This indicates that oxy firing vs. air firing, coal vs. pet coke firing, and limestone variation does not have a significant influence on MBHE heat transfer performance.

10. DATA ANALYSIS

The objective of this task was to review and analyse the results of the three MTF test campaigns. The analysis covered the areas of heat transfer, combustion performance, and emissions.

10.1. Gas-to-Solids Heat Transfer

A major objective of the scaled MTF gas heat transfer tests (see Section 9.4) was to determine the heat transfer coefficients of the falling solids to the rising gas. In particular, this campaign was designed specifically to collect detailed gas-to-solids heat transfer data with optimum gas and solids flow distributions. A second objective was to collect detailed validation data to calibrate a CFD model of the CMB process. This model would then be used to scale up gas flow and temperature profile predictions for large-scale commercial CMB combustors. A third objective was to demonstrate the ability of the MSE Meile laser probe to measure both solids concentration and solids velocity. This section describes the results of the first and third objectives, while results of the second objective are reported in Section 10.2.

Background

Previous MTF heat transfer tests and CFD modelling of the rising flue gas in the combustor to the falling bauxite particles clearly showed the need for a scale-up methodology for design of commercial-sized units. A 250MW_e CMB combustor would have a 40 ft diameter, which is a 144x scaleup in cross-sectional area compared to the MTF.

CFD models of the CMB combustor were being developed as a scale-up tool to predict bauxite particle distribution, temperature profiles, and gas-to-solid heat transfer performance. The models needed benchmark data from the scale-up tests as well as from previous gas-to-solids heat transfer tests. Benchmarking the CFD program required additional information about the bauxite distribution in the combustor beyond the average data obtained in previous tests. An optic probe described in Section 9.4 was used to measure the bauxite concentration profiles and solids migration velocity profiles at four elevations in the combustor. This data was then used to validate the CFD model and to make performance predictions for a commercial sized CMB combustor.

The initial scale-up approach was to use a 10 ft diameter heat transfer test facility having a plan area 9 times larger than the MTF. The larger 250MW_e commercial unit would then require a more manageable scale-up of 16x the cross-sectional area of this proposed larger test facility. One concept of the scale-up facility is shown in Figure 10-1.

Figure 10-1: Proposed Initial CMB Scale-up Heat Transfer Test Facility

The costs of the 10 ft column were excessive, so an alternative scaled modelling approach was used instead that reduced the bauxite particle size rather than increased the column size. This permitted the use of the MTF for the scale-up tests. A scaling procedure suggested by Professor Leon Glicksman¹⁴ was used to select a bauxite particle size for use in the MTF that would provide data for scale-up. This scaling criteria applied to various geometric configurations is shown in Table 10-1.

¹⁴ *Simplified Scaling Relationships for Fluidized Beds*, L. R. Glicksman, M. Hyre, and K Woloshun, Powder Technology, pp 177-199, 1993

Table 10-1: Glicksman Scaling Criteria

CMB Glicksman Scaling													
Parameter		MTF(test)	Commercial	MTF(test)	Commercial	MTF(test)	Commercial	MTF(test)	Commercial	BSF(test)	Commercial		
Combustor Diameter	inches	40	40	40	279	40	862	40	77	96	394		
Combustor Diameter	ft	3.3	3.3	3.3	23.2	3.3	71.9	3.3	6.4	8.0	32.9		
Combustor Height	ft	25	25	14.4	100	5	100	52	100	25	100		
U _o	ft/s	30	30	5.6	14.7	3.2	14.0	10.6	14.7	7.3	14.7		
Particle diameter	mm	0.7	0.7	0.32	0.7	0.22	0.7	0.53	0.7	0.4	0.7		
Terminal velocity UT	m/s	10.5	10.5	2.93	7.72	1.67	7.74	5.58	7.72	3.81	7.72		
Terminal velocity UT β	ft/s	25.3	25.3	9.6	25.3	5.47	25.4	18.3	25.3	12.5	25.3		
Minimum Fluidization velocity U _{mf} β	fps	1.12	1.12	0.16	0.69	0.0773	0.68	0.392	0.69	0.26	0.69		
Minimum Fluidization velocity U _{mf}	m/s	1.0	1.0	0.0	1	0.0	1	0.1	1	0.1	1		
UT/U _o		1.72	1.72	1.72	1.72	1.72	1.72	1.72	1.72	1.7	1.72		
UT/U _{mf}		17.2	10.5	17.2	10.5	17.2	10.5	17.2	10.5	17.2	10.5		
Cluster diameter (assumed)	mm	2.5	10	2.5	10	2.5	10	2.5	10	2.5	10		
UT(cluster) / U _o		8.5	10	8.5	10	8.5	10	8.5	10	8.5	10		
UT(cluster)		255	300	47	147	27	148	90	147	62	147		
Gas Temperature	F	2000	2000	2000	2000	2000	2000	2000	2000	2000	2000		
C _p		0.26	0.26	0.26	0.26	0.26	0.26	0.26	0.26	0.26	0.26		
ρ particle		224	224	224	224	224	224	224	224	224	224		
Glicksman Nu		3.51	5.88	3.51	5.88	3.51	5.88	3.51	5.88	3.51	5.88		
T		1366.6667	1366.667	1366.667	1366.667	1366.667	1366.667	1366.667	1366.667	1366.667	1366.667		
μ gas		3.35E-05	3.35E-05	3.35E-05	3.35E-05	3.35E-05	3.35E-05	3.35E-05	3.35E-05	3.35E-05	3.35E-05		
ρ gas		0.0161442	0.016144	0.016144	0.016144	0.016144	0.0161442	0.016144	0.016144	0.016144	0.0161442		
Re		2.80E+01	2.80E+01	4.86E+00	2.80E+01	1.90E+00	2.81E+01	1.53E+01	2.80E+01	7.90E+00	2.80E+01		
PPL Nu		4.27E+00	4.27E+00	2.34E+00	4.27E+00	1.83E+00	4.28E+00	3.41E+00	4.27E+00	2.72E+00	4.27E+00		
τ transit		5.62	5.62	1.000	3.58	9.41	0.381	2.18	9.41	0.232	6.79	9.43	0.507
τ thermal		0.115	0.115	1.000	0.044	0.115	0.381	0.026	0.115	0.231	0.082	0.115	0.512
Cross Section Area	ft ²	8.7	8.7	8.7	423.1	8.7	4057.3	8.7	32.0	50.3	847.8		
Gas Flow	lb/h	15,216	15,216	2,831	361,483	1,613	3,482,236	5,396	27,428	21,231	725,740		
Solids Flow	lb/h	30,431	30,431	5,662	722,965	3,226	6,964,472	10,792	54,856	42,462	1,451,480		

Scaled Heat Transfer Tests

Six series of test data were taken during the scaled heat transfer tests (as described in Section 9.4). The solids-to-gas ratio was varied from 0.5 to 4.0 during the tests. Detailed temperature, solids concentration, and solids velocity profiles were collected for each test.

Figure 10-2 shows the combustor centerline temperatures at four different elevations over the duration of the MTF scaled heat transfer test series. The heavy red line corresponds to the time period where detailed temperature, concentration, and velocity profiles were measured for the six test series.

Figure 10-2: Combustor Centerline Temperatures

Figure 10-3 shows the gas and bauxite temperature profile along the height of the combustor for each of the six test series. The profiles illustrate how the heat transfer performance was reduced during the higher solids loading test due to solids clustering and gas bypassing. Test 5 had a low solids/gas ratio (0.55) and relatively good heat transfer performance. The bauxite particles were heated up from 120°F to 355°F while the gas was cooled from 430°F to 300°F. Test 2 had a high solids/gas ratio (4.27) and worse heat transfer performance. The bauxite particles were heated up from 175°F to 260°F while the gas was cooled from 445°F to 335°F. One would have expected the gas to be cooled much more with the higher solids flow, but it's likely that the gas and solids were bypassing each other.

Figure 10-3: Average Thermocouple Rake Temperature Profile

Figure 10-4 shows the effect of the gas flow rate on the total heat transferred from the gas to the falling solids for the six test series. The overall heat pickup from the flue gas to the bauxite was roughly proportional to the gas flow, although it's likely that the gas-to-solids heat transfer coefficient was impacted by the bauxite concentration or possible clustering of the bauxite. For example, Test 2 ($\blacktriangle$) had a solids/gas ratio of 4.27 while Test 3 ($\blacksquare$) had a solids to gas ratio of 2.0. Both tests were run at a flue gas velocity of 1.5 ft/sec. Considerably more heat was transferred to the bauxite particles in Test 2, again indicative that there may have been some bypassing or clustering in Test 3 that impacted the heat pickup.

Figure 10-4: Bauxite Heat Pickup vs. Flue Gas Flow

A solids flux probe was used to grab non-isokinetic samples of the local solids flux at three elevations in the combustor. The probe was inserted into the combustor and captured falling bauxite particles for a fixed time duration. Figure 10-5 shows the measured solids weight captured at each elevation for Test 6. At the higher elevations (52ft and 42ft), there was a large spike in solids concentration near the center of the combustor and practically no solids near the wall. This was because the solids distributor released the bauxite particles at the combustor centerline and the particles were still concentrated at the center for the upper combustor elevations. By elevation 34ft, most of the solids had moved towards the wall and very few solids were detected along the centerline.¹⁵

¹⁵ MTF CMB Raining Bed Heat Transfer Test Results II, Bard Teigen, ALSTOM Power Internal Report PPL-03-CT-05, April 2003

Figure 10-5: Solids Flux Probe Measurements - Test 6

Gas-to-Solids Heat Transfer Coefficients

The heat transfer coefficient between the gas and bauxite particle surface was calculated from the total heat flow from the gas to the solids (Q), the surface area of the particles (A_s), and the Log Mean Temperature Difference (LMTD) between the particles and gas. The heat transfer coefficient is determined from the equation:

$$h = Q / A_s / \text{LMTD}$$

where:

h = gas-to-solid heat transfer coefficient

Q - heat transfer from gas to particles

A_s = particle surface area

LMTD = log mean temperature difference (based on bulk gas temperatures)

The heat flow Q was calculated by an energy balance that considered the heat given up by the gas and also accounted for any system heat losses. The surface area A_s was determined by calculating the solids terminal velocity and subtracting the average gas rise velocity to obtain the downward solids migration velocity. The solids flow rate, combustor cross-sectional flow area, and the solids migration velocity were then used to calculate a solids cloud density. This density was used to calculate the surface area of the solids in the combustor:

$$V_{\text{mig}} = V_{\text{term}} - V_{\text{gas}}$$

The solids cloud density was calculated by:

$$r_{\text{ts}} = W_p / A_c / V_{\text{mig}} / 3600$$

The particle surface area was then calculated from the test section solids density as:

$$A_s = 6 * \text{Vol} * 12 * 25.4 * 1000 / D_p * \rho_{\text{ts}} / \rho_{\text{ts}}$$

where:

V_{term} = Terminal Velocity of Particles (ft/s)

V_{mig} = Migration Velocity of Particles (ft/s)

A_s = Particle Surface Area (ft²)

W_p = Mass Flow of Particles (lb/h)

A_c = Combustor Cross Sectional Area (ft²)

Vol = Volume of Combustor (ft³)

D_p = Diameter of particle (micron)

ρ_{ts} = Test section solids density (ft³/lb)

ρ_p = Particle Density (ft³/lb) – Not particle bulk density

The particles and gas flow counter to each other. Heat transfer calculations for this mode of flow use a logarithmic mean temperature difference. The log mean temperature difference was calculated as:

$$\text{LMTD} = ((T_{\text{si}} - T_{\text{go}}) - (T_{\text{so}} - T_{\text{gi}})) / \ln ((T_{\text{si}} - T_{\text{go}}) / (T_{\text{so}} - T_{\text{gi}}))$$

where:

LMTD = Log Mean Temperature Difference (F)

T_{si} = Solids Inlet Temperature (F)

T_{so} = Solids Outlet Temperature (F)

T_{go} = Gas Outlet Temperature (F)

T_{gi} = Gas Inlet Temperature (F)

The solids outlet temperature (T_{so}) was not measured. The solids outlet temperature was calculated from the bauxite flow rate, bauxite inlet temperature, bauxite specific heat, and gas flow heat flow change from the thermocouple rakes shown in Figure 10-3 as:

$$T_{so} = T_{si} + Q_{gas}/Cp_s/W_s$$

and:

$$Q_{gas} = W_{g+a} * Cp_g * (T_{gi} - T_{go})$$

where:

T_{si} = Measured Solids Inlet Temperature

Q_{gas} = Calculated Gas Flow (lb/h)

W_{g+a} = Gas Mass Flow + Recirculated Ash Mass Flow (lb/h)

Cp_g = Calculated Specific Heat of Gas + Ash (Btu/lb/F)

T_{gi} = Measured Gas Inlet Temperature from a Selected Gas Temperature Rake (F)

T_{go} = Measured Gas Outlet Temperature from a Second Gas Temperature Rake (F)

Cp_s = Specific Heat of the Bauxite Solids (Btu/lb/F)

W_s = Bauxite Mass Flow Calculated from a Heat Balance on the Bauxite Transport System lb/h)

The heat transfer coefficient between the gas and bauxite particle surface can now be calculated from the first equation ($h = Q / A_s / \text{LMTD}$). The convective surface heat transfer coefficient is usually expressed in a dimensionless Nusselt number, which in turn is expressed as a function of Reynolds Number:

$$Nu = h * D_p / k$$

where:

Nu = Nusselt number

h = Surface Convective Heat Transfer Coefficient (Btu/hr/ft²/F)

D_p = Particle Diameter (ft)

k = Thermal Conductivity of Gas (Btu/hr/ft/F)

The Nusselt number is related to the particle velocity as a function of the dimensionless Reynolds number. For our application, the Reynolds number based on particle migration velocity down the combustor provides the best relationship:

$$Nu = C_0 + C_1 * Re C_2$$

$$Re = \rho_{\text{gas}} * V_{\text{mig}} * D_p / m_{\text{gas}} \quad (9)$$

where:

C_0, C_1, C_2 = Best Fit Coefficients of the Data to the Equation

Re = Reynolds Number

ρ_{gas} = Gas Density (lb/ft³)

D_p = Particle Diameter (ft)

m_{gas} = Gas Viscosity (lb/ft-sec)

These series test results are presented in Figure 10-6 as a dimensionless Nusselt Number vs. Solids Migration Reynolds Number. Superimposed on this figure are results from an ambient temperature test series conducted at PPL in a 15" ID Octagonal column of 30' in height. Test #5 of the scaled data fit the trend line of earlier tests, while the remaining scaled test data fell significantly below the trend line. This most probably was due to the use of a single inlet solids distributor rather than multiple inlet distributors called for in commercial applications.

The heat transfer coefficient had a scattered trend with increasing rise velocity of gas and ash in the combustor. This was counter to first principle theory for a single particle in a gas stream. The heat transfer coefficient should not vary with gas velocity, but only with terminal velocity of the particle in the gas. The theoretical heat transfer coefficient for a single spherical particle falling through a gas stream is superimposed in Figure 10-6 as a dashed blue line. Experience has shown that the heat transfer coefficient does not follow first principle results, primarily due to particle-particle attraction.

This figure shows that the single particle correlation did not fit the data very well, particularly at low solids migration Reynolds Numbers or high gas velocities. The single particle correlation assumes that the particles fall freely in the gas with a relative velocity between the gas and particle equal to the particle terminal velocity. Changes in gas velocity should not affect the relative velocity or the heat transfer coefficient. However, the data indicated a reduction in heat transfer coefficient with increasing gas velocity and decreasing solids migration velocity. As the solids migration velocity decreased, the particle density increased because of reduced velocity and at a fixed mass flow. This density increase caused more interactions between particles. Particles in close proximity had a tendency to follow in each other's wake due to a drag reduction phenomena. Particles following a lead particle were exposed to gas cooled by the lead particle, and therefore did not have the full gas-particle temperature difference available to the lead particle. This accounted for the reduction in the Nusselt Number at high gas velocities or at low migration Reynolds Number.

At very low gas velocities, the downward velocity of the particles relative to the ground (i.e. V_{term} minus V_{gas}) was close to the terminal velocity. There was little solids holdup and not many particle interactions. For these cases, the observed solid-to-gas heat transfer coefficient was close to the theoretical value prediction.

Figure 10-6: CMB Scale Heat Transfer Rate vs. Particle Migration Velocity

Solids Concentration and Velocity

The MSE Meili optic particle probe was used during the scaled heat transfer test campaign to gather detailed profiles of the solids concentration and solids velocity for CFD model validation. The probe insertion closest to the bauxite distributor at the top of the combustor is shown in Figure 10-7 for tests 1 and 6. The concentration was highest near the centerline close to the distributor and near zero towards the combustor wall. The next lower elevation shown in Figure 10-8 had a more uniform distribution. Finally the lowest elevations in Figure 10-9 and Figure 10-10 showed even more uniform distributions with an increase at the combustor wall. The plug flow average bauxite concentrations calculated from terminal velocity, combustor flue gas velocity, and bauxite flow rate are shown in each figure as a smooth horizontal line. The plug flow concentrations were almost identical for tests 1 and 6.¹⁶

¹⁶ Short Report on Labasys 100 Measurement in CMB-Unit at Elevated Temperatures, Dr Reto Meili, March 2009

Figure 10-7: Optic Probe Bauxite Concentration Elev. 52

Figure 10-8: Optic Probe Bauxite Concentration Upper Elev. 42,3

Figure 10-9: Optic Probe Bauxite Concentration Lower Elev. 42,1

Figure 10-10: Optic Probe Bauxite Concentration Elev. 34

The optic probe data for bauxite particle migration velocity was not so uniform. Figure 10-11 through Figure 10-14 show the measured velocities compared to the plug flow average migration velocity. As a matter of convention, negative velocities indicate that the particles were flowing downward. The top elevation in Figure 10-11 matched the plug flow average under the bauxite distributor, and read a zero velocity outside the distributor zone. The next lower elevation shown in Figure 10-11 had less downward velocity at the center. The next two elevations shown in Figure 10-12 and Figure 10-13 showed a uniform velocity near zero in the center and downward velocities at the combustor wall approaching the average plug flow velocity.

Figure 10-11: Optic Probe Velocity Distribution Elev. 52

Figure 10-12: Optic Probe Velocity Distribution Elev. 42,3

Figure 10-13: Optic Probe Velocity Distribution Elev. 42,1

Figure 10-14: Optic Probe Velocity Distribution Elev. 34

Figure 10-15 through Figure 10-20 show summary tables of the solids concentration and solids velocity profiles measured by the MSE Mielli optic probe at selected radial positions and combustor elevations for all 6 test series. The measurement grid for the optic probe was rather coarse with one to two insertions at 90° with 6 equal areas insertions from the wall. The optic probe was too short to reach completely across the combustor. These limits may account for some of the variation of the velocity results. However, the concentration results were more uniform, indicating the usefulness of the optic probe for concentrations but not for velocity.

Measurements 'CMB Combustor' - MTF-CMB 'Gas only'

Test 1 - 6/7.12.04

Customer: Alstom Inc., Power Plant Laboratories
 Product: Bauxite 300 μm
 Experiment: Labasys® 100 in 'CMB-MTF' Unit, tip '3401500 mixed'

Load = 2.21, $v/v_t = 0.66$

Figure 10-15: MSE Meili Solids Concentration and Velocity Profiles – Test 1

Measurements 'CMB Combustor' - MTF-CMB 'Gas only'

Test 2 - 8.12.04

Customer: Alstom Inc., Power Plant Laboratories
 Product: Bauxite 300 μm
 Experiment: Labasys® 100 in 'CMB-MTF' Unit, tip '3401500 mixed'

Load = 4.91, $v/v_t = 0.30$

Figure 10-16: MSE Meili Solids Concentration and Velocity Profiles – Test 2

Measurements 'CMB Combustor' - MTF-CMB 'Gas only'

Test 3 - 8./9.12.04

Customer: Alstom Inc., Power Plant Laboratories
 Product: Bauxite 300 μm
 Experiment: Labasys® 100 in 'CMB-MTF' Unit, tip '3401500 mixed'

Load = 2.46, $v/v_t = 0.30$

Figure 10-17: MSE Meili Solids Concentration and Velocity Profiles – Test 3

Measurements 'CMB Combustor' - MTF-CMB 'Gas only'

Test 4 - 8.12.04

Customer: Alstom Inc., Power Plant Laboratories
 Product: Bauxite 300 μm
 Experiment: Labasys® 100 in 'CMB-MTF' Unit, tip '3401500 mixed'

Load = 1.11, $v/v_t = 0.66$

Figure 10-18: MSE Meili Solids Concentration and Velocity Profiles – Test 4

Measurements 'CMB Combustor' - MTF-CMB 'Gas only'

Test 5 - 9.12.05

Customer: Alstom Inc., Power Plant Laboratories
 Product: Bauxite 300 μm
 Experiment: Labasys® 100 in 'CMB-MTF' Unit, tip '3401500 mixed'

Load = 0.55, $v/v_t = 0.66$

Figure 10-19: MSE Meili Solids Concentration and Velocity Profiles – Test 5

Measurements 'CMB Combustor' - MTF-CMB 'Gas only'

Test 6 - 9.12.04

Customer: Alstom Inc., Power Plant Laboratories

Product: Bauxite 300 μm

Experiment: Labasys® 100 in 'CMB-MTF' Unit, tip '3401500 mixed'

Load = 2.21, $v/v_t = 0.66$

Figure 10-20: MSE Meili Solids Concentration and Velocity Profiles – Test 6

10.2. CFD Modeling of Gas Heat Transfer

One of the objectives of the MTF scaled heat transfer test program was to collect the detailed validation data necessary to calibrate a CFD model of the CMB process. Once calibrated, the CFD model would be used to scale up the results to predict CMB performance at commercial scale. This section describes the CFD model validation and then the application of the CFD model for commercial predictions.

A CFD model was developed in Fluent 6.1 for the CMB configuration of the MTF. This model began at the 20-foot elevation of the MTF elevation and continued to the top of the combustor. This configuration was selected because the MTF was heated by natural gas during the heat transfer scale-up tests. It was assumed that all of the combustion was completed by the time the gases had reached the 20-foot elevation. This simplified the CFD model because combustion did not have to be included in the model calculation.

The model used a very fine grid with about one million cells. This was significant because the size of the cell could affect the calculation of the particle flight. With a larger unit, there were cell count limitations, which made it necessary to use larger cells.

The model was run for six different cases, which used the input values shown in Table 9-13. Then the model results were compared to the data that was collected for the six MTF test series. As expected, the results from the CFD model were not a good match with the measured data. The heat transfer observed in these tests and in previous tests was considerably less than that predicted by the default heat transfer mechanism in the Fluent model.

The default heat transfer coefficient correlation used in the model is based on a single particle moving relative to the gas. The following correlation is used:

$$Nu = hD/k = 2.0 + 0.6Re^{0.5}Pr^{0.33}$$

where

Nu = Nusselt number.

h = Convective coefficient.

D = Particle diameter

k = Thermal conductivity

Re = Reynolds number

Pr = Prandtl number

The particle model in Fluent does not account for interactions between particles. These include effects such as particle collisions and the possibility that the particles are traveling together in groups. The overall heat transfer in large groups of particles was observed in past MTF tests to be significantly less than the sum total of all particles using the Fluent default heat transfer coefficient correlation.

A practical result from using a higher than measured heat transfer coefficient in the Fluent CFD model is that overall heat transfer can actually be less than expected. This is because the local higher heat transfer calculated (with the higher heat transfer coefficient) affects the gas flow. The temperature, pressure, and velocity are artificially changed by a larger amount

than the data would indicate. The gas velocities affect the calculated flight path of the particles. In extreme cases, the gas and particles are separated because of the high velocities. Separating the particles from the high gas flows reduces the amount of heat that can be transferred. In some cases the overall heat transferred, as shown by the resultant overall outlet temperatures, can coincidentally match the data. When this happens, the velocity flow field won't match the data. This situation happened with previous models of the CMB process and with the earlier tests.

MODELING RESULTS

In the MTF scaled heat transfer tests, the data did not match either the temperature profiles nor the velocity profiles calculated by Fluent using the convective heat transfer coefficient built into the model. Therefore, the model was calibrated by employing a user-defined function to modify the value of the convective heat transfer coefficient. The models for each MTF test condition were run with different values for the convective heat transfer coefficient until both the calculated temperatures and the velocities at each test elevation were close to the data measured at those elevations in the MTF.

The results of these calculations are presented Figure 10-21 through Figure 10-26.¹⁷ The first three symbols (♦, ▲, ✕) in Figure 10-21 represent the temperature data from the thermocouple rakes at elevation 34ft, 42 ft, and 52ft. The next three symbols (+, ■, —) represent the CFD model predictions for temperature at each of these three elevations. The three solid lines represent the best fit for the measured temperature data for all three thermocouple rake elevations. For five of the six test series, the convection coefficient was best modelled by using 0.1 ± 0.01 times the model default heat transfer coefficient. For the remaining test, the best value was 0.2 ± 0.01 times the model default heat transfer coefficient. Trying to get a more precise number was subjective, given the need to match several elevations of test data. The one test that indicated a higher convective heat transfer coefficient was test series five. This test used a solids-to-gas ratio of only 0.5 whereas the other tests used a solids-to-gas ratio of 1.2 or higher. This result makes sense because as the total number of particles decreases, the heat transfer coefficient would be expected to rise towards the default value. It would take considerably more testing to determine the actual relationship between the heat transfer coefficient (used in Fluent) and the solids-to-gas ratio.

¹⁷ *Analysis of CMB Scaled Heat Transfer Tests*, Paul Thibeault, ALSTOM Power Internal Report, January 2009

Figure 10-21: Temperature Profiles for Test 1 Run 2

Figure 10-22: Temperature Profiles for Test 2 Run 18

Figure 10-23: Temperature Profiles for Test 3 Run 25

Figure 10-24: Temperature Profiles for Test 4 Run 11

Figure 10-25: Temperature Profiles for Test 5 Run 32

Figure 10-26: Temperature Profiles for Test 6 Run 39

Figure 10-27 through Figure 10-29 present an alternative view of the CFD results. They show contour plots of velocity, temperature, and particle concentration at various elevations. The three elevations below the injector are the same as the data planes in the MTF tests.

Figure 10-27: MTF Test 11 - CFD Contours of Velocity

Figure 10-28: MTF Test 11 - CFD Contours of Temperature

Figure 10-29: MTF Test 11 - CFD Contours of Solid Concentration

At the plane closest to the solids injector, the CFD model predicted that there was a large gradient in all of the variables plotted directly below the injector. There was less of a gradient in these planes closer to the walls. The gradients were not as obvious in the other planes. This effect was shown quantitatively in the plots of the CFD results and test data. At the elevation nearest the injector, the CFD model predicted a higher gradient in temperature and velocity than what was observed in by the data. The most likely explanation for this result was that the CFD model did not have enough resolution to properly define the gradient in these values. However, the CFD model did a reasonable job of describing the flow field velocities and temperatures in most of the test rig.

Another aspect of the MTF scaled heat transfer test was to investigate the use of the laser probe developed by MSE Meile. This probe could be used to calculate solids velocity and solids concentration. It was previously used to measure solids velocity in a moving bed application for another test program at ALSTOM Power Plant Labs (PPL). In this application, an attempt was made to use this instrument to measure the velocity of the falling solids and to measure the solids concentration of the falling solids in the gas flow field.

Velocities calculated by the CFD model did not correlate very well with the measurements taken on the laser probe. They were not reported here because the analysis discrepancy was excessive. It was felt that the probe needed to be calibrated and the work required to do that was beyond the scope of this project.

The CMB predictions of solid concentrations are compared to the laser probe results in Figure 10-30 through Figure 10-34. Test 5 was not reported here because there were inconsistencies with the solids concentration data. In all of the cases, the solids concentration measurements at the plane closest to the injector did not match the probe at all. The CFD model predicted a much larger gradient than the probe measurement and also a much higher average concentration. At the other two data elevations, the model predictions were much closer to the probe data, and the gradients seemed to roughly follow each other. The CFD models were still predicting higher concentrations than the probe. In some cases the CFD reading was seven to eight times higher.

Figure 10-30: Solids Concentration (Vol %) Laser Probe Data Compared to CFD Results for Test 1

Figure 10-31: Solids Concentration Laser Probe Data Compared to CFD Results for Test 2

Figure 10-32: Solids Concentration Laser Probe Data Compared to CFD Results for Test 3

Figure 10-33: Solids Concentration Laser Probe Data Compared to CFD Results for Test 4

Figure 10-34: Solids Concentration Laser Probe Data Compared to CFD Results for Test 6

The CFD predictions at the highest data plane were probably over-predicting the solids concentration gradient. However, the CFD model was forcing a mass balance on the models so that the average concentration over the cross section was probably more accurate than the probe readings.

The probe seemed to match the concentration gradient in the lower two planes, so that the probe was actually measuring something predicted by the CFD. It was felt that the probe needed to have more calibration work done to allow it to be used to measure free-falling solids. This would be a valuable thing to do but calibrating the probe was beyond the scope of this project.

COMMERCIAL PLANT MODEL SCALE-UP

The Fluent CFD model was developed to scale up the results from small-scale tests such as the MTF to larger commercial sized units. The original CFD models developed during the CMB Phase I project predicted that commercial sized CMB combustors might have a gas flow field with large gradients in velocity that would cause the falling solids to segregate from the gas. This would cause several severe problems for the commercial size unit:

- The gas-to-solids heat transfer would be less than predicted. The combustor size and cost would increase significantly as heat transfer decreased.
- The divergence of falling solids and rising gases within the combustor would also be exacerbated as the heat transfer decreased, resulting in excessive combustor gas outlet temperatures.

- Local gas temperatures in the combustor might exceed the ash fusion temperature, resulting in slagging, agglomeration, and potential defluidization issues.

As described in the previous CFD model validation section, using Fluent's default heat transfer correlation resulted in a heat transfer coefficient that was higher than the actual measured value and could lead to the problems just discussed. The CFD model validation with the MTF test data showed that for gas and solids flows with a solids to gas ratio about 1.0 or higher, the heat transfer coefficient that most closely matched the test data was about 0.1 times the default value in the Fluent CFD code.

The commercial model developed in Phase I was therefore revised to incorporate the lower heat transfer coefficient. Two different scenarios were run. One scenario used secondary air injected tangentially just over the bubbling bed part of the process. The second scenario used radially injected secondary air to better mix the gas and solids. The results for scenario 1 are shown in the Figure 10-35 through Figure 10-37 and the results for scenario 2 are shown in Figure 10-38 through Figure 10-40.

Figure 10-35: CFD Model of Commercial Plant Case 1 - Surface of Solid Concentration = 0.25lb/ft³

Figure 10-36: CFD Model of Commercial Plant Case 1 - Contours of Velocity

Figure 10-37: CFD Model of Commercial Plant Case 1: Contours of Temperature

Figure 10-38: CFD Model of Commercial Plant Case 2 - Surface of Solid Concentration = 0.25lb/ft³

Figure 10-39: CFD Model of Commercial Plant Case 2 - Contours of Velocity

Figure 10-40: CFD Model of Commercial Plant Case 2 - Contours of Velocity

Figure 10-35 and Figure 10-38 show the flow patterns of the solids. This was represented by a contour surface of solids concentration of 0.025 lb/ft^3 . The colors indicated temperature. The flow pattern of the falling solids was roughly a cone shape with the solids gradually moving toward the walls. Case 1 showed a slight swirl in the falling solids flow pattern. The rest of the commercial plant results showed the concentration of solids, temperature, and velocity at various elevations.

The solids were flowing downward in a gently expanding cone pattern. The gas had a higher velocity in the center of the unit but had to travel through the falling solids to exit at the top in three cyclone outlet ducts. This flow pattern resulted in very good heat transfer even with the use of a low heat transfer coefficient. The results were greatly improved as compared to the previous results from the original commercial CFD models, which showed varying degrees of gas and solids flow separation.

The results from these cases showed that with a solids temperature of 1040°F entering at the top, the gas outlet temperature was 1238°F for case 1 and 1358°F for case 2. This was actually very good heat transfer performance.

The temperatures on the other hand were not as good as were needed for the CMB process. The maximum temperature predicted in each case was well over 2800°F in the center of the bed where the solids were not concentrated very heavily. This may not be as bad as predicted, however. These cases did not model the bubbling bed at the bottom of the unit and the concentration of heat predicted by these models was most certainly spread out by the action of the lower bubbling combustor.

Overall, the commercial models provided a reasonable representation of the CMB gas and solids flow field when modelled with the heat transfer coefficient derived from the MTF scaled heat transfer test results. There did seem to be some additional work needed to refine the model to include a better representation of the bottom of the CMB combustor.

CONCLUSIONS

- The MTF test results showed that the average convective heat transfer coefficient that should be used in Fluent is 0.1 times the default value when the solids to gas ratio is greater than 1.0. It should be 0.2 times the default value when the solids to gas ratio is about 0.5.
- A CFD model can give a reasonable representation of the CMB gas and solids flow field. However, the area around the injector seems to over predict the concentration gradient.
- The MSE Meile laser probe was promising as a device to measure gas velocities and solid concentration in a falling solid stream but a significant amount of additional calibration may be required. At present, it is not calibrated well enough for this application but has been calibrated for a moving solid bed.
- The CFD model can give a reasonable representation of a commercial sized CMB. There are still development issues especially about representing the bubbling bed combustor portion of the CMB.
- The commercial plant performance is much better than predicted in the previous CFD models in regards to gas to solids heat transfer. There is still some question about the maximum gas velocities and temperature.

10.3. MBHE Heat Transfer

MBHE heat transfer was investigated during a series of cold flow tests and MTF tests in both CMB and O₂-fired CFB mode. MBHE heat transfer was later investigated during the extended field tests at the Grant Town CFB. This section describes the results of cold flow and MTF tests. The heat transfer results from the Grant Town tests are described later in Section 12.

10.3.1. Cold Flow Moving Bed Heat Transfer Tests

A series of ambient temperature tests were conducted to expand the measured range of the heat transfer coefficients of horizontal spiral finned tubing with a bed of ash or particles moving slowly past the tubing. The velocity of the particles ranged from 15-300 fph. This low velocity minimizes erosion of the MBHE tubes.

The tests had four objectives:

- 1) Determine the effect of tube bundle depth on MBHE heat transfer performance
- 2) Determine the effect of ash or particle composition on performance
- 3) Determine the effect of transverse tube spacing on performance
- 4) Extend the ash velocity range beyond that achieved in previous PPL MBHE tests

The test section consisted of a slice model at ambient temperature conditions. A 3 tube wide x 10 tube deep bundle was installed in the slice model to simulate the effect of bundle depth on heat transfer performance. Several tube bundle arrangements were evaluated; 6" S_T X 6" S_L staggered, 8" X 6" staggered, 6" X 6" staggered with 8 degree offset, and 6" X 6" staggered with 8 degree inclined. The slice model was nominally 4" X 16" in plan area and 8' in height. Inserts were added to adjust the plan area for different tube arrangements. The arrangement of the tube bundles and test section is shown in Figure 10-41.

The circumferentially finned tubes were electrically heated by cartridge heaters in the core of the tube. Three of the tubes had thermocouples attached to the fin tip and base at four circumferential locations to measure heat flow through the fin to the ash. Additional thermocouples were installed in the ash near each instrumented tube to measure ash temperatures at the top, side, and bottom of the instrumented tube.

Two tube/fin arrangements were tested. The first had 1.27" OD T22 tubing with A209 circumferential fins of 0.852" height and 0.102" thickness. The second had 2.00" OD T22 tubing with A302 circumferential fins of 0.500 height and 0.096" thickness. Both arrangements had circumferential spiral fins on ½ " spacing.

Ash or solids were supplied to the test section from a 1600-2000 lb hopper on load cells to provide a measured solids flow rate. The flow rate was controlled by a variable area distribution orifice under the tube bundle. Solids draining from the test section were collected in a lower hopper. At the end of the test, solids were pneumatically transported from the lower hopper to the upper hopper.

Figure 10-41: MBHE Cold Flow Tube Arrangements

Tube fin to ash heat transfer coefficients for a 6"x6" staggered tube array are shown in Figure 10-42 for the top, middle, and bottom portions of the tube bundle. There is little difference with bed depth at velocities below 100 fph and a small decrease in heat transfer at the bottom of the tube bundle at velocities of 100-300 fph. There were two fin geometries at the top of the tube bundle and different fin alloys, CF01 and ST01. The data shows that fin geometry had little effect on the heat transfer coefficient.

Heat transfer coefficients for an 8"x6" staggered tube array are also shown in Figure 10-42, superimposed on the 6"x6" results. The 8"x6" results were significantly lower than for the 6"x6" array. This was because visual observation indicated that some of the ash flowed down between adjacent tubes without being touched by a heat transfer surface. In effect there was little or no mixing or diffusion of the ash particles. In addition, the heat transfer coefficients showed more deterioration with the 8"x6" array as you moved lower into the bundle.

The ambient heat transfer tests are summarized in Table 10-2 as the average ratio of measured divided by calculated values using the same calculation for all data.

Figure 10-42: MBHE Fin Heat Transfer Test Results for an 6" x 6" Staggered Tube Array Comparing Results with a 8" x 6" Array

Table 10-2: Comparison of Measured/Calculated Ratio of Fin Heat Transfer

6X6 array		Avg	Std dev
CFB Ash	Top	1.00	0.04
CFB Ash	Mid	1.01	0.10
CFB Ash	Bot	0.92	0.13
Bauxite 700	Top	1.09	0.06
Bauxite 701	Mid	0.92	0.07
Bauxite 702	Bot	1.37	0.19
Bauxite 500	Top	1.15	0.11
	Mid		
Bauxite 500	Bot	1.38	0.16
8X6 Array			
CFB Ash	Top 8	0.63	0.02
CFB Ash	Mid 8	0.32	0.01
CFB Ash	Bot 8	0.43	0.02

10.3.2. MTF MBHE Operation In CFB Mode

Results of the ambient test series were used to modify the MTF MBHE for the O₂ fired CFB tests (which were part of a separate DOE program) by reducing the tube spacing to match the 6"x6" array. This gave the same fin tip to fin tip clearance of 0 as was provided in the 6"x6" ambient tests. A schematic and photo of the rebuilt MBHE is shown in Figure 10-43.

Figure 10-43: Modified MTF MBHE with Reduced Tube Spacing

A series of MBHE tests were then conducted during the MTF O_2 firing CFB tests. Details of these tests are described in Section 9.5. Results of the tests are shown in Figure 10-44 and Figure 10-45. Reducing the horizontal S_T tube spacing from 3.5" to 2.5" resulted in significant improvements in heat transfer performance as shown in Figure 10-44. The heat transfer results from the MTF O_2 -fired program (green data points) matched the predicted results from the ambient tests.

Figure 10-44: MTF CMB O₂ MBHE Performance for CFB Ash and Sand With Change of Tube Pitch

Figure 10-45: Comparison of MBHE Test Performance with Predictions for Various Fuel/Combustion Gas and Operating Conditions

10.3.3. MTF MBHE Bauxite Test Results

Additional MBHE heat transfer data was gathered during the CMB MTF test campaigns that were described in Sections 9.2 and 9.3. During the CMB Phase I program, a remote MBHE was installed adjacent to the MTF and bauxite from the MTF combustor was pneumatically transported to the top of the MBHE. Bauxite cooled in the MBHE was then pneumatically transported back to the top of the MTF combustor. A schematic and photo of the MBHE for these tests is shown in Figure 10-46.

Figure 10-46: CMB MBHE for Hot Bauxite Tests

A major objective of the CMB MTF test campaigns was to evaluate solids-to-tube heat transfer coefficients in the moving bed heat exchanger with bauxite as the heat transfer media. A heat transfer model developed prior to this program was based on moving bed heat exchanger tests with fluidized bed ash material as the heat transfer media. The MBHE design was refined in this test campaign with a staggered tube array. The heat exchanger was tested over several operating periods, with solids inlet temperatures to the heat exchanger ranging from 1620°F-1800°F. The solids flow rates ranged from 5,200 to 19,300 lb/hr during these periods.

The test series were conducted with the data acquisition system recording data every 3 minutes. The test data was imported into a preliminary analysis spreadsheet and plotted in a series of charts of solids temperatures, solids flow rates, heat flow, and coolant temperatures to determine periods of steady state operation as shown in Figure 10-47.

Figure 10-47: MBHE Solids Temperatures for Typical Test Condition

Once the steady state conditions were determined, average values of each instrument measurement were calculated for a period of at least 30 minutes or ten measurement readings. Figure 10-47 shows a typical set of solids temperatures in the MBHE starting on November 5, 2003. The four highest trend lines represent the 4 temperatures at the top of the MBHE (TE1017-TE1020). The remaining 16 trend lines are the solids temperature measurements located just above each of the 16 solid outlet distributor openings at the bottom of the MBHE. A description of the different thermocouples shown in this figure is provided in Table 10-3.

Table 10-3: Identification of MBHE Thermocouples

Tag #	Description	Units
PT4475	MBHX Top Pressure	" H2O guage
TE1072	MBHX Gas Outlet Temperature	°F
TE1017	Top TC 1 (NW)	°F
TE1018	Top TC 2 (NE)	°F
TE1019	Top TC 3 (SW)	°F
TE1020	Top TC 4 (SE)	°F
TE1021	Bottom TC 1,1 (NW)	°F
TE1023	Bottom TC 1,2	°F
TE1024	Bottom TC 1,3	°F
TE1025	Bottom TC 1,4 (NE)	°F
TE1026	Bottom TC 2,1	°F
TE1028	Bottom TC 2,2	°F
TE1029	Bottom TC 2,3	°F
TE1030	Bottom TC 2,4	°F
TE1031	Bottom TC 3,1	°F
TE1032	Bottom TC 3,2	°F
TE1034	Bottom TC 3,3	°F
TE1035	Bottom TC 3,4	°F
TE1036	Bottom TC 4,1 (SW)	°F
TE1037	Bottom TC 4,2	°F
TE1038	Bottom TC 4,3	°F
TE1039	Bottom TC 4,4 (SE)	°F

Average values were calculated from the data in Figure 10-47 and pasted into a summary spreadsheet for all measurements. The average values were then used to calculate and compare measured performance with predicted performance. The following figures indicate results of these average values for the duration of the test program.

The solids inlet temperatures entering the top of the MBHE are shown in Figure 10-48. This figure shows that the MBHE solids inlet temperature never got above 1800°F even though the combustor bed was operating at or above 2000°F. A water-cooled siphon seal between the combustor and horizontal transport pipe absorbed a significant amount of heat from the bauxite particles, which cooled them well below the bed temperature. A Maxon natural gas burner was intended to make up for this heat loss by preheating the transport air in the bauxite transport line. Unfortunately, refractory problems with the burner box casing limited preheating the transport air to about 1800°F.

Figure 10-48 shows that non-uniform solids distribution through the MBHE affected the solids inlet temperatures. This issue became very pronounced after the outlet size area of the lower solids distribution grid was increased (about day 3 in Figure 10-48). This modification was made just before the third operational period to prevent blockage of the MBHE outlet hoppers by refractory pieces that had spalled off from the refractory lined pipe from the MBHE combustor to the top of the MBHE. The flow mal-distribution stemmed from blockage on the south outlet side of the MBHE and extended all the way up to the top of the MBHE. An active distribution grid with large flow passages was installed before cumulative day 7. Initial problems with the air vented from the active distribution grid caused agglomeration in the tube bundle. This was caused from sintering of coal particles transported from the combustor that reacted with the vented air. Improving the vent removal

of the grid air before it reached the hot coal particles in the top of the tube bundle eliminated this problem.

Figure 10-48: MBHE Solids Inlet Temperatures

Figure 10-49 shows the same trends for the MBHE solids outlet thermocouples over the duration of the test campaign. The temperature measurements on the south side of the MBHE initially tended to be lower than average and very unresponsive, indicative of being in a slow or non-moving region of solids. The highest temperature measurements were on the north side and center where most of the solids flow was occurring. Adding the active distribution grid improved the outlet temperature solids distribution.

Figure 10-49: MBHE Solids Outlet Temperatures

Figure 10-50 shows the MBHE cooling water outlet temperatures for the top two bundles over the test period. The water entered at the bottom center of each bundle and exited at the top of the bundle on both the west and east sides. In the initial operating period, the bottom east side of the MBHE was plugged with refractory rubble from the refractory-lined transport line. The cooling water outlet temperatures for the top two bundles were much hotter on the west side (TE1081 and TE1079) during the first two days of operation than on the east side. This was because the solids were preferentially flowing through the MBHE along the west side. The rubble screen was installed above the MBHE outlet distributor at this point to prevent this pluggage. The trends show that the solids flow was still biased on one side for periods of time during the remaining operation.

The accumulation of rubble in the distributing grid required that it be replaced by an active grid design. This grid had the same spacing, but the bottoms of the first section were enlarged and a pneumatic ash flow control valve (ACV) was placed under each grid opening. The ACV was large enough to pass a 2" diameter particle.

Elevation clearances at the bottom of the MBHE did not permit enough ash to form an air seal between the high-pressure pneumatic transport return pipe back to the combustor. Therefore, a pig-tail loop seal was installed under the MBHE to seal the MBHE from the transport back pressure.

Initially a vent was installed under the grid to remove the ACV activation air from flowing up into the tube bundle. This vent was prone to plug and the result was a sintered

agglomeration of ash in the tube bundle. An improved vent eliminated this problem. The sequence of these distributor grid changes is identified in Figure 10-50 to Figure 10-52.

Figure 10-50: MBHE Cooling Water Outlet Temperatures

ALSTOM Power's MBHE design spreadsheets developed through a previous program were used to design the MBHE for the MTF modifications. Derivatives of the same spreadsheets were also used to calculate predicted solids side heat transfer coefficients for the bundles at each MTF test condition. This gave a predicted heat flow for the tube bundles that could be compared with the measured bundle heat flow.

The measured solids inlet temperatures, coolant inlet temperatures, solids flow rate, and coolant flow rates were input as fixed values for these calculations. The solids and coolant outlet temperatures were iteratively adjusted until the coolant and solids heat flows were equal. The temperature effect on fin and tube thermal conductivity was also included in the iterative evaluation.

Figure 10-51 compares the ratio of the measured vs. the predicted heat flow for the top three MBHE tube bundles over the duration of the test period. The measured heat flows for the top tube bundle averaged around that predicted for the initial MBHE operation. After the bottom of the distributing grid was enlarged, the performance started at the 100% improvement value but then started to slowly deteriorate. This same trend is shown by the solids inlet temperatures in Figure 10-48. The heat transfer performance for the lower tube bundles was not as good as the top bundle. The second bundle had an initial 20% lower

rate than predicted (after the distributing grid was enlarged). There was not sufficient bauxite flow to provide measurements for the lower two bundles at all times.

The top tube bundle always had higher heat transfer performance than the lower bundles. This phenomena was also observed in tube bundle test results by Jörg Niegsch.¹ However, the magnitude of the difference between bundles for this test was probably caused by mal-distribution of the solids flowing through the distributing grid at the bottom of the heat exchanger. The bauxite was free flowing, but rubble in the grid caused flow interruptions in individual grid locations. The solids flow tended to redistribute at higher elevations in the heat exchanger. The angle of friction for 700 micron bauxite particles is 55° . Bauxite flows uniformly above this angle. This means that tube bundles located 45 inches above the grid should have relatively uniform flow. Those below will not and will have a degraded heat transfer performance as shown in Figure 10-51. Note that even with non-uniform flow conditions, the measured heat flow for the top tube bundle was the same or better than predicted by the design spreadsheet.

Figure 10-51: MBHE Measured vs Predicted Heat Flow Showing Effect of Solids Agglomeration

Figure 10-52: MBHE Operation with Agglomerated Performance Deleted

The bauxite flow distribution within the heat exchanger was not very uniform during some of the operation. Figure 10-51 shows deteriorated heat flow starting at day 3. Correspondingly, the solids inlet temperatures in Figure 10-48 show that the south half of the MBHE had reduced flows at day 3. Mal-distribution of the solids flow indicated by temperatures was reflected in decreased heat flows.

Control of the solids flow through the MBHE was by a device at the bottom of the heat exchanger, ensuring that it remained full of solids and operated in the packed bed mode as shown in Figure 10-52. Initially, a variable speed rotary valve was used to control the solids flow. A 4X4 passive pyramidal grid above the rotary valve was used to provide a uniform flow distribution at the bottom of the MBHE. The openings in the grid were sized to distribute the flow evenly. However, reduced flow rates did not provide a uniform flow. Refractory rubble and rocks from the coal fed to the combustor plugged the grid openings. A perforated screen located above the grid was of little help in reducing the plugging problem.

An active grid was developed in June 2003 with large openings to prevent the plugging problem. Each opening of the 4X4 pyramidal grid had a pneumatic flow control valve under the opening. The openings in the grid were enlarged to a 2"X2" size and rubble of up to that size could pass through the grid. Compressed air connected to a solenoid bank actuated the flow through each opening in the grid. Air was sequentially supplied to each grid flow control valve for a short period, followed by a rest time. Varying the difference between operating and rest time controlled the flow through the valve. This timing was controlled by a computer solenoid control system and was manually controlled to provide uniform temperature profiles at the bottom of the MBHE. The intermittent flow through the grid did not significantly affect the MBHE heat transfer performance.

Solids exiting from the bottom of the MBHE passed through a pigtail loop seal, into a pneumatic transport pipe, back to the top of the combustor. Backpressure from the transport pipe was sealed by the loop seal. However, grid activation air could not pass through the pigtail seal and needed to be vented. Initially, an under grid vent was used, but was prone to pluggage by solids. This pluggage caused grid activation air to pass upward through the tube bundle. Flow distribution was uneven and in one case grid activation air permeating up through the bed caused sintering of coal particles transported in the MBHE.

A revised vent consisting of two rakes over the active grid was installed October 2003, and two standpipes passing through the grid enabled the grid activation air to be vented to the top of the MBHE and out the transport air vent. This rake/standpipe modification eliminated upflow of air from the grid to the tube bundle. The MBHE was operated for 3 days with the combustor firing the West Virginia coal and the petroleum coke with no deterioration in performance.

Figure 8-4 shows the MBHE Active Grid Rake Vent System in the right figure and the MBHE Under Grid Vent in the center figure. The left figure shows the original passive grid with rotary valve control.

Summary

The MBHE test results showed that the performance of the tightened pitch tube bundles exceeded the performance of the previous inline tube bundles by up to 60%. It is possible that the tube bundle performance could have been even higher than observed since the tube bundle showed signs of significant solids flow mal-distribution even when the best heat transfer rates were obtained. One promising result was that the tube bundle fins were largely rock-free, with only a few instances of rocks lodged in the fins. The staggered tube bundle performance could now be predicted for both hot and cold conditions as well as for variations in tube bundle configuration.

10.4. Agglomeration

The CMB Phase I program showed that the heat transfer media (bauxite particles) gradually accumulated an ash coating during their exposure in the 2000°F CMB bed. A commercial CMB would need to continually remove some of these particles and replace them with fresh bauxite to maintain constant bed inventory. Given the high cost of the sintered bauxite particles used in these MTF tests, one of the tasks of Phase II was to explore alternative bed materials. Section 8.2 discussed the preliminary evaluation of a number of alternate bed materials that could be used for CMB heat transfer media. One of these materials might have been selected for further testing in a future MTF test campaign. However, the Phase II MTF tests were still conducted using the sintered bauxite particles because the MTF test campaigns were scheduled in advance of the preliminary evaluation of alternate heat transfer media.

One objective of the CMB test campaigns was to continue evaluating the effect of high bed temperatures on ash agglomeration and ash build-up on the bauxite particles. Bed samples were collected at frequent intervals during the tests to monitor the growth of ash deposits on the bauxite particles. Selected samples were then sent for scanning electron microscope (SEM) analysis in order to monitor the build-up of ash coating on the bauxite particles.

The September 2003 MTF test campaign was run with a Pittsburgh #8 bituminous coal for about 80 hours of operation at 2000°F. The November 2003 MTF test campaign was run with both a West Virginia bituminous coal and petroleum coke for 88 hours of operation at 2000°F. Figure 10-53 summarizes the bed temperature history for the duration of the November 2003 test campaign with the West Virginia bituminous coal and a petroleum coke. For comparison, the light blue band in this figure indicates the normal operating temperature range in a conventional CFB boiler.

Figure 10-53: MTF Temperature History for Nov 2003 MTF Test

The local bed ash composition may play a key role in the ash coating growth rate observed on the bauxite particles. The CMB system is designed to operate with a bed that is primarily bauxite. However, the bed material was not classified during these tests, which allowed rocks and coal ash to build up in the bed and increased the presence of ash constituents that contribute to agglomeration and ash coating growth.

A scanning electron microscope was used to analyse polished cross-sections of coated bauxite particles which had been embedded in an epoxy resin. The SEM was a JOEL 840A electron probe microanalyzer with Thermo-NORAN's Vantage microanalysis system. Energy dispersive spectroscopy (EDS) provided quantitative elemental analysis of the coatings, which in turn identified the elements used in the elemental X-ray mapping of several coated samples. Using the Vantage's digital imaging capabilities, backscattered electron images were generated to contrast the coatings from the bauxite substrate. Thickness measurements were made from the coating images using the software's generated measuring capabilities. Thickness measurements were taken at many different locations for each particle, with at least five different particles measured for each sample. All

of the thickness measurements were then averaged for a composite thickness for that time period.

10.4.1. Ash Coating Growth

The MTF operation for the September 2003 test with the Pittsburgh #8 coal used the remaining bed material left over from the Phase I test campaign with West Virginia coal as the initial charge for both the bed and the MBHE. The ash buildup on the bed particles during this campaign was therefore a mixture of ash deposits from two coals. Figure 10-54 shows the increase in the coating thickness surrounding the bauxite particles for the duration that they had been used for testing. The data through the first 108 hours was from Phase I while burning the West Virginia bituminous coal. The data beyond 108 hours was from the September 2003 MTF test with Pittsburgh #8 coal. Also included in blue (data points and trend line) are the results from a 2001 MTF test campaign with West Virginia coal and no MBHE.

Figure 10-54: Bauxite Coating Thickness Through September 2003 MTF Test

As noticed during the Phase I CMB program, some of the bauxite particles behaved differently than others. The particles seemed to be divided into three distinct groups. The furnace was initially loaded with bauxite that was obtained before this program. This bauxite visually looked black. After running out of this material, we then used some newer bauxite from the same supplier. Although it was the same material, it visually had more of a brownish color. A third group of bauxite particles had a reddish cast to them. These were particles that had been exposed to the bed and had developed a coating that gave them some color.

As shown in Figure 10-54, the brownish particles had virtually no coating develop on them over the entire duration of the test campaign. The black particles developed a coating up to around 25 microns. The reddish particles, which were not very abundant, had coatings of around 50 microns by the end of the September 2003 test.

Table 6.3-1 shows an EDS analysis of the original brown (divided into light and dark) and the black bauxite particles from the prior September 2002 MTF test. The composition for all of the particles was very similar, with about 80% Al_2O_3 , 7% each of SiO_2 and Fe_2O_3 , and 4% TiO_2 . There were no obvious differences to account for the unusual coating growth behaviors between these different bauxite particles.

Table 10-4: EDS Analysis of Bauxite Particles

Compound Formula	Compound Wt.% (Ave)		
	Light Brown	Dark Brown	Black
Al_2O_3	80.46	77.62	80.78
SiO_2	7.62	8.07	7.46
TiO_2	3.85	4.45	3.58
Fe_2O_3	7.09	8.58	7.37
MnO	0.31	0.44	0.00
ZrO_2	0.67	0.85	0.67
CaO	0.00	0.00	0.13

In order to compare the coating growth data from this campaign with the earlier MTF test program, the operating hours need to be compared on the same basis. The 2001 MTF test was run as a 2000°F bubbling bed with no bauxite recirculation. The bauxite particles therefore were exposed to the high temperature bed conditions for the full duration of the test. The September 2002 CMB test had a bauxite recirculation loop, but no MBHE. Since it took some time for the bauxite particles to pass through the solids separation hopper on top of the combustor, the bauxite particles in the September 2002 test only were exposed to the 2000°F bed about 57% of the operating time. The September and November 2003 MTF tests both had the MBHE in operation. The bauxite particles now spent a long time passing through the MBHE. Thus, the actual solids exposure time to high temperature conditions was considerably less than the total residence time. During these tests, the bauxite particles were exposed to the high temperature bed for only about 12 minutes every hour. The particles spent the balance of the time passing through the transport system, the MBHE, and the solids separation hopper on top of the combustor. Figure 10-55 normalizes the particle's residence time from each test as time spent at the full 2000°F bed temperature.

The coating growth rate was expected to be slower for these tests compared to the earlier 2001 tests due to the mechanical abrasion and friction the particles experience in the transport loop and passing through the MBHE. The 2001 MTF test was run as a bubbling bed only and the particles were therefore not subjected to any attrition in the transport system. However, these results are difficult to compare with the earlier results because of the different characteristics of the three different colored bauxites. The brown bauxite clearly had dramatically less coating growth, the black had about the same or slightly less, and the reddish particles had more coating.

Figure 10-55: Bauxite Coating Thickness vs. Exposure Time (West Virginia Coal)

Figure 10-56 restates the coating growth rate as a % volume increase in particle size per hour, with the duration normalized as actual high temperature exposure time. There was considerable scatter in the results for the three different types of bauxite. However, the coating growth rate was the same or in many cases less than the growth rate observed in the earlier testing without a transport system. The difference was at least partially attributed to attrition in the transport line wearing down the coating.

Figure 10-56: Coating Growth Rate vs. Exposure Time Through September 2003 MTF Test

Figure 10-57 shows SEM photos at four levels of magnification of a coated bauxite particle after 42 hours of operation while burning the West Virginia coal. These photos were from the Sept 2002 MTF test, but is typical of all of the SEM photos taken throughout this program. This particle was one of the “red” bauxite coated particles, which showed the thickest coating of the three types previously mentioned. The red boxes in this figure show the area that is magnified in the next photo.

Figure 10-57: SEM Photos after 42 Hrs Exposure (West Virginia Coal)

Figure 10-58 shows an EDS elemental mapping of several different regions in this coating. The coating was primarily a mixture of calcium, aluminum, silica, and iron oxides in varying proportions. There were only minor percentages of other compounds, including MgO and Na₂O. There was virtually no sulfur present in the coating. This could be significant, as the University of North Dakota had observed that sulfate bonding was a key mechanism in the second stage of agglomeration, where particles begin to stick together. With the high CMB bed temperatures, it was unlikely that much sulfation was taking place.

Compound	Compound Wt %			
Formula	Area 1	Area 2	Area 3	Area 4
SiO ₂	24.67	20.15	11.47	0.45
Al ₂ O ₃	25.53	28.92	11.84	31.95
Fe ₂ O ₃	8.17	10.12	54.20	38.97
CaO	40.95	37.76	18.20	0.28
MgO	0.68		0.81	0.33
TiO ₂		2.17	0.84	24.91
P ₂ O ₅		0.88		
Cr ₂ O ₃			0.55	
MnO			0.69	2.42
NiO			1.41	
ZrO ₂				0.70

Figure 10-58: EDS Elemental Mapping of Bauxite Particle after 42 Hrs Exposure

The X-ray mapping shown in Figure 10-59 shows the distribution of oxygen, aluminum, silica, calcium, iron, and titanium throughout the coating of the same particle. Calcium and silica were both distributed uniformly throughout the coating.

Figure 10-59: X-Ray Mapping of Bauxite Coating after 42 Hrs Exposure

The results in Figure 10-56 showed that the coating growth rate was slower in the MTF test campaigns when the MBHE and bauxite transport system were in operation. The attrition in the recirculation loop removed some of the coating and slowed although not eliminated the buildup of ash on the particle surface. Although the growth rate was slowing, the MTF was not run long enough to determine if the particle ash coating would eventually reach a steady state level, where any further growth was offset by attrition. There was some evidence from prior work that some of the coating was spalling off when the coating became thicker. Further operation is needed to look at the long-term effects of coating growth and its effect on CMB operation.

The effect of the several different coals on the coating growth rate is shown in Figure 10-60. The ash coating growth rate from the Pittsburgh #8 coal tests in February 2003 are shown as the green data points and trend line. The ash coating growth rate from the West Virginia coal tests in September 2002 are shown as the brown, black, and red data points (up to a high temperature exposure of 60 hours), representing the three different colored bauxite samples that behaved quite differently. A blue trend line is also shown for the black colored bauxite while firing the West Virginia coal. The light blue squares (which are circled) are from the November 2003 test campaign, which first burned West Virginia coal and then petroleum coke.

Despite the scatter with the West Virginia coal results, it's apparent that the ash coating layer grew more rapidly on the bauxite particles during the Pittsburgh #8 coal fired tests. This was not surprising, as the ASTM Initial Softening Temperature (see Table 5.4-4) was only 2140°F for this ash, compared to over 2700°F for the West Virginia bituminous coal ash. At one point in this test, the average bed temperature was even above the Initial Softening Temperature for the Pittsburgh #8 coal ash. Furthermore, the actual temperatures at the burning coal particle surface could be several hundred degrees hotter than the bed temperature and thus well above the ash softening temperature for this coal ash. This coal also has high iron content in the ash and is a high sulfur coal, both factors that increase its propensity to slag.

Figure 10-60: Coating Growth Rate for West Virginia, Pittsburgh #8, and Petroleum Coke

10.4.2. Implications for CMB Operation

Growth Rates

The MTF results show that an ash coating grows on the bauxite surface as the particle is exposed to the high bed temperature. The coating is established very quickly and then grows at a slower rate, as evidenced by the declining growth rate curves in Figure 10-54 through Figure 10-56, and in Figure 10-60. In all cases, the operating hours were normalized to an actual high temperature exposure time to compare data between different MTF test campaigns.

However, the high temperature exposure time of the bauxite particles is a much shorter portion of the total particle residence time in a commercial CMB boiler. The particles are only exposed for 8 minutes in the high temperature bed in a commercial sized CMB boiler before they pass through the moving bed heat exchanger and are then returned to the combustor. One complete circuit through the CMB solids circulation loop takes 2 hours. Thus, the total operating time is 15 times longer in the commercial CMB boiler than the normalized bed exposure time in the MTF.

The coating growth rate was previously defined as the volumetric growth rate of the particle. The growth rate observed for the particles during the West Virginia coal tests was about 0.3% particle volume per hour exposure, based on the black bauxite particles after 67 hours of hot exposure time (or 112 hours of total operating time). Note that the growth rate was initially very high and was steadily declining with time. Furthermore, the black bauxite particles had a much higher growth rate than the brown bauxite particles, which showed practically no coating growth. It is reasonable to expect that the long-term growth rate could be considerably less than the 0.3% rate. For discussion purposes, assume that the 0.3%/hr rate is an upper growth rate and that the lower growth rate is one-third that rate, or 0.1%/hr.

For a commercial CMB boiler, these rates would be 15 times lower if expressed on an elapsed time basis. In this case, the upper and lower growth rates are 0.02% and 0.007% particle volume per hour elapsed time.

Based on these growth rates, we can infer some implications for CMB operation; particularly as they impact bed hydrodynamics and heat transfer in the moving bed and combustor. Figure 10-61 shows the minimum fluidizing velocity (U_{mf}) and terminal velocity (U_{term}) for various bauxite particle sizes. Assume that the CMB bed operates at a minimum of 12 ft/sec (during turndown conditions) and that we want the bed superficial velocity to be at least 4 times greater than U_{mf} to ensure good fluidization. This requires a maximum U_{mf} of 3 ft/sec and correspondingly, a maximum bed particle size of 1300 microns.

Figure 10-61: Minimum Fluidization and Terminal Velocities for Bauxite

Figure 10-62 shows the total time necessary for a bauxite particle to grow to a given size in a commercial CMB boiler based on particle growth rates of 0.02% and 0.007% particle volume per hour elapsed time. It would take between 3 to 10 years for the bauxite particles to grow to a maximum size of 1300 microns while firing West Virginia coal, based on the upper and lower growth rates, respectively. Based on this, the hydrodynamic impact of particle growth on bed fluidization is not a significant problem.

Figure 10-62: Exposure Time Required for Final Particle Size

Moving Bed Heat Transfer

The particle size growth should also have only a minor impact on moving bed heat transfer performance. Cold flow moving bed heat transfer tests conducted prior to the CMB Phase I program showed only a slight decrease in the heat transfer coefficient when the bed particle size was increased by 40%. This was equivalent to a bed particle growing from 700 to 1000 microns. Based on Figure 10-62, the particle growth time from 700 to 1000 microns is much quicker, taking between 1 and 3 years while firing West Virginia coal, respectively. Circulating the bauxite at a higher rate would offset this reduction in heat transfer coefficient. This would increase the heat transfer coefficient slightly but also increase the log mean temperature. For the CMB design case, increasing the bauxite recirculation rate less than 2% would offset the reduction in heat transfer due to particle size growth.

Solids-to-Gas Heat Transfer

The effect of particle growth on the gas-to-solids heat transfer in the combustor may be more significant. The larger particle size has a higher terminal velocity and thus a shorter residence time for heat transfer. There will also be less heat transfer surface area due to fewer (but larger) particles if the solids inventory is also reduced to maintain a constant bed pressure drop. Figure 10-61 shows that the terminal velocity increases from 23 ft/sec to 33 ft/sec as the particle diameter grows from 700 to 1000 microns. This results in a reduction in heat transfer to the particles and a corresponding increase in the gas outlet temperature if process conditions aren't adjusted.

Bed Pressure Drop

The ash coating growth will also have a significant impact on bed pressure drop and thus solids management. The coating increases the weight of the bauxite particles and will gradually increase the bed pressure drop if the total solids inventory is not controlled. The bed inventory will increase by 3% to 9% over a month's operation if the total inventory of

bauxite remains unchanged. This is based on the lower and upper growth rates while firing the West Virginia coal, respectively, and assuming an ash coating density of 160 lb/ft³. Some bed inventory will thus need to be removed on a regular basis to control the bed pressure drop.

Solids Management

The bauxite particle size will need to be managed to some maximum size level to maintain acceptable boiler performance and to control bed inventory. There are several ways of controlling particle size, such as through attrition, growth rate control, or solids replenishment. These methods are briefly discussed below.

There is considerable mechanical abrasion on these particles as they circulate through the moving bed heat exchanger and transport lines. It was hoped that this alone would be sufficient to control the particle growth rate. The cold flow attrition testing conducted in the CMB Phase I project showed that the coating could be completely attrited away during cold transport conditions. The hot MTF tests also showed a reduction in the coating growth rate compared to previous tests without a bauxite recirculation system. However, there was still a gradual increase in the coating with exposure time despite the attrition. This rate may slow down or stop if a steady state condition is reached between growth and attrition rates.

It may also be possible to modify the local bed operating conditions to slow down the coating growth rate. One possibility is to minimize the presence of CaO in the bed, which can act as a fluxing agent. Grinding the limestone fine and injecting it above the bed may reduce the coating growth rate, while at the same time improving the quality of the lime for NID performance.

The circulating solids size can also be managed by continuously removing some of the coarser material and replacing it with fresh material. This includes the removal of some bauxite along with the ash. The bauxite inventory that is removed will eventually need to be replenished. One cost-savings option is to sell the used bauxite to an oil company. Fresh bauxite particles are pumped into oil wells for enhanced oil recovery. One oil company purchases 100,000 tons/yr of bauxite for this application. The 4000 tons from a 300Mw CMB could easily be consumed by one of these companies and may be attractive to them at a somewhat reduced price. The physical properties of the coated bauxite should still be very similar to what the oil companies pump into their wells, as sphericity, durability, and sizing are their primary criteria.

Another option is to consider the use of alternate heat transfer media if the rate of bauxite replenishment is too high. Section 8.2 discusses a preliminary evaluation of eleven potential candidates as alternate heat transfer media. Hot MTF tests with the alternate material would be needed to ensure that the bed still operates reliably without agglomeration. Successful operation would be evidence that a less expensive bed media could also work in this environment. Dolomite is an example of a relatively low-cost material that could be considered for the heat transfer media. It is also frequently cited for its ability to minimize the formation of low-temperature melting eutectics, which could be beneficial in mitigating the formation of the coating on the particles.

10.5. Combustion Efficiency

One objective of the MTF test campaigns was to assess the critical operating parameters that affect CMB combustion performance. The CMB process is quite different from CFB

combustors. Although bed temperature is several hundred degrees hotter than CFBs, the combustor temperature is not isothermal as in CFBs. The CMB combustor temperature profile decreases with combustor height and the actual gas residence time at temperatures above 1500°F is less than a CFB. Furthermore, the ash recirculation rate from the main cyclone is very low compared with a CFB.

Combustion Efficiency

Combustion performance is often characterized by the combustion efficiency or the carbon heat loss. The combustion efficiency is defined by

$$\eta_{\text{comb}} = 100 - [(C_{\text{Drain}} + C_{\text{Flyash}} + C_{\text{FBHE}} + C_{\text{Hopper}}) * 14086 + \text{CO} * 4347] / (W_{\text{Coal}} * \text{HHV}) * 100$$

where

C_{Drain}	= carbon loss in bed drain (lb/hr)
C_{Flyash}	= carbon loss in flyash (lb/hr)
C_{FBHE}	= carbon loss in FBHE drain (lb/hr)
C_{Hopper}	= carbon loss in bauxite separation hoppers (lb/hr)
CO	= carbon monoxide in flue gas (lb/hr)
W_{Coal}	= coal feed rate (lb/hr)
HHV	= coal higher heating value (Btu/lb)

The major loss that needs to be considered in the combustion efficiency calculation is the flyash carbon loss. This loss was found by measuring the flyash rate and carbon concentration through a high-volume, pseudo-isokinetic sampling technique at the combustor cyclone discharge.

Carbon Heat Loss

Since the primary heat loss is through the carbon in the flyash, it is easier to spot trends by evaluating the carbon heat loss (C_{loss}). This is expressed as a percentage of the coal heat input through the equation

$$C_{\text{loss}} = (C_{\text{Drain}} + C_{\text{Flyash}} + C_{\text{FBHE}} + C_{\text{Hopper}}) * 14086 / (W_{\text{Coal}} * \text{HHV}) * 100$$

10.5.1. September 2003 Test – Pittsburgh #8 Bituminous Coal

The Pittsburgh #8 bituminous coal was fired during the September 2003 test campaign. Results from these tests are summarized in Table 10-5. The table also includes selected operating parameters, including bed temperature, combustor outlet temperature, and oxygen content in the flue gas.

Table 10-5: Combustion Performance With Pittsburgh #8 Bituminous Coal

<u>Sample Time</u>	<u>Unburned Carbon in Flyash (%)</u>	<u>Flyash flow lb/hr</u>	<u>Carbon loss lb/hr</u>	<u>Coal Flow, lb/hr</u>	<u>Carbon Heat Loss (%)</u>	<u>Bed Temp °F</u>	<u>Furnace Outlet Temp, °F</u>	<u>O₂, % dry</u>
9/07/03 01:20	23.23	41	10	610	1.80	1980	800	8.0
9/07/03 14:00	12.42	92	11	750	1.74	2010	1250	4.4
9/07/03 19:32	9.78	113	11	750	1.68	1985	1295	4.6
9/07/03 22:58	10.68	101	11	750	1.64	1975	1330	4.4
9/08/03 18:22	12.87	114	15	750	2.24	1990	1340	4.8
9/08/03 20:30	11.00	122	13	720	2.14	1990	1350	4.0
9/08/03 22:00	10.52	127	13	750	2.03	1980	1300	4.2
9/09/03 03:35	2.30	1505	35	750	5.28	1985	1300	4.4
9/09/03 06:30	4.05	1032	42	750	6.37	1985	1325	4.7
9/10/03 18:29	7.30	192	14	750	2.14	2005	1380	4.5
9/11/03 00:05	7.04	156	11	750	1.67	1990	1375	4.7
9/11/03 02:10	6.69	125	8	750	1.28	1990	1420	4.0

Figure 10-63 shows the Carbon Heat Loss for the cyclone outlet samples. For most of the samples, the carbon heat loss was about 2%. During the two tests on the morning of September 9, the carbon loss was higher. In these tests the cyclone efficiency was deliberately degraded in order to get more of the limestone in the furnace carried over to the backend sulfur cleanup system. As seen in Table 10-5, the percentage of unburned carbon in these two samples was lower than the others, but the overall carbon loss was higher.

The last point in the table was run at a lower excess air. The oxygen content of the flue gas was lower. With the reduced air flow, the temperature at the top of the combustor was higher – see Figure 10-64. This caused a reduction in the carbon loss.

Note that the measurement of the flyash flow was not rigorous, and might be even more uncertain for the high loss test points. The carbon heat loss values are considered to be approximate.

Figure 10-63: Carbon Heat Loss With Pittsburgh #8

Figure 10-64: Effect Of Furnace Exit Temperature On Carbon Loss – Pittsburgh #8

10.5.2. November 2003 Test – Waynesburg Bituminous Coal And Petroleum Coke

Results from these two fuels are summarized in Table 10-6. The table also includes bed temperature, combustor outlet temperature, and oxygen content in the flue gas.

Table 10-6: Combustion Performance From November 2003 Tests

<u>Sample Time</u>	<u>Unburned Carbon in Flyash %</u>	<u>Flyash flow lb/hr</u>	<u>Carbon loss lb/hr</u>	<u>Coal Flow lb/hr</u>	<u>Carbon Heat Loss %</u>	<u>Bed Temp °F</u>	<u>Outlet Temp °F</u>	<u>O₂, % dry</u>
<i>Cyclone Outlet Samples</i>								
11/4/03 0:10	12.38	188	23	802	3.68	1980	1160	5.5
11/4/03 6:34	11.42	135	15	802	2.44	1990	1115	5.8
11/4/03 14:52	11.73	48	6	801	0.89	1985	1225	5.6
11/4/03 18:28	13.92	52	7	801	1.14	1980	1290	4.5
11/5/03 1:50	9.30	206	19	850	2.86	2000	1245	3.7
11/5/03 11:16	6.30	148	9	850	1.40	1995	1320	3.0
11/5/03 14:55	5.03	149	7	850	1.12	1995	1360	3.5
11/5/03 22:05	5.41	128	7	850	1.03	2005	1405	4.2
11/6/03 17:34	12.90	134	17	700	2.40	2010	1310	5.6
11/6/03 23:45	12.88	145	19	715	2.53	2000	1410	4.1
11/7/03 4:25	12.01	138	17	715	2.26	2005	1420	3.7
11/7/03 7:00	12.90	146	19	715	2.55	1985	1400	3.7

Figure 10-65 shows the Carbon Heat Loss for the cyclone outlet samples. The carbon heat loss varied from about 1% to 3.5% of the coal higher heating value for the Waynesburg bituminous and was about 2.5% for the petroleum coke. Figure 10-66 shows that the carbon heat loss increased with decreasing furnace exit temperature for the bituminous coal. There were no flyash samples taken at the lower exit temperatures when firing petroleum coke.

Figure 10-65: Combustion Performance For November 2003 Test

Figure 10-66: Effect Of Furnace Exit Temperature On Carbon Loss – November 2003

Total Heat Losses

The preceding discussion focused on the heat loss due to carbon in the flyash, which was the major combustion inefficiency. There were a few other smaller sources of unburned carbon losses that have also been briefly mentioned:

- unburned carbon loss in other bed drain streams
- unburned carbon loss from bauxite transport lines
- heating value of carbon monoxide (CO) in the flue gas.

Bed Drain

The only other drain stream in the MTF tests was from the fluid bed heat exchanger (FBHE). For long-term commercial operation, there would also need to be some drain from the lower furnace to remove coarse material. This was not needed for these tests. The bed carbon concentration was very low (less than 0.1%) and should be an insignificant loss.

The carbon heat loss in the FBHE drain can be estimated from the drain rate and the carbon content in the ash stream. The amount of FBHE drain material was calculated by a system mass balance (total ash generated - measured flyash). The carbon content in FBHE drain samples was generally less than 1.0% wt.%. Based on the carbon content and FBHE drain rate, the heat loss was typically about 0.1% for each fuel tested.

Transport Losses

The carbon heat loss from the bauxite transport lines was about 0.4% of the coal heat input. However, this carbon loss was specific to the MTF equipment layout. In a commercial system, this ash will be returned to the combustor, giving another chance for the carbon to burn out.

Carbon Monoxide Loss

For either of the bituminous fuels burned in the MTF tests, a CO emission of 100 ppm (dry at 3% O₂) corresponds to about 0.082 lb CO/MMBtu fired. Since the heating value of CO is 4,347 Btu/lb, the heat loss is 0.036% heat loss per 100 ppm CO.

During the West Virginia coal tests in November 2003, the CO emissions were typically around 200ppm at normal conditions (with generally high O₂). During the Pittsburgh #8 bituminous tests in September 2003, the CO emissions generally ranged from 200 to 400 ppm. This makes a heat loss of 0.07 to 0.14% of the coal heating value.

Combustion Efficiency

Based on the above losses, the overall combustion efficiency can be estimated for each fuel. The combustion efficiency during the West Virginia coal tests ranged from 96.1 to 98.9%. This is slightly lower than the 98.5 to 99.6% range observed during the CMB Phase I tests. The combustion efficiency during the Pittsburgh #8 coal tests ranged from 97.6 to 98.3%. This was also slightly lower than the 98.4 to 99% range observed during the Phase I tests. The slight reduction in combustion performance was because the focus of the testing was more directed towards evaluating environmental performance, sometimes in non-optimum conditions for combustion. Finally, the combustion efficiency for the petroleum coke tests ranged from 97.1 to 97.6%.

10.6. NO_x and CO Emissions

An objective of the MTF test campaigns was to assess the environmental performance of the CMB process with respect to NO_x and CO. CMB combustor temperature profiles are significantly different from conventional CFBs, with higher bed temperatures and a declining temperature profile up the combustor. Ammonia injection was also used during one test sequence to evaluate its impact on NO_x reduction.

10.6.1. Environmental Performance With Pittsburgh #8 Bituminous

Table 10-7 summarizes the emissions during the September 2003 test campaign with Pittsburgh #8 bituminous coal. The NO_x and CO emissions in this table have been normalized to 3% O₂.

Table 10-7: NO_x and CO Emissions With Pittsburgh #8 Coal

Test	Start	End	Dur.	Sorbent	Firing Rate	T Bed	T Top	NH ₃ Flow	NH ₃ Elev.	Steam flow	% O ₂ dry	NO _x	CO
1	9/7 09:01	9/7 11:15	02:14	EcoCal	9.25	1991	1237	-		-	3.97	134	316
2	9/7 11:22	9/7 11:51	00:29	EcoCal	9.25	1991	1279	0.15	12N	-	3.09	120	421
3	9/7 11:57	9/7 12:14	00:17	EcoCal	9.25	1986	1284	-		-	3.49	119	397
4	9/7 12:21	9/7 14:13	01:52	EcoCal	9.25	2001	1273	0.13	12N	-	3.78	86	368
5	9/7 14:25	9/7 15:39	01:14	EcoCal	9.25	1982	1283	-		-	4.33	142	205
6	9/7 15:44	9/7 17:12	01:28	EcoCal	9.25	1980	1292	0.14	12N&W	-	4.34	112	213
7	9/7 17:16	9/7 17:58	00:42	EcoCal	9.25	1984	1296	-		-	5.00	138	206
8	9/7 18:04	9/7 20:04	02:00	EcoCal	9.25	1978	1293	0.14	22N	-	4.49	57	367
9	9/7 20:34	9/7 21:18	00:44	EcoCal	9.24	1960	1348	0.14	22N	-	4.34	45	229
10	9/7 21:20	9/8 00:07	02:47	EcoCal	9.25	1968	1326	-		-	4.30	82	161
11	9/8 16:13	9/8 16:56	00:43	EcoCal	9.25	1969	1331	-		92.6	4.25	98	277
12	9/8 17:02	9/8 17:23	00:21	EcoCal	9.25	1978	1361	-		-	4.26	95	325
13	9/8 17:28	9/8 17:37	00:09	EcoCal	9.25	1982	1358	-		47.8	4.21	98	310
14	9/8 17:45	9/8 18:09	00:24	EcoCal	9.25	1977	1340	-		93.8	4.34	102	298
15	9/8 18:11	9/8 18:34	00:23	EcoCal	9.24	1986	1342	0.07	22N	93.5	4.61	70	381
16	9/8 18:38	9/8 18:51	00:13	EcoCal	9.25	1990	1365	0.07	22N	-	4.23	75	360
17	9/8 18:53	9/8 19:08	00:15	EcoCal	9.24	1966	1368	-		-	4.70	101	283
18	9/8 19:08	9/8 19:19	00:11	EcoCal	9.11	2010	1360	-		-	3.29	126	557
19	9/8 19:23	9/8 19:48	00:25	EcoCal	8.89	1984	1354	0.07	22N	-	3.44	94	413
20	9/8 19:53	9/8 20:45	00:52	EcoCal	8.88	1988	1358	-		-	4.00	163	371
21	9/8 20:56	9/8 21:15	00:19	EcoCal	9.08	1975	1294	-		-	4.75	126	329
22	9/8 21:17	9/8 22:02	00:45	EcoCal	9.24	1989	1296	-		-	4.45	102	401
23	9/8 22:04	9/8 22:59	00:55	EcoCal	9.24	1980	1324	-		-	4.28	96	316
24	9/8 23:01	9/8 23:33	00:32	EcoCal	9.25	1963	1307	-		-	4.37	101	374
25	9/9 00:10	9/9 01:44	01:34	EcoCal	9.25	1946	1317	-		-	4.31	93	456
26	9/9 08:01	9/9 09:59	01:58	EcoCal	9.01	2013	1198	-		-	4.73	110	914
27	9/9 10:06	9/9 10:12	00:06	EcoCal	9.00	2009	1227	-		-	4.14	95	764
28	9/9 15:37	9/9 16:38	01:01	EcoCal	9.25	1989	1407	-		-	4.24	98	283
29	9/10 00:00	9/10 03:40	03:40	EcoCal	9.25	1984	1406	-		-	4.54	97	250
30	9/10 10:39	9/10 12:19	01:40	EcoCal	9.25	1993	1385	-		-	4.37	83	397
31	9/10 12:23	9/10 13:42	01:19	EcoCal	9.25	1981	1369	0.14	22N	-	4.69	51	556
32	9/10 13:46	9/10 14:44	00:58	EcoCal	9.25	2001	1358	0.08	22N	-	4.70	68	406
33	9/10 15:30	9/10 16:01	00:31	Marblewhite	9.25	1983	1381	-		-	4.32	88	343
34	9/10 21:38	9/10 22:06	00:28	Marblewhite	9.25	1988	1367	-		-	5.32	105	220
35	9/10 22:19	9/10 22:42	00:23	Marblewhite	9.25	1973	1377	-		-	4.87	112	382
36	9/11 00:01	9/11 00:57	00:56	Marblewhite	9.25	1988	1375	-		-	4.71	102	338
37	9/11 01:03	9/11 02:14	01:11	Marblewhite	9.25	1979	1409	-		-	3.67	75	633

Ammonia Injection

A 30% solution of aqueous ammonia was injected into the furnace on several occasions to assess its effect on reducing NO_x emissions and the impact on CO emissions. The average exit temperature, NO_x, and CO emissions for each test point are plotted in Figure 10-67.

The first sequence of ammonia tests was on September 7, Tests 1 – 9. In each of the tests with ammonia, the NO_x levels dropped from the prior baseline test. The drop was more pronounced in Test 8 and 9, where the ammonia was added higher in the combustor – level 22' vs. 12'. In Tests 2 and 8, the CO levels increased when ammonia was added. This did not occur in Tests 4 and 6. Note that in Test 9, the air staging was altered – some of the overfire air was added up higher at elevation 22'. This increased the temperature profile in the combustor, thus the CO dropped while the NO_x reduction continued with the ammonia addition.

Another sequence with ammonia and with steam injection took place on September 8, Tests 10 – 17. The first few tests with steam injection without ammonia had little impact on the NO_x or CO levels. Test 15, which included with steam and ammonia injection, showed a decrease in NO_x and an increase in CO. Each change was less than that in the earlier Test 8 where ammonia was injected at level 22', likely because the ammonia flow rate was half that of Test 8. In Test 16, the ammonia continued while the steam was shut off. There was little change in the NO_x or CO.

The final sequence with ammonia was on September 10, Tests 30 – 32, with a baseline, full ammonia flow, and half ammonia flow. The NO_x reduction and CO increase varied as expected with ammonia load. Ammonia slip was measured in Tests 31 and 32. No slip was observed.

Figure 10-67: NO_x and CO Emissions With Pittsburgh #8 Coal

Temperature

The NO_x and CO emissions are plotted versus the temperature at the top of the combustor in Figure 10-68. The NO_x emissions do not vary much with the exit temperature. The CO is similarly insensitive as long as the exit temperature is above about 1230°F.

Figure 10-68: Influence of Furnace Outlet Temperature on NO_x and CO Emissions With Pittsburgh #8 Coal

Effect of Overfire Air Variations

As mentioned above, the spreading out of the overfire air in Test 9 resulted in a slight additional drop in NO_x with ammonia injection. The primary air flow was varied on September 8, Tests 17 – 19. In Test 18, the proportion of primary air through the grid plate was increased. With less deep staging, the NO_x increased over Test 17, as expected. With the addition of ammonia in Test 19, the NO_x dropped back down.

Curiously, the CO increased with higher primary air and decreased in this case with added ammonia.

10.6.2. Environmental Performance with Waynesburg Bituminous Coal

Table 10-8 summarizes the emissions during November 2003 test campaign with Waynesburg bituminous coal. Tests were done with both EcoCal (Tests 1 – 14) and Marblewhite (Tests 15 - 45) limestones. The NO_x and CO emissions are both normalized to 3% O₂. The furnace temperatures, NO_x and CO emissions are plotted for each test in Figure 10-69.

Table 10-8: NO_x and CO Emissions With Waynesburg Coal

Test	Start	End	Dur.	Fuel	Sorbent	Firing Rate	T Bed	T Top	NH ₃ Flow	Steam flow	% O ₂ dry	N3NO _x	N3CO	N3SO ₂
1	11/3 20:14	11/3 20:40	00:25	W	EC	8.89	2007	1105	-	-	4.70	264	301	1625
2	11/4 00:16	11/4 00:32	00:16	W	EC	8.89	1982	1105	-	-	4.99	268	398	1572
3	11/4 00:53	11/4 01:24	00:30	W	EC	8.89	1986	1096	-	-	5.39	273	393	1496
4	11/4 04:12	11/4 04:29	00:16	W	EC	8.89	1966	1122	0.08	-	5.10	172	397	1597
5	11/4 04:36	11/4 04:46	00:10	W	EC	8.89	1959	1126	0.09	-	5.25	163	470	1609
6	11/4 04:55	11/4 05:05	00:10	W	EC	8.89	1937	1126	0.11	-	5.37	166	503	1753
7	11/4 05:20	11/4 05:30	00:10	W	EC	8.89	1985	1127	0.11	-	5.42	186	521	1876
8	11/4 16:47	11/4 16:56	00:09	W	EC	8.88	1976	1210	0.10	-	5.11	96	618	2004
9	11/4 17:07	11/4 17:12	00:04	W	EC	8.88	1987	1195	0.12	-	4.92	90	638	2117
10	11/4 17:15	11/4 17:20	00:05	W	EC	8.88	1990	1195	0.12	-	5.19	90	641	2056
11	11/4 17:21	11/4 17:29	00:08	W	EC	8.88	1993	1210	0.12	-	4.83	98	690	2033
12	11/4 17:45	11/4 18:15	00:30	W	EC	8.88	1983	1216	-	-	4.89	130	499	1812
13	11/4 18:31	11/4 18:39	00:08	W	EC	8.88	1968	1289	-	-	5.13	118	637	1666
14	11/4 18:47	11/4 18:58	00:11	W	EC	8.88	1974	1278	-	-	4.97	127	490	1696
15	11/4 22:53	11/4 23:03	00:09	W	MW	9.42	1995	1218	-	-	4.28	111	659	1739
16	11/5 01:40	11/5 01:50	00:09	W	MW	9.42	2005	1239	-	-	3.78	102	583	1903
17	11/5 02:06	11/5 02:19	00:13	W	MW	9.43	1990	1245	0.13	-	3.91	78	688	1868
18	11/5 03:46	11/5 03:53	00:07	W	MW	9.42	2016	1249	0.13	-	4.23	90	678	1395
19	11/5 03:55	11/5 04:01	00:06	W	MW	9.42	1989	1300	0.13	-	4.07	76	790	1439
20	11/5 04:03	11/5 04:19	00:15	W	MW	9.42	1969	1295	0.13	-	3.83	75	638	1485
21	11/5 10:15	11/5 10:30	00:14	W	MW	9.42	1984	1265	-	-	3.56	164	230	1719
22	11/5 11:00	11/5 11:15	00:14	W	MW	9.42	1985	1291	-	-	3.06	118	256	1699
23	11/5 11:40	11/5 11:50	00:09	W	MW	9.42	1988	1295	-	-	3.48	138	208	1533
24	11/5 12:10	11/5 12:20	00:10	W	MW	9.42	1956	1313	-	-	3.35	117	203	1517
25	11/5 13:27	11/5 13:37	00:09	W	MW	9.42	1921	1415	0.11	-	3.14	77	254	1256
26	11/5 13:45	11/5 13:55	00:09	W	MW	9.42	1930	1451	0.11	-	3.51	86	223	1204
27	11/5 14:01	11/5 14:10	00:08	W	MW	9.42	1979	1425	0.11	-	3.21	97	228	1407
28	11/5 14:19	11/5 14:29	00:09	W	MW	9.42	1971	1459	0.11	-	3.64	98	222	1352
29	11/5 14:37	11/5 14:42	00:04	W	MW	9.42	1982	1368	0.11	-	3.64	120	235	1228
30	11/5 15:09	11/5 15:19	00:09	W	MW	9.42	1991	1357	-	-	3.53	139	196	1460
31	11/5 16:20	11/5 16:40	00:19	W	MW	9.42	1907	1433	0.11	-	3.91	124	124	1257
32	11/5 16:49	11/5 16:53	00:04	W	MW	9.42	1899	1370	0.11	-	3.63	128	108	1343
33	11/5 17:08	11/5 17:23	00:14	W	MW	9.42	1935	1357	0.11	-	3.92	120	119	1300
34	11/5 19:01	11/5 19:14	00:12	W	MW	9.42	1995	1348	-	-	3.70	132	247	1415
35	11/5 20:00	11/5 20:27	00:27	W	MW	9.42	2008	1346	-	42.2	4.09	149	231	1404

36	11/5 20:34	11/5 20:49	00:14	W	MW	9.42	2010	1330	-	-	3.75	145	307	1417
37	11/5 21:16	11/5 21:44	00:27	W	MW	9.42	2001	1390	-	61.0	4.01	145	277	1380
38	11/5 23:46	11/6 00:14	00:27	W	MW	8.53	1970	1371	-	-	5.96	205	191	1188
39	11/6 00:35	11/6 01:02	00:27	W	MW	8.54	1970	1378	-	-	5.55	194	167	1280
40	11/6 01:09	11/6 01:13	00:04	W	MW	8.56	1951	1362	-	-	6.29	242	152	1333
41	11/6 02:01	11/6 02:29	00:27	W	MW	8.54	1984	1372	-	-	5.68	195	173	1288
42	11/6 02:42	11/6 02:54	00:12	W	MW	8.53	1977	1441	-	-	6.09	220	181	1252
43	11/6 04:01	11/6 04:14	00:12	W	MW	8.54	1959	1443	-	-	5.56	188	175	1246
44	11/6 10:15	11/6 10:28	00:12	W	MW	9.42	1956	1357	-	-	4.37	211	486	1547
45	11/6 11:00	11/6 11:19	00:18	W	MW	9.42	1967	1351	-	-	3.81	218	424	1534
46	11/6 13:15	11/6 13:27	00:12	PC	MW	9.85	1969	1233	-	-	5.93	189	220	2328
47	11/6 13:51	11/6 14:10	00:18	PC	MW	9.85	1937	1283	-	-	5.41	164	217	2300
48	11/6 14:12	11/6 14:18	00:06	PC	MW	9.85	1918	1362	-	-	4.85	140	190	2135
49	11/6 14:20	11/6 14:29	00:08	PC	MW	9.85	1936	1436	-	-	4.21	115	142	1882
50	11/6 14:33	11/6 14:48	00:14	PC	MW	9.84	1976	1468	-	-	3.87	118	103	2159
51	11/6 16:01	11/6 16:29	00:27	PC	MW	10.16	1975	1346	-	-	5.23	174	161	2209
52	11/6 17:40	11/6 18:01	00:21	PC	MW	10.16	2004	1298	0.12	-	5.68	107	259	2361
53	11/6 18:16	11/6 18:24	00:08	PC	MW	10.18	1981	1283	0.12	-	7.43	105	397	2690
54	11/6 18:26	11/6 18:35	00:08	PC	MW	10.17	1945	1262	0.12	-	6.72	93	412	2289
55	11/6 18:37	11/6 18:49	00:12	PC	MW	10.17	1987	1259	0.12	-	6.05	94	343	2375
56	11/6 18:52	11/6 18:54	00:02	PC	MW	10.17	2011	1278	0.12	-	5.78	87	321	2307
57	11/6 18:56	11/6 19:00	00:04	PC	MW	10.18	2008	1278	0.12	-	5.55	77	263	2388
58	11/6 19:23	11/6 19:42	00:18	PC	MW	10.16	2005	1258	-	-	5.54	154	173	2431
59	11/6 20:21	11/6 20:40	00:18	PC	MW	10.17	1994	1264	-	-	5.48	172	177	2343
60	11/6 20:45	11/6 20:59	00:14	PC	MW	10.17	1995	1298	0.09	-	5.23	107	260	2367
61	11/6 21:10	11/6 21:33	00:23	PC	MW	10.16	1994	1308	0.12	-	5.16	93	315	2353
62	11/6 21:36	11/6 21:46	00:10	PC	MW	10.16	1992	1312	0.12	-	4.84	85	320	2406
63	11/6 21:48	11/6 22:09	00:21	PC	MW	10.19	1960	1335	0.12	-	4.78	79	325	2390
64	11/6 22:16	11/6 22:41	00:25	PC	MW	10.38	1966	1366	0.12	-	4.47	75	247	2306
65	11/6 22:45	11/6 22:58	00:12	PC	MW	10.39	1990	1386	0.12	-	4.20	71	209	2254
66	11/6 23:21	11/6 23:41	00:20	PC	MW	10.38	1998	1408	0.12	-	4.08	75	252	2199
67	11/7 00:00	11/7 00:30	00:29	PC	MW	10.39	2003	1404	0.09	-	3.88	74	192	2304
68	11/7 01:12	11/7 01:20	00:08	PC	MW	10.39	2005	1419	0.09	-	3.76	58	192	2273
69	11/7 01:28	11/7 01:49	00:20	PC	MW	10.38	1992	1419	0.24	-	3.86	60	215	2363
70	11/7 02:14	11/7 02:29	00:14	PC	MW	10.38	1989	1407	0.12	-	4.13	69	153	2317
71	11/7 02:46	11/7 03:03	00:16	PC	MW	10.39	1989	1409	0.12	74.0	4.31	71	157	2221
72	11/7 03:07	11/7 03:24	00:17	PC	MW	10.38	1972	1448	0.12	84.1	4.43	73	151	2160
73	11/7 03:50	11/7 04:09	00:19	PC	MW	10.38	2004	1425	-	84.1	3.69	108	160	2313
74	11/7 04:31	11/7 04:50	00:19	PC	MW	10.38	2007	1439	-	55.9	3.49	108	124	2306
75	11/7 06:01	11/7 06:55	00:54	PC	MW	10.38	1990	1403	-	-	3.62	116	131	2229

Figure 10-69: NO_x, CO Emissions, and Temperature With Waynesburg Coal

Ammonia Injection

There were several sequences of ammonia injection with Waynesburg coal. After the first three baseline tests, Tests 4-7 were done using the 1765 injector tip with a 65° spray angle. This reduced the NO_x from about 270 to about 170 ppm (all ppm expressed normalized to 3% O₂ in the flue gas).

The next tests (8 – 12) used the 1200 injector with a 0° spray angle, but with also a change in overfire air staging. The NO_x dropped to below 100 ppm. The CO jumped from about 500 to over 600 ppm, despite the increase in furnace exit temperature due to the staging change. Comparing Tests 25 – 29 with tip 1200 and Tests 31 – 33 with tip 1765, we again see that the 0° tip has much lower CO (about 110 vs. 220 ppm). The effect on NO_x is again unclear since the NO_x was rising during Tests 25-29.

Steam Injection

Steam was injected without ammonia at elevation 12 (Test 35) and elevation 22 (Test 37). There was no apparent effect on the NO_x emissions and at most a very small drop in CO with the added steam.

Temperature

The NO_x and CO emissions are plotted versus the temperature at the top of the combustor in Figure 10-70. The first seven test points had very low air staging; the combustion occurred low in the furnace. This accounted for the high NO_x and the low CO. The other tests had the combustion more spread out. The expected trend of higher NO_x (without ammonia) and lower CO with increasing temperature is seen in the figure.

Figure 10-70: NO_x and CO Emissions vs. Furnace Exit Temperature

10.6.3. Environmental Performance with Petroleum Coke

Table 10-8 also summarizes the emissions during the November 2003 test campaign with petroleum coke. Tests were done with Marblewhite limestone. The furnace temperatures, NO_x and CO emissions are plotted for each test in Figure 10-71.

Figure 10-71: NO_x, CO Emissions, and Temperature With Petroleum Coke

Ammonia Injection

There were two long sequences of ammonia injection with petroleum coke. Tests 52 - 57 were done using the 1200 injector tip with a 0° spray angle. This reduced the NO_x to below 100 ppm and in Test 57 reached 77 ppm, which is 0.11 lb/MMBtu. Tests 60 – 67 used the 1715 injector tip with a 15° spray angle; Test 68 – 72 used the 65° 1765 tip. Test 69 also used the 15° tip with ammonia at two locations. These test got the NO_x down to below 60 ppm, or about 0.083 lb/MMBtu. The CO was lowest with the 1765 tip, down to about 150 ppm or 0.13 lb/MMBtu.

Steam Injection

Steam was injected in Tests 71 and 72 with ammonia and Tests 73 and 74 without ammonia. There was no obvious effect on NO_x or CO emissions.

Temperature

The NO_x and CO emissions are plotted versus the temperature at the top of the combustor in Figure 10-72. In this case, the NO_x also decreased with increasing furnace exit temperature.

Figure 10-72: NO_x and CO Emissions vs. Furnace Exit Temperature

10.7. Sulfur Capture

Another objective of the MTF test campaigns was to assess the sulfur capture performance of the CMB system in both the combustor and in the backend NID system. The combustor was intended to primarily calcine the sorbent, which would then be used in the NID system for sulfur capture. Significant sulfur capture was not expected in the combustor due to the very high temperatures in the lower combustor. However, the sorbent does pass through a temperature window where sulfur capture can take place.

The NIDTM system is basically a modern version of spray drying technology. One of the main controlling process parameters is the humidity in the system, which is controlled by the water amount added to the NID mixer. Commercially, a NID system is normally operated at a relative humidity around 50%. Higher humidity's were also tested in this program to assess its effect on sulfur capture.

The NID system is operated with a very large solids inventory that takes a long time to replace and to get to steady state conditions. Due to the available pilot plant test time, many tests were conducted for shorter durations with the objective to find trends rather than running longer terms to establish actual steady state values.

10.7.1. Limestones Tested in the MTF

During the September and November 2003 test campaigns, two limestones were evaluated: Eco-Cal 5050 and Marblewhite 325, which are two sizes of North Adams limestone from western Massachusetts supplied by Specialty Minerals.

The size distributions for these sorbents are shown in Figure 10-73.

Figure 10-73: Sorbent Particle Size Determined By CILAS Laser Technique

10.7.2. Carryover of the Calcium to the Backend NID™ System

The sulfur capture results from the Phase I program showed that the sulfur capture performance in the NID system was limited by how much calcium oxide was carried over to the backend equipment. The MTF cyclone was very efficient and captured a large portion of the recirculating calcium products. Some of the ash included significant amounts of calcium that then had to be drained from the system to maintain a proper solids inventory in the combustor. The utilization of the calcium in the drain was not very high because of the high temperatures in the combustor. This limited the overall sulfur performance. An objective of the Phase II test campaign was therefore to try to improve the overall sulfur capture performance by increasing

the carryover of calcium to the NID system. This was first accomplished by reducing the size of the initial sorbent from a D_{50} of 50 microns for the EcoCal 50/50 limestone down to a D_{50} of 9 microns for the Marblewhite 325 limestone. Further attempts to increase the sorbent carryover were also made by intentionally disrupting the cyclone to decrease its collection efficiency.

Pseudo-isokinetic samples were taken of the flyash leaving the furnace cyclone throughout the MTF tests. The flyash flow rate was then determined from the amount of sample collected over the sample duration. Past results showed that this calculated flyash flow rate matched reasonably well with the flow rate determined by other methods, including mass balance on the furnace, long-term solids collection in the baghouse, or other isokinetic sampling methods (i.e. EPA Method 5). Nevertheless, the results were considered here mainly for approximate values.

Several of these flyash samples were analysed for calcium and sulfur. The amount of calcium (total and unreacted) that was carried over to the NID system was calculated from these analyses and from the calculated flyash flow rates. Figure 10-74 shows the amount of calcium in the flyash as a fraction of the total calcium fed with the limestone for the September 2003 test with the Pittsburgh #8 bituminous coal. The two samples from early September 9th were taken when the cyclone efficiency was deliberately degraded in order to increase the amount of calcium carried over to the NID system. The overall sulfur capture for these two tests jumped from about 70% to over 90% due to the increased calcium carryover. Apart from these two samples, the carryover was also higher with the finer Marblewhite 325 limestone.

The utilization of the calcium in the flyash samples is shown in Figure 10-75, along with the utilization of the baghouse drain, which represents the additional capture in the NID system.

The calcium carryover and utilization from the November test with Waynesburg bituminous coal and petroleum coke are shown in Figure 10-76 and Figure 10-77.

Figure 10-74 Fractional Calcium Carryover For Pittsburgh #8

Figure 10-75 Calcium Utilization Of Flyash And Baghouse Drain For Pittsburgh #8

Figure 10-76 Fractional Calcium Carryover For Waynesburg Coal And Petroleum Coke

Figure 10-77 Calcium Utilization Of Flyash And Baghouse Drain For Waynesburg And Petroleum Coke

10.7.3. Sulfur Capture Performance

Figure 10-78 shows the relative humidity at the NID outlet and the sulfur capture for each test point during the September test with Pittsburgh #8 bituminous coal. Tests 1 – 32 were with the Eco-Cal 50/50 limestone; Tests 33 – 37 were with the finer Marblewhite 325 limestone.

Figure 10-78 NID Relative Humidity And Sulfur Capture Performance For Pittsburgh #8

The calcium to sulfur mole ratio was 2:1 for the first 25 tests; at Test 26 it was raised to 2.2:1 for the rest of the week. The sulfur capture in the combustor was roughly 40% to 50% for most of the test points – other than Tests 9 and 10 with some of the overfire air staged higher to elevation 22 ft. The overall sulfur capture reached 60 – 70% through Test 25. (Note that the SO₂ analyzer was not working during Tests 18 – 20). Tests 26 and 27 reached 80% relative humidity in the NID system and the overall sulfur capture exceeded 90%. With the finer Marblewhite 325

limestone in the last 5 tests, the overall sulfur capture reached 98% when the relative humidity was about 80%.

The NID relative humidity and the sulfur capture performance for the November 2003 test are shown in Figure 10-79 for the Waynesburg bituminous coal and Figure 10-80 for the petroleum coke.

Figure 10-79 NID Relative Humidity And Sulfur Capture Performance For Waynesburg Bituminous Coal

Tests 1 – 14 with the Waynesburg coal were with the coarser Eco-Cal limestone; the remaining tests with the Waynesburg coal and the petroleum coke used the Marblewhite 325 limestone.

Figure 10-80 NID Relative Humidity And Sulfur Capture Performance For Petroleum Coke

The calcium-to-sulfur ratio was 2:1 for all the November tests with the Waynesburg bituminous coal and petroleum coke. The in-furnace sulfur capture reached 30 - 40% with the Waynesburg coal and 40% with pet coke.

The Waynesburg coal was similar to the Pittsburgh #8 coal in that the overall sulfur capture exceeded 90% with the coarser Eco-Cal limestone at a relative humidity of 80% and reached 99% with the Marblewhite 325 limestone. With the petroleum coke, the sulfur capture also reached 99% at 80% relative humidity.

11. COMMERCIAL PLANT CONCEPT

The CMB was originally proposed as a cost effective design for high temperature Rankine cycles. The concept was also adopted for nearer term applications as a low cost CFB alternative with conventional steam conditions. ALSTOM Power established a goal of developing a nearer term CMB that had boiler island material costs that were 30% less expensive than a conventional CFB.

The objective of this task was to develop an updated commercial concept of the nearer term CMB boiler, including boiler island arrangement and costs. Two cost comparison studies were completed during this phase – a preliminary 300Mw CMB design and then a more detailed 2x150Mw_e CMB design.

11.1. 300Mw_e CMB Commercial Boiler

The preliminary cost comparison study involved a comparison between a 300Mw_e CMB and a commercial CFB boiler design. The steam conditions for this study were 2520psig/1004°F/1004°F. This study included the evaluation of a wide range of CMB arrangements, with the intent of identifying a configuration that would then be used for a more detailed cost study. Figure 11-1 shows thumbnail figures of many of the CMB arrangements that were considered during this study. A brief summary of the arrangements is listed in Table 11-1.

Figure 11-1: 300Mw_e CMB Boiler Arrangements

Table 11-1: 300Mw CMB Boiler Arrangements

Configuration	Description
Base	1x300Mw _e rectangular combustor, single MBHE, 3 cyclones
Alt 2	1x300Mw _e rectangular combustor, 3 MBHEs, 3 cyclones
Alts 3-4	3x100Mw _e rectangular combustor, 3 MBHEs (tall vs. squat), 3 cyclones
Alts 5-6	1x300Mw _e cylindrical combustor, 2 MBHEs, 3 cyclones (in 2 arrangements)
Alts 7-8	1x300Mw _e cylindrical combustor, 1 MBHE, 3 or 6 cyclones
Alts 9, 11, 13	1x300Mw _e cylindrical tapered combustor, 1 MBHE, 3,4, or 6 cyclones
Alts 10,12,14,15	1x300Mw _e cylindrical combustor, 1 MBHE, 4 or 6 cyclones, support structure variations

Preliminary costs were then developed for the boiler material costs for the various CMB designs and for a conventional 300Mw_e CFB. A key assumption was that both the CMB and CFB designs would require a backend NID system. This was a reasonable assumption since the design sulfur capture requirement was 98%. Figure 11-2 shows the resulting relationship between total pressure parts weight and the estimated CMB boiler material savings percentage compared to the CFB design. The various points on the trend line represent the cost savings for different CMB arrangements relative to the CFB design and for different assumptions related to heat transfer performance. The study showed that the CMB boiler material costs were between 19 – 31% lower than the CFB design. The CMB auxiliary power was about 0.6% less than for the CFB and the O&M costs were about the same for both designs. No erection issues were identified. Figure 11-3 compares a side elevation view of the CFB and CMB (Alternate 14) boilers.

Figure 11-2: 300Mw_e CMB Boiler Material Cost Savings vs CFB

Figure 11-3: 300Mw_e Side Elevation Comparison

11.2. 2x150MW_e CMB Commercial Plant Design

A more detailed CMB design study was completed in 2005 for a 2x150MW_e plant arrangement. The steam conditions for this study were 2130psig/1009°F/1009°F. This study built upon the previous CMB cost study and also upon the results from the CMB Phase I results. Five CMB designs were considered with varying degrees of combustion gas cooling in the combustor. Seven different sulfur control technologies were evaluated for each CMB configuration. A CO control system was not included at this time. All of these CMB designs were then compared against a 2x150MW_e CFB design, both without and with a backend NID system.¹⁸

11.2.1. 150MW_e CFB and CMB Design Basis

Fuel Analysis

The fuel selected for this study was a lignite that was used in a previous ALSTOM Power study, but had its sulfur content artificially increased. This fuel had an uncontrolled SO₂ level (if not captured in the furnace or downstream of the boiler) of 5478ppm. The as-fired fuel analysis used in this study is shown in Table 11-2.

Table 11-2: Fuel Analysis (as-fired)

	Wt %
Carbon	31.6
Hydrogen	3.8
Oxygen	14.7
Nitrogen	0.3
Sulfur	3.2
Ash	14.6
Moisture	31.8
HHV (Btu/lb)	6040

SO₂ Emission Control

SO₂ emission control was an important parameter for this study. Given the target sulfur removal requirement of 98% and the high sulfur content in the fuel, both the CMB and CFB would need some form of SO₂ emission control downstream of the boiler. The CFB boiler could meet this target with either a CFB with only in-furnace sulfur capture (but high Ca/S mole ratio) or by a CFB with a backend NID system. As discussed in Section 8.4, a range of SO₂ emission control technologies were considered in this study for the CMB cases. They included commercial technologies (dry and wet scrubbers) and various combinations of technologies under development (NID system with high relative humidity, steam reactivation, Cyclone NID). Table 11-3 shows the expected sulfur capture performance (total and in-furnace), required Ca/S ratio, and resulting boiler efficiency for each combination.

¹⁸ CMB Proof of Concept Design Review Issues – Technology Approach and Cost Analysis, Glen Jukkola, ALSTOM Power Internal Report, August 2009

Table 11-3: 2x150Mw_e Study Design Conditions

	CFB No NID Baghouse	CFB With NID	CMB NID Hi H ₂ O	CMB Stm Reac NID
Coal Sulfur - Uncontrolled SO ₂ (ppm)	5478	5478	5478	5478
Total Sulfur Capture %	97.5	98	98.1	99.1
Furnace Sulfur Capture %	97.5	92.5	50	92.5
Sorbent	CaCO ₃	CaCO ₃	CaCO ₃	CaCO ₃
Ca/S Ratio	2.75	1.95	1.95	1.95
Backend SO ₂ System	None	NID	NID	NID
Boiler Efficiency	81.23%	82.31%	80.67%	82.31%

	CMB Bagfilter Wet Scrub	CMB Dry Scrub Lime	CMB Cyc NID CaO/Hi H ₂ O	CMB Cyc NID Hi H ₂ O	CMB Stm React Cyc NID
Coal Sulfur - Uncontrolled SO ₂ (ppm)	5478	5478	5478	5478	5478
Total Sulfur Capture %	98	98	98	98	98
Furnace Sulfur Capture %	0	0	0	50	92.5
Sorbent	CaCO ₃	CaO	CaO	CaCO ₃	CaCO ₃
Ca/S Ratio	1.40	1.80	1.80	1.95	1.95
Backend SO ₂ System	Wet Scrubber	Dry Scrubber	CycNID	CycNID	CycNID
Boiler Efficiency	82.14%	82.14%	82.14%	80.67%	82.31%

CFB Boiler Configuration

Two CFB boiler configurations were evaluated for this study. One was a CFB without any backend sulfur control and the other was a CFB with a NID system. Both configurations were 2x150Mw_e. Each CFB was designed with 20% excess air, a Lungstrom airheater, and a 287°F gas temperature leaving the airheater.

CMB Boiler Configuration

Five CMB boiler configurations were evaluated for this study. Each one was 2x150Mw. The main variation between cases was the furnace gas outlet temperature. It was varied from 630°F (original CMB design) to 1300°F. As discussed in Section 8.3, the higher gas outlet temperature may be needed to control CO emissions leaving the boiler. Increasing the gas outlet temperature had significant impacts on the CMB arrangement, as all of the heat transfer duty could no longer be done in the MBHE.

Case 1 (Base) was the original CMB design with a 630°F gas temperature leaving the furnace. All of the heat transfer duty took place in the MBHE. This case was the tallest design, with a furnace height of 100ft. Case 2 had a 750°F gas temperature leaving the furnace. This arrangement required a small backpass with a partial economizer. Case 3 had a 908°F gas outlet temperature. This arrangement required a backpass with a full economizer. Case 4 had a 1159°F gas outlet temperature. This arrangement required a backpass with a LTSH and full economizer. Case 5 had a 1300°F gas outlet temperature. This arrangement required a backpass with a LTSH, LTRH, and full economizer. This case was the most compact design,

with a furnace height of 65ft. Figure 11-4 compares the boiler sizing of the 630°F and 1300°F CMB cases.

Figure 11-4: Comparison of CMB 630°F and 1300°F Boiler Sizing

All of the CMB cases used a once through steam/water system and a Lungstrom airheater. Cases 1 through 4 used both solids temperature control and LTRH with steam bypass for the reheat steam temperature control. Case 5 used a split backpass for reheat steam temperature control.

CMB MBHE Design and Steam/Water Circuits

All of the MBHEs were designed with the same approach. The MBHEs were designed with water-cooled bounding walls and hanger tubes. All of the vapour-heated sections used stainless steel material. A once through steam/water system was used for each design. The MBHE used spiral-finned tubing in a staggered arrangement. The tubes were 1.75 inch OD and had 2 fins per inch. The fins varied from 0.5" to 0.75" and were made of 304SS with a 1400°F oxidation limit.

Figure 11-5 shows the MBHE surface arrangement for the 630°F and 1300°F CMB designs. All of the tube sections had 48 assemblies in a staggered arrangement with transverse spacing of 6.25" and a longitudinal spacing of 3.0". Each staggered assembly was supported by hanger

tubes at 3 locations. Figure 11-6 and Figure 11-7 show the steam/water circuits for the 630°F case (without any backpass) and the 1300°F case (with a split backpass).

Figure 11-5: MBHE Arrangement for 630°F and 1300°F CMB Designs

Figure 11-6: 150MW_e CMB Steam & Water Flow Schematic – 630°F Case

Table 11-4: Comparison of 150Mw_e CFB vs CMB Performance

		CFB Regular Sorbent With NID	CMB - Base Fine Sorbent With NID
<u>Combustor Performance</u>			
Ca/S		1.95	1.95
Combustor Sulfur Capture	%	92.5	50.0
Combustor Utilization	%	47.0	26.0
Boiler Efficiency	%	82.31	80.67
<u>NID Performance</u>			
NID Sulfur Capture	%	73.3	96.0
<u>Total Performance</u>			
Percent Sulfur Capture	%	98.0	98.1
<u>Air Flows to Boiler</u>			
Primary Air Flow	lbm/hr	684,078	943,849
Secondary Air Flow	lbm/hr	546,874	353,169
Fluidizing Air Flow	lbm/hr	69,355	40,573
Infiltration Air Flow	lbm/hr	19,802	0
Total Combustion Air	lbm/hr	1,320,109	1,337,591

Note: CMB Bauxite Transport System PA required flow is 242,647 lbm/hr

Table 11-5 compares the fluidizing air requirements for the FBHE, MBHE, and seal pots for the two designs, while Table 11-6 compares the air pressure requirements. These tables show that the MBHE requires considerably less airflow than a FBHE. Furthermore, the MBHE airflow for solids transport can be provided by a PA fan while the FBHE airflow must be provided by a high pressure fluidizing air (FA) blower.

Table 11-5: Comparison of 150Mw_e CFB vs CMB Fluidized Air Flow Requirements

		CFB	CMB - Base
Seal pot	lb/hr	26,555	9,072
Total grease	lb/hr	12,320	6,160
RH FBHE - nozzles	lb/hr	20,200	0
RH FBHE - related	lb/hr	8,431	0
MBHE Flow Distribution	lb/hr	0	25,341
Total	lb/hr	67,506	40,573

Table 11-6: Comparison of 150Mw_e CFB vs CMB Air Pressure Requirements

<u>Combustor Pressure Profile</u>		CFB	CMB - Base	Delta CMB -CFB
At grate	in-wg	43.0	36.0	-7.0
At secondary air port	in-wg	35.0	5.0	-30.0
<u>Fan Pressure Requirements - both units have Lungstrom</u>				
PA	in-wg	79.17	68.6	-10.6
PA used for bauxite transport	in-wg	n/a	80.6	
PA required	in-wg	79.17	80.6	1.4
SA	in-wg	58.77	28.77	-30.0
FA for FBHE	in-wg	239.0	n/a	
FA for seal pots and grease	in-wg	222.2	93.8	
FA for bauxite distribution	in-wg	n/a	100.0	
FA required pressure	in-wg	239.0	100.0	-139.0

Table 11-7 compares the fan power requirements for the two designs. Although the CMB design requires more airflow from the PA fan for transporting the bauxite, the overall fan power consumption is about 8% less for the CMB design.

Table 11-7: Comparison of 150Mw_e CFB vs CMB Power Consumption (both with Lungstrom Air Heater)

		CFB	CMB - Base
<u>Fan Pressure Head</u>			
PA	in-wg	79.17	80.60
SA	in-wg	58.77	28.77
FA	in-wg	239.00	100.00
ID	in-wg	-17.74	-18.21
<u>Fan Inlet ACFM</u>			
PA	ft3/min	191,002	251,951
SA	ft3/min	123,158	78,639
FA	ft3/min	16,272	9,519
ID	ft3/min	581,560	547,352
<u>Fan Power</u>			
PA	kW	2,723	3,656
SA	kW	1,216	380
FA	kW	655	160
ID	kW	1,955	1,884
Total	kW	6,549	6,080

11.2.3. 150Mw_e CMB Boiler Arrangements

General arrangement drawings were developed for each of the CMB boiler designs and for the CFB boiler design. Figure 11-8 shows the side elevation view of the CFB boiler and the CMB boilers with combustor gas outlet temperatures of 1300°F, 1130°F, 908°F, and 630°F. Figure 11-9 shows the plan view for the same four CMB cases. Each of the CMB cases used a cylindrical combustor with four cyclones. As the combustor gas outlet temperature increased, the duty in the MBHE decreased and the size of the backpass increased. At the same time, the height of both the MBHE and combustor decreased.

Figure 11-8: 150Mw_e CFB vs CMB Boiler Side View

Figure 11-9: 150Mw CMB Plan View

Figure 11-10 shows an isometric view of the CMB cases with combustor gas outlet temperatures of 1300°F, 1130°F, 908°F, and 630°F. Not shown in these views are the following items:

- Fans, blowers, pumps
- Air ducting from the PA and SA to the Ljungstrom Airheater
- Gas ducting from the Ljungstrom Airheater
- Fuel and limestone silos and handling systems
- Bauxite bunker and makeup system
- Furnace drains, standpipes to the MBHE, rock classifiers, bauxite distributors
- Ash removal systems

- Startup burners, miscellaneous air/gas piping, solids valves
- Support steel and platforms
- Sootblower systems

Figure 11-10: 150Mw_e CMB Isometric View

11.3. 150Mw_e CMB Commercial Plant Cost Comparison

11.3.1. 150Mw_e CMB Material Weights

The three-dimensional CMB designs were used for developing weights and costs for the steel, refractory, pressure parts, and insulation/lagging. Figure 11-11 and Figure 11-12 show an isometric view of the metal surfaces and refractory surfaces for each CMB case. As the CMB combustor gas outlet temperature increases, the amount of structural and fabricated steel decreases because both the combustor and MBHE height become shorter, although a partial backpass becomes necessary. The total pressure part weight also is significantly reduced at the higher gas outlet temperatures. However, the cyclones become much larger and considerably more refractory is required for the combustor, cyclones, and gas outlet ducts.

Figure 11-11: 150Mw_e CMB Isometric View - Metal Surfaces Only

Figure 11-12: 150Mw_e CMB Isometric View - Refractory Only

Table 11-8 compares the percentage weight savings of the CMB boiler island material weights relative to the CFB design. The CMB has significantly lower weights than the CFB design, with the largest weight savings in pressure parts and also in refractory for the low temperature CMB cases. The CMB boiler design with a 630°F combustor outlet temperature weighs 32.8% less (not including the bauxite heat transfer media) than the CFB design. The weight savings in the CMB with the 1300°F combustor outlet temperature is reduced to 21.1% because of the much

larger requirement for refractory. Note that the CMB total weight savings compared to the CFB case is reduced to about 10% for all CMB cases when the weight of the bauxite for heat transfer media is considered.

Table 11-8: 150MW_e CMB Boiler Island Material % Weight Differential Relative to CFB

	630F	908F	1139 F	1300F
Summary	CMB	CMB	CMB	CMB
Pressure Parts	-51.9%	-56.6%	-46.1%	-44.5%
Structural Steel	-9.0%	-9.6%	-8.8%	-8.4%
Fabricated Steel	-34.1%	-28.9%	-35.9%	-37.3%
Refractory & Insulation	-51.9%	-29.6%	-28.3%	-15.2%
Additional Purchased Items	-51.2%	-40.9%	-34.9%	-31.2%
Additional Mechanical Items	0.0%	0.0%	0.0%	0.0%
Total Equipment	-32.8%	-26.4%	-25.0%	-21.1%
Total Equipment + Bauxite	-9.1%	-9.8%	-10.1%	-9.5%

11.3.2. 150MW_e CMB Boiler Island Material Costs

Table 11-9 compares the percent CMB boiler island material cost savings relative to the CFB design for most major cost categories. The CMB design shows significant savings in pressure parts, fabricated steel, and refractory when compared with the CFB design. The summary does not include any costs for backend SO₂ capture, limestone preparation and handling, nor any costs associated with additional CO reduction. The cost of the sintered bauxite as the heat transfer media is included in the second total line. The CMB with the 630°F combustor outlet temperature has material costs (including heat transfer media) that are 22.7% less than the CFB design. The CMB material savings for the 1300°F gas outlet temperature were only 15.1% less than the CFB design, largely because of the large increase in refractory weight. None of the CMB designs evaluated in this study met the 30% materials cost savings target against the CFB design.

Table 11-9: 150Mw_e CMB Boiler Materials % Cost Differential Relative to CFB

	630F CMB	908F CMB	1139 F CMB	1300F CMB
Summary				
Pressure Parts	-31.2%	-39.1%	-27.0%	-27.4%
Structural Steel	-9.6%	-7.9%	-2.6%	2.3%
Fabricated Steel	-30.3%	-17.6%	-31.7%	-32.8%
Refractory & Insulation	-53.6%	-31.5%	-26.0%	-10.0%
Additional Purchased Items	-65.4%	-58.1%	-53.8%	-51.2%
Additional Mechanical Items	-3.1%	-3.6%	-3.6%	-3.1%
Total Equipment	-24.2%	-21.3%	-18.5%	-15.8%
Total Equipment+Bauxite	-22.7%	-20.2%	-17.6%	-15.1%

A sensitivity study was also made on the impact of the cost of the heat transfer media. The sintered bauxite used for the MTF test campaigns is used by industry in large quantities for enhanced oil recovery, but is still very expensive. This material would cost about \$0.27/lb (in 2005\$) when ordered in the bulk quantities needed for the commercial CMB boiler. Alternative heat transfer medias were also evaluated (see Section 8.2) that would have significantly lower costs. One alternate material could be obtained in bulk quantities at prices as low as \$0.05/lb. Figure 11-13 compares the CMB material cost savings relative to the CFB design at two different combustor outlet temperatures and with two different heat transfer media. There is almost a 9% variation in cost savings depending upon design conditions. However, reducing the price of the heat transfer media by over 80% still only improved the total boiler island material savings (as compared to the CFB) by an additional 0.7% - 1.5%. Even with the reduced cost of the heat transfer media, none of the CMB designs met the 30% materials cost savings target against the CFB design.

Figure 11-13: Design Impact on CMB Boiler Material Cost Savings Relative to CFB

11.4. 150MW_e CMB Commercial Plant Economic Evaluation

11.4.1. Economic Criteria and Design Criteria

Economic Criteria

The economic evaluation criteria used in this study is shown in Table 11-10. The study was carried out over a 15-year evaluation period and assumed a 7% cost of money and 3% escalation rate. The criteria included values for limestone, lime, scrubber ash sale price, and bauxite price. The evaluation also assumed that 25% of the bauxite is replaced each year.

Table 11-10: 2x150Mw_e Study Economic Evaluation Criteria

Coal	\$1.00	\$/mm Btu
Limestone	\$20.00	\$/ton
Limestone for Wet Scrubber	\$30.00	\$/ton
Lime	\$65.00	\$/ton
Ash Disposal	\$20.00	\$/ton
Scrubber Ash Disposal/Revenue	\$10.00	\$/ton
Power	\$0.02	\$/kw
Steam	\$0.001	\$/lb
No. Boilers	2	
Capacity Factor	80%	
Operation Time	7,008	Hrs/Yr
Cost of Money	7%	
Escalation Rate	3%	
Erection/Materials Ratio	1.00	
Bauxite Price \$/lb	\$0.27	\$/lb
Bauxite Makeup/yr	25%	%/yr
Number of Years	15	

Design Conditions

Table 11-11 summarizes the design conditions for the 9 primary cases considered in this study, including 2 CFB cases and 7 CMB cases with various SO₂ control technologies. All of the CMB cases were based on the CMB design with the 630°F combustor gas exit temperature. This case had the lowest boiler island capital costs and was considered the most favourable design. However, it did not consider the additional costs necessary for further CO reduction. Additional cases were also evaluated with the CMB designs with the higher combustor gas exit temperatures.

All of the designs in Table 11-11 were based on a high sulfur lignite fuel with a 98% sulfur capture requirement. The table shows that there is a significant impact of in-furnace sulfur capture on boiler efficiency. CMB designs with only 50% in-furnace sulfur capture had boiler efficiencies over 1.6% less than designs with 92% in-furnace sulfur capture. Note that the designs with sorbent feed directly to the dry scrubber/wet scrubber did not have the same boiler efficiency penalty because there is no limestone calcination in the furnace.

Table 11-11: 2x150Mw_e Study Design Conditions

	CFB No NID Baghouse	CFB With NID	CMB NID Hi H2O	CMB Stm Reac NID
Coal Sulfur - Uncontrolled SO ₂ (ppm)	5478	5478	5478	5478
Total Sulfur Capture %	97.5	98	98.1	99.1
Furnace Sulfur Capture %	97.5	92.5	50	92.5
Sorbent	CaCO ₃	CaCO ₃	CaCO ₃	CaCO ₃
Ca/S Ratio	2.75	1.95	1.95	1.95
Backend SO ₂ System	None	FDA	FDA	FDA
Boiler Efficiency	81.23%	82.31%	80.67%	82.31%

	CMB Bagfilter Wet Scrub	CMB Dry Scrub Lime	CMB Cyc NID CaO/Hi H2O	CMB Cyc NID Hi H2O	CMB Stm React Cyc NID
Coal Sulfur - Uncontrolled SO ₂ (ppm)	5478	5478	5478	5478	5478
Total Sulfur Capture %	98	98	98	98	98
Furnace Sulfur Capture %	0	0	0	50	92.5
Sorbent	CaCO ₃	CaO	CaO	CaCO ₃	CaCO ₃
Ca/S Ratio	1.40	1.80	1.80	1.95	1.95
Backend SO ₂ System	Wet Scrubber	Dry Scrubber	CycNID	CycNID	CycNID
Boiler Efficiency	82.14%	82.14%	82.14%	80.67%	82.31%

11.4.2. 150Mw_e Commercial Boiler Capital Costs

Capital costs were developed for a 2x150Mw_e boiler island and gas cleanup systems. The boiler island capital cost includes all boiler island materials, services (freight, insurance, engineering, field services), risk, and G&A. The CFB boiler island erection estimate was based on the total boiler island weight and included factors for direct labor, tools, supervision, G&A, and second plant erection. The CMB erection estimates were then prorated from this estimate based on total boiler island material costs (but not including the weight of the heat transfer media). Applying this methodology, the CMB boiler island D&E cost was \$18.9 million less than the CFB design. This result is consistent with CMB's reduced weight and simplicity of the pressure part design.

The capital costs for the limestone preparation and gas cleanup equipment were estimated in a similar manner. Table 11-12 shows the capital cost difference relative to a CFB with NID system for the complete 2x150Mw_e boiler island and gas cleanup systems for seven different CMB configurations and for a CFB without the NID system. The CFB design with NID™ system was used as the base design in this table. The original CMB design (with NID system/Hi H₂O) for a 2x150Mw_e boiler costs \$17.4 million less than the CFB design with NID system. There were three variants of the CMB design with CycNID, each one costing between \$25.6 million and \$29.0 million less than the CFB design with NID system.

Table 11-12: 2x150Mw_e Boiler Island Capital Cost Difference Relative to a CFB with NID System (\$2005)

	CFB No NID Baghouse	CMB NID Hi H2O	CMB Stm React NID	CMB Bagfilter Wet Scrub
Delta \$	-\$9,425,400	-\$17,360,400	-\$18,650,400	\$25,923,800
Delta %	-6.8%	-12.6%	-13.5%	18.8%
	CMB Dry Scrub Lime	CMB Cyc NID CaO/Hi H2O	CMB Cyc NID Hi H2O	CMB Stm React Cyc NID
Delta \$	-\$22,761,600	-\$28,971,000	-\$25,560,400	-\$26,850,400
Delta %	-16.5%	-21.0%	-18.6%	-19.5%

Table 11-13 summarizes the auxiliary power requirements for a single 150Mw_e boiler for each CFB and CMB case. The power savings was not as great as expected in this study because the 150Mw CFB boiler only used a single small FBHE and thus had relatively small FA power consumption. The CMB also used some fluidizing air to transport the bauxite. Overall, the CMB used about 1.5% - 2% more power than the CFB designs for the equipment included. The percent differential would be smaller if the power for the remaining common equipment was also included. The CMB is still expected to show some power savings relative to larger CFB designs that use multiple FBHEs and have larger FA blower power requirements.

Table 11-13: 1x150Mw_e Boiler Island Auxiliary Power Summary (kilowatts)

	CFB No NID Baghouse	CFB With NID	CMB NID Hi H2O	CMB Stm React NID	CMB Bagfilter Wet Scrub
Total kw	5,890	6,290	6,420	6,420	9,020
Delta %	-6.4%	0.0%	2.1%	2.1%	43.4%
	CMB Dry Scrub Lime	CMB Cyc NID CaO/Hi H2O	CMB Cyc NID Hi H2O	CMB Stm React Cyc NID	
Total kw	6,390	6,390	6,420	6,390	
Delta %	1.6%	1.6%	2.1%	1.6%	

11.4.3. 150Mw_e Commercial Boiler Operating Costs

Table 11-14 summarizes the annual operating costs for a single 150Mw_e boiler for each CFB and CMB case. All of the CMB designs had higher annual operating costs than the base CFB with NID design. Each CMB design had extra costs associated with bauxite makeup and O&M on a condensate polishing system. Some CMB designs also had added costs for steam (steam

reactivation), higher power consumption (wet scrubber), and higher sorbent costs. The CMB designs with the lowest operating costs (only about 1.5% higher than the CFB design) were the two CMB designs with steam reactivation and the CMB design with the wet scrubber.

Table 11-14: 1x150Mw_e Boiler Island Annual Operating Cost Summary (\$2005)

	CFB No NID Baghouse	CFB With NID	CMB NID Hi H2O	CMB Stm Reac NID	CMB Bagfilter Wet Scrub
Total Operating	\$22,309,400	\$19,934,700	\$20,581,900	\$20,264,600	\$20,235,900
Delta %	11.9%	0.0%	3.2%	1.7%	1.5%
	CMB Dry Scrub Lime	CMB Cyc NID CaO/Hi H2O	CMB Cyc NID Hi H2O	CMB Stm React Cyc NID	
Total Operating	\$22,378,700	\$22,378,700	\$20,581,900	\$20,259,400	
Delta %	12.3%	12.3%	3.2%	1.6%	

11.4.4. 150Mw_e Commercial Boiler Total Evaluated Costs

Total evaluated or life cycle costs were developed for a 2x150Mw_e boiler for each CFB and CMB design. The evaluated costs included boiler and gas cleanup system capital costs, and life cycle costs for auxiliary power and all operating costs. The evaluated cost savings are expressed as a delta cost compared to the base CFB with NID design. They do not include any CO mitigation costs, which could add another \$3-5 million cost to the CMB cases with low combustor gas outlet temperatures. A sensitivity study was also completed with parametric variations in fuel sulfur content, CMB furnace outlet temperature (630°F and 1300°F), and heat transfer media cost (\$0.05/lb and \$0.27/lb).

Table 11-15 summarizes the total evaluated costs savings as compared to the CFB with NID system case for the high and moderate sulfur fuel scenarios (5478 and 3000 ppm uncontrolled SO₂). The cases are ordered by most favorable evaluated cost savings for the two fuel sulfur concentrations as compared to the CFB with NID design. All of the cases in this table were based on the CMB design with a 630°F combustor gas outlet temperature and a heat transfer media cost of \$0.27/lb.

Table 11-15: 2x150Mw_e Boiler Island Evaluated Cost Comparison vs CFB (\$2005)

Design Basis	5478 ppm SO ₂ Inlet	3000 ppm SO ₂ Inlet
CMB w Steam Reactivation, CycNID	-\$20,986,184	-\$21,844,445
CMB w CycNID, Hi H ₂ O	-\$13,535,406	-\$20,986,184
CMB w Steam Reactivation, NID	-\$13,530,709	-\$13,530,709
CMB w NID, Hi H ₂ O	-\$5,335,406	-\$10,547,369
CFB w NID	\$0	\$0
CMB w CycNID, CaO, Hi H ₂ O	\$22,163,731	\$5,395,665
CMB w Dry Scrubber, CaO	\$28,373,208	\$11,605,142
CMB w Wet Scrubber	\$30,418,676	\$40,830,496
CFB w/o NID	\$42,262,776	\$24,274,812

The results show that several CMB configurations look promising compared to the CFB with NID design. The CMB with steam reactivation and CycNID showed a \$21 million evaluated cost savings compared to the CFB with NID design for both fuel sulfur concentrations. The CMB with CycNID at high humidity and the CMB with steam reactivation and NID system also showed savings of \$13.5 million with the high sulfur fuel and up to \$21 million savings with the moderate sulfur fuel. The original CMB design with NID system at high humidity showed savings of \$5.3 million and \$10.5 million with the high and moderate sulfur fuels, respectively.

Not surprising, the CMB designs with lime sorbents (CMB with CycNID+CaO and CMB with Dry Scrubber) and the CMB with wet scrubber had much higher evaluated costs than the base CFB design with NID system. The low Ca/S mole ratio with these designs was overwhelmed by the high sorbent price for lime and the high capital and auxiliary power costs for the wet scrubber.

These results show that there are a number of CMB configurations that hold promise for total evaluated cost savings compared to the CFB with NID design. However, the most promising configurations all had environmental control technologies that were also under development and not commercially ready. Furthermore, the potential savings would be offset somewhat by the need for a CO mitigation system because of the low combustor outlet temperature – an additional cost of \$3–5 million.

The need for the CO mitigation system might be avoided by using a CMB design with the higher (1300°F) combustor gas outlet temperature. However, the sensitivity study showed that the total evaluated savings with the 1300°F CMB design relative to the CFB with NID design would decrease by \$8.5M or \$9.2M when the heat transfer media price was \$0.27/lb or \$0.05/lb, respectively – regardless of the environmental control method or fuel sulfur content. Thus, the CMB design with the 630°F combustor gas outlet temperature would be more attractive than the higher temperature CMB design even with the added expense of a CO mitigation system.

11.4.5. Potential CMB Cost Savings

The CMB was originally developed as a cost effective design for high temperature Rankine cycles. The proceeding analysis showed that there is also the potential to adapt the CMB as a low cost CFB alternative, despite the fact that the target of boiler island material costs being at least 30% less expensive than a conventional CFB was not met. Table 11-16 summarizes some

of the potential areas where cost savings can be realized with the CMB design. However, these savings need to be tempered by the need for further development of some of the environmental control methods proposed in Section 8.4 and by the need for a CO mitigation system.

Table 11-16: Potential CMB Cost Savings

Lower Capital Costs

- Reduced Pressure Parts
- No Hi-Temp Cyclone
- Reduced Structural Steel
- Smaller footprint/building
- Standard Pressure Parts
- Shorter Links, No RH Sprays
- No Sootblowers
- No Pulverizers

Lower Operating Costs

- Low Gas Side Pressure Drop
- Low Steam Side Pressure Drop
- Reduced Auxiliary Power
- Low Cost Fuels
- Reduced Fuel Consumption

Lower Environmental Control Costs

- No Pressure Parts in SNCR/SCR Temperature Windows

11.4.6. Recommendations from Cost Study

The CMB cost study showed that the CMB as currently designed would not meet the near term goal of a 30% cost reduction target relative to a conventional CFB even if the high CO emissions could be resolved with no additional cost. There were some scenarios where the total evaluated costs of the CMB design were favourable compared with the CFB with NID design. However, these scenarios all relied upon environmental control technologies that were also still under development. Given the technical risks associated with CMB, these potential savings did not make a compelling argument for continued development of CMB as a low cost CFB replacement.

Further development work to address these technical issues would be costly and outside the scope of the current CMB program. Although CMB's long-range potential was for high temperature Rankine cycles, its near-term use as a lower cost CFB was considered to be a necessary stepping stone in its development path. The total evaluated cost of a near term CMB design was not significantly lower than ALSTOM Power's conventional CFB designs. ALSTOM Power therefore did not see a compelling justification to continue CMB development at this time.

ALSTOM Power plans to continue development of the MBHE technology as a lower cost replacement for a CFB MBHE. The MBHE eliminates the use of high pressure fluidizing air and results in significant auxiliary power savings in addition to capital cost reductions. From the DOE's perspective, the MBHE is an enabling technology for both chemical looping and O₂-fired CFBs. Continued development of this technology will reduce the risk of future demonstrations of both chemical looping and O₂-fired CFBs. ALSTOM Power therefore requested and received the DOE's agreement to redirect the CMB program to continue development of the MBHE only. The focus of this work was therefore directed towards a long-term field test of the MBHE at a commercial CFB plant, which is discussed in Section 12.

12. MBHE COMMERCIALIZATION

The objective of this task was to accelerate MBHE commercialization by installing and then testing a MBHE slipstream in a commercial power plant for extended testing. A customer was first identified for slipstream field-testing of the MBHE. Plant arrangement drawings were then prepared for the field test and the MTF MBHE was modified and shipped to the commercial CFB plant. Extended duration field-testing was then conducted to assess long-term heat transfer performance and MBHE operational reliability.

12.1. Overview

The CMB concept was deferred indefinitely once cost studies (described in Section 11) showed that the CMB near-term goal of a 30% cost reduction target relative to a conventional CFB could not be realized. The decision at that time was to refocus the CMB project towards accelerating MBHE development. As shown in Table 12-1, the MBHE has many attractive characteristics, including the elimination of high-pressure fluidizing medium for heat transfer from ash to steam and reduced capital and operating costs. The MBHE is a viable alternative to external heat exchangers for large size CFB boilers, advanced coal combustion processes (oxy-fired CFBs and chemical looping), CFB retrofit applications, and selected industrial processes.

Table 12-1: Benefits of MBHE Technology

- Reduced Capital Costs
 - Pressure Part Reductions
 - Reduced Refractory
 - Potential Bundling of MBHE and Seal Pot
- Reduced Operating Costs
 - Eliminate Fluidizing Air Flow
 - Reduced Head/Power Control Air
- Eliminates Maintenance Issues Associated With FBHEs
 - No Moving Parts in Solids Flow Stream
 - Reduced Solids Flow Vibration and Clamp Forces
- Application to Non-Standard Processes

Figure 12-1 compares the layout of a commercial CFB boiler with both a MBHE and a FBHE. The MBHE arrangement is much simpler and more compact.

Figure 12-1: CFB Comparison with MBHE vs FBHE

12.2. MBHE Field Testing Objective

Favorable MTF results indicated the need for a long-term demonstration test using commercial CFB ash. The CMB program was therefore modified with the DOE's agreement to find a commercial CFB plant willing to test a MBHE slipstream for an extended duration. The objectives of the slipstream test program were to:

- Demonstrate Long-Term MBHE Reliability and Heat Transfer Performance
The MBHE test period was defined as 90 days of operation
- Demonstrate Ability to Control Ash Flow Rate
- Demonstrate Ability to Operate Without Ash Agglomeration nor Blockage
- Develop Scale-up Design Criteria

The plan was to use the existing MTF MBHE for the field test. However, it was decided to reconfigure the MBHE from a 32" x 32" plan area to a 16" by 84" slice plan area in order to better evaluate the effect of solids distribution on MBHE heat transfer performance. This would increase the width of the MBHE to almost half the width of a commercial 250Mw MBHE. This larger width MBHE would enable the solids distribution through the MBHE to be evaluated and would provide valuable information for developing a MBHE scale-up design criteria.

12.3. Grant Town CFB Description

Five different CFB plants were visited in an attempt to find a suitable host site to test the revised MBHE. The first four CFB sites had various arrangement issues that interfered with the proposed MBHE installation. The final CFB site was a 2x40Mw_e CFB in Grant Town, West Virginia that was owned by American Bituminous Power Partners (AMBIT). This plant burned a bituminous coal waste and began commercial operation in 1993. The plant had sufficient room

for the MBHE installation, plus the CFB was small enough that the MBHE could potentially provide some process benefits to the plant.

The Grant Town CFB tends to operate with excessive combustor temperature – often up to 1650°F. As a result, NO_x emissions are high and excessive limestone consumption is required for SO₂ control. The cyclone refractory also requires frequent maintenance due to the high temperatures. The MBHE slipstream will cool recirculating solids from the cyclone sealpot and potentially reduce the combustor temperature by up to 50°F at full solids flow. The lower temperature could improve combustor performance and lead to savings in ammonia and limestone consumption. The MBHE would also give the plant the ability to trim combustor temperature to meet fuel property variations. The plant management was therefore willing to host the MBHE slipstream testing and the AMBIT Grant Town CFB plant was selected for the demonstration. Figure 12-2 shows an aerial view of the Grant Town facility.

Figure 12-2: AMBIT Grant Town CFB Facility

12.4. MBHE Slipstream Redesign and Fabrication

The MTF MBHE was reconfigured into a slice arrangement to better represent a commercial MBHE. The tubes were rearranged end-to-end to provide a slice model with a 7' length, which would be ½ of the width of a 250 MW_e MBHE. The MBHE consisted of two parallel pairs of four tube bundles each. Each bundle was five tubes wide. The bundles consisted of 2.0" OD tubes with 0.5" or 0.625" high by 0.12" thick circumferential fins on 0.5" spacing. The first two tube bundles were three tubes deep in the vertical orientation. The third and fourth tube bundles were four and two tubes deep respectively. The fin tip to fin tip clearance between adjacent tubes was 1" in order to permit rubble of that size to pass through the bundle. The tubes were arranged in a staggered array so that ash passing vertically down could not bypass any region without contacting a finned surface.

Figure 12-3 shows a cross-section of the Grant Town MBHE slipstream design. It shows the tube arrangement, the solids distributor below the tubes, and the ash control valves. Figure 12-4 compares the arrangements of the MTF MBHE and the Grant Town Slipstream MBHE. The

MTF tube bundles were laid out side by side in a 32" x 32" arrangement. The Grant Town tube bundles were laid out end to end in a 16" x 84" arrangement.¹⁹

Figure 12-3: Grant Town MBHE Slipstream Configuration

¹⁹ *MBHE Grant Town Commercial CFB Test*, Bard Teigen, ALSTOM Power Internal Report PPL-05-CT-10, April 2010

**PPL CMB MBHE Installation
Bundles Side by Side 32"x 32"**

**Grant Town Slip Stream Installation
Bundles End to End 16" x 84"**

Figure 12-4: Comparison of MTF and Grant Town MBHE Arrangements

The MTF MBHE was reconfigured in ALSTOM Power's Power Plant Laboratory before shipping to the Grant Town CFB site. The tube bundles were removed from the MTF MBHE and installed into a new MBHE refractory-lined casing. The left picture in Figure 12-5 shows the front side of the MBHE after the tube bundles were installed but before the front wall was put in place. The right photo shows the instrumentation mounted on the backside of the MBHE.

MBHE Front Showing Tube Bundle

MBHE Back Showing Instrumentation

Figure 12-5: MBHE Slipstream Modifications

The four photos in Figure 12-6 shows the MBHE in preparation for shipping to the Grant Town site. The upper left and right photos show the MBHE before and after installing the front wall. The bottom left photo shows the MBHE being filled with sand. The sand was added to protect the tube bundles during shipping. The final photo shows the MBHE loaded on the truck and ready for shipping to Grant Town.

MBHE Before Installing Front Wall

MBHE With Front Wall

MBHE Being Filled With Sand
Before Shipping

MBHE Ready For Shipment to Grant Town

Figure 12-6: MBHE Slipstream Fabrication and Shipping

12.5. Grant Town MBHE Installation

The fabrication of the MBHE slipstream was not completed in time for installation during the Grant Town CFB's September 2007 annual outage. The plan was to install the MBHE in a slipstream of ash from the south cyclone loop seal of the AMBIT Grant Town A-boiler. The main tie-ins were therefore installed during the outage. This allowed the MBHE to be installed at a later date while the plant was in operation. A few minor connections were not completed before the plant restarted operation, but these were made while the plant was operating.

The MBHE was finally installed in August 2008. The Grant Town MBHE slipstream installation took recirculated ash from the cyclone seal pot. The cooled ash leaving the MBHE was pneumatically returned to the combustor overbed air port. Condensate return water was used for cooling, with the heated water returned to the plant deaerator inlet so that no heat was lost

from the system. Figure 12-8 shows a schematic of the MBHE installation at Grant Town. Figure 12-8 shows a closer view of the MBHE installation layout.

Figure 12-7: Grant Town MBHE Slipstream Installation

Figure 12-8: Grant Town MBHE Slipstream Installation

After installation, initial shakedown tests indicated the need for revisions in order to achieve the desired MBHE operating conditions. Unfortunately, plant schedules delayed the revisions until September 2009. Operation at higher ash flows and heat flows started in September 2009 and continued depending on boiler availability. The balance of this section describes some of the equipment and modifications that were made prior to extended MBHE operation.

Ash was supplied through a 24"OD x 8" ID refractory lined ash inlet pipe to the top of the MBHE from the Grant Town A-boiler south cyclone loop seal. Two 24" knifegate valves were installed in the pipe for isolation of the MBHE from the boiler if needed. The first was a manually operated valve. The second was a pneumatically operated valve that could be operated by the plant operator from the DCS control panel for emergency isolation from the boiler if needed. The expansion joint in the inlet pipe shown in Figure 12-9 allowed for differential expansion between the boiler and the stationery MBHE. The perforated rubble screen with 0.75" openings shown in Figure 12-10 was installed in the cyclone loop seal at the inlet of the ash pipe to prevent rubble and agglomerates of larger size from entering the MBHE. The perforated screen was 3/8" thick and was made of alloy 253.

Figure 12-9: MBHE Ash Inlet Expansion Joint

Figure 12-10: MBHE Ash Inlet Pipe Screen

Cooling water for the tube bundles was supplied by the plant condensate booster pump and returned to the #2 feedwater heater discharge prior to the dearator water inlet in order to recover heat from the cooled ash. An orifice meter at the inlet of each bundle measured cooling water flow rates. The inlet and outlet temperature was measured by Type K thermocouples. The water made 2-3 horizontal passes and 2-4 vertical passes, then exited the heat exchanger, and passed by an outlet water thermocouple and vertically upward to a collecting header and back to the plant feedwater system.

Successful operation of the MBHE required a uniform solids flow distribution through the tube bundle. A flow distribution baffle shown in Figure 12-11 was installed at the bottom of the MBHE for this reason. The distribution baffle consisted of a series of orifice plates, the top having sixteen openings, the second having four openings, and the bottom having two seal legs leading to the ash control valves (ACV). A series of tests were conducted with orifices installed in the top distributor plate to reduce the opening diameter. Figure 12-12 shows the ash distributors installed below the tube bundles.

Figure 12-11: MBHE Ash Distribution Grid

Figure 12-12: MBHE Ash Distribution Grid

Two ash control valves at the bottom of the MBHE controlled the solids flow through the MBHE. Solids discharged from the ACV were pneumatically transported through a 10" Schedule 40 A404 transport pipe to the A-boiler combustor and through an 8" primary over-fire air port. Transport air was supplied by either the 90 inwg primary air fan or by the plant 10 psi high pressure ash transport spare fluidizing blower. Figure 12-13 shows the arrangement of the MBHE ash return back to the combustor.

Figure 12-13: MBHE Ash Return to Combustor

Figure 12-14 shows four photos of the Grant Town MBHE installation. The upper left photo shows the ash discharge pipe leaving the sealpot and before entering the MBHE and the two solids isolation valves. The upper right photo shows the location of the MBHE between the PA duct and the building sidewall. The lower left photo shows the two ash control valves below the MBHE feeding into the ash transport pipe back towards the combustor. The lower right photo shows the ash transport pipe from the MBHE and just before the combustor. Isolation valves can be seen on either side of the cross.

Ash Discharge Pipe from Sealpot to MBHE

MBHE Located Between PA Duct and Building Sidewall

Ash Control Valves Below MBHE

Ash Transport Pipe from MBHE to Boiler

Figure 12-14: Grant Town MBHE Installation Photos

An instrumentation and control system was installed to control and measure performance of the MBHE. The MBHE was controlled by a PC based ladder control logic feeding an Allan Bradley PLC. All instruments were read by the Alan Bradley PLC data acquisition and controller and a Labview data acquisition system. The Labview system converted the instrument output to engineering units, displayed the results on the PC, and then recorded the results on the PC hard drive for later analysis. Recorded results were transmitted hourly by Internet to ALSTOM Power Plant Laboratory for data analysis.

Instruments were installed on the MBHE to measure its performance. The primary instruments were type K thermocouples to measure solids inlet and outlet temperatures and to measure the tube bundles' coolant inlet and outlet temperatures. In addition, six orifice flow meters were installed to measure cooling water flow rates to each tube bundle. The instrumentation list is summarized in the Appendix in Section 15.1.

The MBHE had eight individual tube bundles, four on each side, with each bundle having independent cooling manual water flow control valve and flow meter. Each tube bundle was fed water from an orifice flow meter and inlet thermocouple. Each outlet section had a thermocouple to measure the water discharge temperature. Heat transfer from the ash to the tube bundles was measured by the water flow through each tube bundle and the inlet-outlet temperature difference of the water.

Ash flow through the MBHE was determined by the heat flow in the tube bundles and by the ash inlet-outlet temperature difference. Two ash inlet outlet temperature differences were used. The first was the difference between the A-boiler south cyclone dipleg temperature and the corresponding ACV seal leg temperature. The second was the difference between the average ash inlet temperature inside the MBHE and the average discharge temperature below the bottom tube bundles. The first method was used for subsequent data analysis. Both a left and right ash flow was calculated to determine ash flow splits between the right and left side of the heat exchanger.

In order to determine the solids flow distribution through the MBHE ash distributor grid, a thermocouple grid was installed below the tube bundles but above the distribution grid. Six thermocouples were placed in three positions on each side with two thermocouples in each pair at the 1/3 penetration from the front to back wall. Deviations from the average temperature by individual thermocouples indicated that a higher or lower than average solids flow was passing by the particular thermocouple. Each pair of thermocouples was located approximately at the mid point between the top distributor grid openings.

12.6. Grant Town MBHE Slipstream Operation

A comprehensive MBHE Development Log Key is shown in the Appendix in Section 15.2 This table documents all of the Grant Town MBHE events and activities, starting from the initial Grant Town site visit in March 2007 through the suspension of MBHE testing in May 2010. The following section summarizes some of the experiences and events that took place during the MBHE operation.

Cooling Water Modifications

The revised MBHE slipstream was installed in the Grant Town CFB boiler facility in September 2008. Several problems encountered during shakedown delayed initial operation of the MBHE. The tube bundle cooling water was supplied from the plant condensate booster pumps and was initially returned directly to the plant dearator. Introduction of the relatively cold cooling water into the dearator caused severe water hammer. The water hammer was caused by steam bubble collapse resulting from the 120°F cooling water contacting the 30psi steam. This problem was solved by rerouting the cooling water return to the discharge of the #2 feedwater heater prior to the dearator. Cooling water was now mixed with the feedwater, eliminating the water hammer. The heat recovery was not affected by this change. Both approaches reduced steam requirements of the dearator, leaving more steam to pass through the low-pressure stages of the turbine. Figure 12-15 shows the location of the MBHE and the cooling water discharge from the MBHE.

Figure 12-15: MBHE Cooling Water Discharge

Isolation Valve Modifications

A further problem delaying startup was related to the two 24" slide plate isolation valves that were installed in the ash inlet pipe below the sealpot to isolate the boiler from the MBHE. The first valve was installed during the September 2007 plant outage. The second valve was installed in the summer of 2008 along with the MBHE while the boiler was in operation. Figure 12-16 shows the two isolation valves below the sealpot. An attempt to open the first valve was made in December 2008. The valve was locked closed by hydrated fly ash lodged in the cavity of the slide plate. The ash may have gotten wet from moisture evolving off the refractory-lined inlet pipe.

Several attempts to open the valve were in vain, including the use of a jack to apply 40 tons of force on the plate. In February of 2008, the first slide-plate was removed and replaced with a spool piece. Subsequent operation of the remaining 24" pneumatically operated valve failed because of a lodged slide plate due to ash carried into the plate pocket. The valve was modified by installing a cleanout port. Prior to opening the valve, the port was opened, cleaned out, and purge air was applied to the valve to clear the plate pocket. This revision enabled the valve to be opened at will. Note that it is not expected that these large isolation valves would be needed in full-scale commercial practice. They were installed at Grant Town to allow the MBHE to be isolated from the plant. This would permit it to be worked on while the plant was in operation and to ensure that MBHE startup problems would not impact the plant's availability.

Figure 12-16: MBHE Solids Isolation Valves

Ash Flow Issues

Initial ash flow to the MBHE was achieved in March 2009. A series of shakedown tests were conducted. The MBHE operated as expected, except that the inlet ash flow was not sufficient to achieve the high loads required. The initial inlet height of the ash takeoff point from the cyclone loop seal was dictated by structural steel interference and was located in the splash zone of the sealpot. This was not the ideal location as the loop seal did not supply sufficient ash to the take off point. Also, the ash screen at the takeoff point was not properly installed resulting in insufficient openings for the ash flow. A series of tests were conducted with the limited ash flow to confirm low load performance. During these tests, several additional problems were encountered.

Pluggage of the MBHE ash discharge distributor and ACV occasionally occurred due to the presence of hydrated ash agglomerates during the low load operation. Three 4" access ports shown in Figure 12-17 were installed in the bottom of the MBHE to provide access to clean out the MBHE if pluggage occurred. A fourth port over the tube bundle was also considered, but not installed because of limited deck access to the proposed area. The cleanout procedure was to isolate the MBHE from the A-boiler, open the clean out ports, and vacuum the ash from the interior of the MBHE.

Figure 12-17: MBHE Inspection and Cleanout Ports

The ash transport pipe from the MBHE to the A-boiler combustor inlet plugged on several occasions. A control logic trip automatically stopped the ash flow whenever the transport pipe inlet pressure exceeded a set point. However, the boiler operating pressure and resulting PA fan pressure drifted ± 15 inwg. This drift required a limited trip set point that restricted ash flow. Setting the trip for normal conditions caused the transport pipe to plug before the trip point during extreme pressure swings. Transport pipe cleanout was often delayed due to lack of plant personnel for the cleanout. The control logic was therefore changed to set the trip on the pressure difference between the ash inlet pipe and the combustor average bed pressure. This eliminated many of the trips.

Preliminary tests with the ash return pipe to the combustor indicated that the transport pressure drop at extrapolated high ash flows would exceed the primary air fan head. A parallel connection to an auxiliary blower having a head of 10 psi was installed during the Fall 2009 outage. The high transport pressure drop was because the existing plant layout required a difficult transport pipe layout to return the solids back to the combustor, including bends, reducers, and vertical pipe sections. Ideally, a commercial MBHE design would use only horizontal or slightly downward angled return piping. The pressure drop through the horizontal portion of the transport line was very low and supported the design philosophy of using only PA Fan air for the transport air.

The solenoid valves controlling pulse air to the ACV failed. The first failure was an air leakage in the solenoid that stopped operation. Subsequent failures caused the solenoid to stay open. This led to poor ash control in the first case and to transport pipe plugs in the second case. The solenoids were replaced with high cycle duty units.

A decision was made to relocate and revise the MBHE ash inlet pipe. The ash inlet was originally located in the splash zone of the cyclone loop seal discharge chamber. The elevation selected was governed by structural interferences outside the loop seal. Review of the interferences showed that the elevation could be dropped by an additional 18". This modification shown in Figure 12-18 was made in the Fall 2009 plant outage. In addition, the ash inlet screen was revised to utilize the entire set of openings. Finally, the slide seal expansion joint on the ash inlet pipe was replaced with an offset metal bellows expansion joint. This eliminated a problematic ash leakage at the top of the MBHE.

Figure 12-18: MBHE Ash Inlet Pipe Modification

Agglomerates

During the installation of the ash inlet pipe, agglomerates at the top of the tube bundle were broken up and air lanced to the bottom of the MBHE. The agglomerates were then removed from the bottom of the MBHE. The ACV had ports cut in strategic locations for inspection and cleanout. Figure 12-19 shows a SEM analysis of one of the agglomerates. The analysis showed that there was no indication of any ash softening in the sample or any evidence of recarbonation. The presence of high Ca concentration in the ash, long periods of MBHE inactivity, and high moisture flue gas support the premise that the agglomerates were a result of calcium oxide or calcium sulfate hydration. A typical agglomerate analysis was loaded with free

lime, with 78%, 16%, and 4% of the calcium distributed as CaO , CaSO_4 , and CaCO_3 , respectively. The agglomerates tended to be friable and easily broken up.

Figure 12-19: SEM Analysis of MBHE Ash Agglomerates

There were a number of unique situations that made the Grant Town MBHE susceptible to moisture condensation and then ash hydration. The combustor operated with relatively high limestone feed rates because the combustor tends to run hot (up to 1650°F). In addition, there were many instances when the MBHE was not operating due to either plant-related problems or

MBHE issues where the cooling water through the MBHE tubes was still on. With cooling water at 120°F and plenty of moisture in the flue gas, there was the potential for moisture condensation on the tubes. This would not be an issue for commercial MBHEs because the tubes would have high temperature steam instead of cooling water passing through them. The MBHE operating procedures were modified to minimize the impact of condensation. This included isolating the MBHE for two days after a CFB startup (because a fresh limestone startup bed would lead to high CaO concentrations in the MBHE) and to turn off the cooling water whenever the MBHE was not in service.

Operation at Higher Solids Flow Rates

Tests with the revised ash inlet and other Fall 2009 revisions had significantly higher ash flowrates. A series of tests were conducted to evaluate MBHE performance at the higher flow conditions. During these tests, the right side ash control valve started to perform erratically and then stopped completely. The tests were continued with ash flow through the left side only. Solids flows up to 18,000 lb/hr or 90% of design solids flow rate were achieved through the left side for about 1 ½ hours before flow had to be stopped because of plant schedules.

Operation at the high ash loads was interrupted by ash leaks at an ash return pipe reducer from 10-inch to 8-inch pipe diameter at the combustor inlet. This was initially repaired with a refractory patch. Concern about the mechanical integrity of the reducer/refractory patch resulted in a shut down of the MBHE until the reducer could be replaced with an impact reducer design.

A review of the high ash flow tests indicated that the ratio of measured to predicted heat transfer was still low. A decision was therefore made to shut down the MBHE, remove the access door, inspect the MBHE interior, and mechanically clean the entire tube bundle. The inspection showed an accumulation of solids on top of the tube bundle, but the bundle itself was free of any solids build-up and did not need much cleaning.

While the MBHE access door was off, a review of the layout of the tube bundles shown in Figure 12-20 showed that the return bends occupied 30% of the plan area. The return bends had no fins so that heat transfer surface was greatly reduced. Ash bypassing through this region would not be cooled sufficiently. A series of baffles were therefore installed in the bend region as shown in Figure 12-21 to reduce the bypassing of ash from the finned region of the tube bundle.²⁰

²⁰ *Extended Surface Heat Transfer*, D. Kern and A. Kraus, McGraw-Hill Book Co. 1972, pg 179

Figure 12-20: MBHE Tube Bundle Ash Bypass Region

Figure 12-21: MBHE Tube Bundle Baffles

An attempt to restart the MBHE after the front door was installed was unsuccessful because of ash agglomeration in the ash inlet pipe. The MBHE ash inlet valve was closed during the period up to and during the door removal. Hot moist flue-gas permeating through the ash ahead of the valve formed agglomerates that prevented ash from flowing into the MBHE after the valve was opened. Removing a spool in the ash inlet pipe during an A-boiler forced shutdown solved this problem. Removing the spool permitted the ash inlet pipe to be cleaned up to the ash inlet screen without entering the cyclone loop seal.

Cyclone Sealpot Return Chute Failure

Ash flow to the MBHE was resumed after a boiler outage in May 2010 and continued for another 14 days until May 15th. This brought the total accumulated MBHE operating time to 121 days. The MBHE operated with solids flows mostly between 10 -16 klb/hr during this time. Operation was somewhat erratic, with periodic solids flow stoppages on both the left and right sides. The problem was related to the ash control valve (ACV) actuator. At times the actuator did not open the ACV (which supplies fluidizing air), preventing ash flow from that side. Installation of an air lubricator on the control air supply to the actuators seemed to reduce the instances of flow stoppage.

During the previous Boiler A tube leak outage on April 22-24, an ash leak was noticed from the rear wall wallbox. Grant Town Plant Operations made a temporary patch with Blu-Ram refractory, but the leak was not fully investigated. Another Boiler A tube leak outage occurred on

May 15th, and the rear wall ash leak became more noticeable. Grant Town Operations personnel stripped lagging and insulation from the bottom of the rear wall wallbox and found that the bottom panel of the wallbox had pulled away from the lower rear wall tube header. Figure 12-22 shows that the bottom plate had pulled away from the header edge bar at the interface seal weld attaching the wallbox bottom plate to the lower rear wall tube header.

Figure 12-22 is the Boiler A rear wall south wallbox, where the south loop seal ash return enters the combustor. The Boiler A south loop seal discharge is also the location of the slipstream takeoff connection for the MBHE ash inlet line. The face of the wallbox was not distorted, but the bottom plate showed significant warpage. This was an indication that the carbon steel bottom plate was severely overheated. If hot ash had leaked into the wallbox from the combustor, then the castable refractory inside the wallbox would have crumbled from overheating. The photo in Figure 12-23 shows crumbled refractory inside the wallbox. Plant Operations replaced the bottom plate during the May 15th forced outage.

Figure 12-22: Wallbox Bottom Panel Separation from Lower Rear Wall Tube Header

Figure 12-23: Condition of Refractory in Wallbox

Grant Town Plant Operations decided to isolate the MBHE to take the load of the MBHE ash supply line off of the loop seal by unbolting the downstream side of the MBHE inlet isolation valve. Upon inspection by Plant Operations, it appeared that the expansion joint on the MBHE box inlet was free and able to expand, although it appeared to be fully compressed with Boiler A in service and with the MBHE out of service. The spring hangers on the MBHE ash supply line were unlocked and in the correct position for the load they were supporting. There was no evidence that the MBHE ash supply line was not able to move thermally with the boiler and loop seal.

Plant Operations decided to leave the MBHE ash supply line disconnected from the loop seal while the situation was investigated further. ALSTOM Power asked Plant Operations to make a few observations to help diagnose the problem. The plant agreed to:

- Measure the distance between a platform and a reference point on the front and rear of the south loop seal under both hot and cold conditions to confirm uniform expansion and contraction
- Look for "hot spots" on the rear wall south wallbox bottom plate that was replaced during the Boiler A outage on May 15.

If the newly replaced bottom plate again overheats and warps, this may indicate there were cracks in the hard face refractory inside the loop seal ash return duct or boiler opening, allowing hot bed and recycle ash to leak into the wallbox. In that case, thermal stress from overheating and ash jacking may be possible causes of wallbox bottom plate separation from the lower rear

wall tube header. However, it may take a few thermal cycles of Boiler A start-up and shut-down for the damage to occur again now that the bottom plate has been replaced.

Figure 12-24 shows the bottom of Grant Town Boiler A and the south loop seal, with the location of the MBHE ash supply line and the MBHE box. The MBHE ash supply line and valving is supported by spring hangers (orange oval in figure), and the MBHE itself is bottom supported by structural steel. There is a stainless steel bellows expansion joint in the ash supply line at the inlet to the MBHE body (orange circle in figure).

Figure 12-24: MBHE Ash Supply form Loop Seal and Support Arrangement

The MBHE ash inlet line bellows expansion joint was designed to accommodate the movement of the ash inlet line. The expansion profile of the inlet line was up to 5" movement downward and up to 1" movement laterally when hot. The loop seal expansion profile was 11/16" south and west (1" lateral), and 4 3/16" downward. The type and size of the ash inlet line bellows expansion joint was chosen to allow more expansion than experienced by the loop seal.

If the MBHE ash inlet line expansion joint was close to fully compressed but frozen in place, the inlet line would not be able to move back to the cold position when Boiler A comes off-line. In this case, the inlet line would exert a downward force on the loop seal as the boiler cooled. The resulting reaction force would be at the top of the wallbox, not at the bottom. Based on this

analysis, the MBHE ash inlet line could not be the failure mechanism for the wallbox bottom plate.

Alternatively, if the MBHE ash inlet line expansion joint was fully compressed with Boiler A in service but the MBHE off, then there would be no additional expansion capability as the MBHE is brought into service. However, both the MBHE ash inlet line and box were lined with refractory and could be touched while in service, so there was essentially no expansion when the inlet line and MBHE were filled with hot ash. The only way for the inlet line to exert an upward force on the loop seal was if there was insufficient expansion joint capacity prohibiting the loop seal from moving downward as Boiler A was brought on-line and heated up. The reaction force would be at the bottom of the wallbox. If this were the case, then the weld between the MBHE ash inlet pipe and either the bellows expansion joint or the root weld connection to the loop seal would fail before the weld between the wallbox bottom plate and bottom header.

Based on the available observations, it is unlikely that the MBHE ash inlet line caused the damage to the bottom plate of the rear wall wallbox. A possible cause may be recycle and bed ash leaking into the wallbox through cracks in the hard face, causing overheating of the bottom plate. At this point, the only way to prove this is for the problem to occur again without the MBHE connected to the loop seal, or for a Boiler A outage where Plant Operations can enter the boiler and inspect the hard face refractory inside the boiler at the loop seal return leg inlet to the combustor. Typically there is no time during a forced outage for boiler internal inspection. Additionally, it may take a few thermal cycles of Boiler A start-up and shut-down for the damage to occur again now that the bottom plate has been replaced.

Additional MBHE testing would have been significantly delayed as Grant Town Operations required several thermal cycles of Boiler A start-up and shut-down to convince themselves that the wallbox failure was unrelated to the MBHE. As a result, ALSTOM Power decided to suspend further testing of the MBHE in May 2010 after completing 121 days of cumulative MBHE operation.

12.7. MBHE Data Analysis Procedure

The MBHE test series were conducted with the local plant data acquisition system recording data at a set rate to the local PC hard drive. Data rates could be set at 1sec, 10sec, 1min, and 10min. Most operation used a 1-minute logging rate. Data was displayed on the PC screen with a 1 second refresh rate for observation of performance.

Two programs were available for viewing the data; the Alan Bradley (AB), and the Labview. The AB program contained trip and control functions, and tabular and trend functions. The AB trend functions could be changed to extended periods such that data trends of 1-24 hours could be viewed on the control PC display screen. The Labview trends were more flexible in selecting values to trend, but had only a 20-minute trend window.

An ALSTOM Power TOPS system was used to transmit data from Grant Town to Windsor for post processing. This system transmitted data recorded on a one-minute interval. The data was sent to Windsor hourly. The data could be pulled from the Windsor server in intervals of one or more minutes depending on the object of the post processing. Most of the data was procured with 10-minute intervals.

The test data was pasted into a preliminary review spreadsheet and plotted as shown in Figure 12-25. Once the steady state conditions were determined, average values of each instrument measurement were calculated for a period of 1-8 hours, consisting of 25-50 data points. These average values were then copied into a data analysis spreadsheet. This spreadsheet was prepared for selected time periods and conditions.

The MBHE test results were reviewed in a series of charts of solids temperatures, solids flow rates, heat flow, and coolant temperatures vs. time to determine periods of steady state operation. Graphic charts such as that of Figure 12-25 have a time scale in hours beginning at 00:00:00 of the first day of the month.

Figure 12-25: Grant Town MBHE Test Data Time Trend

A series of worksheets were then used to calculate the predicted heat transfer performance for given input conditions for each tube bundle. The predicted heat transfer performance was dependant on the calculated outlet ash properties. This required an iterative calculation process for each tube bundle with calculated data values from the worksheet.

The predicted performance was calculated for the temperature difference between the ash and cooling water. The ash outlet and cooling water outlet temperature was required for the calculation. These values were iteratively calculated. At the end of the iterative procedure, predicted heat transfer performance was calculated for each tube bundle. A separate worksheet was used to summarize the closeness of the fit between measured and predicted heat pickup by each bundle for each of the six tests.

12.8. Grant Town MBHE Test Results

Analysis of the Grant Town MBHE test results are based on the measured tube bundle waterside heat flows and ash solids temperatures. There was not a simple technique to measure the ash flow. The ash flow was therefore calculated by a heat balance on the ash determined by the heat removed by the cooling water and the inlet-outlet temperature difference of the ash. There was a significant variation of measured ash temperatures across the MBHE for a given time period during the test as shown in Figure 12-26 and Figure 12-27. Heat flow from the ash to the tube bundles was dependent on the log mean temperature difference between the inlet and outlet temperatures as shown in Figure 12-28. The spatial outlet ash temperatures for one of the tests (on the 104th day of operation) is shown in Figure 12-29. The temperatures shown are the six solids outlet thermocouples along the front and rear of the MBHE. The X and Y axis shows the relative position of the temperature measurement in inches from the MBHE wall. Judgement had to be made in how the temperature variation was accounted for.

Figure 12-26: Left Tube Bundle Solids Outlet Temperature vs. Time

Figure 12-27: Right Tube Bundle Solids Outlet Temperature vs. Time

Figure 12-28: Comparison of Measured Temperatures vs. Ash Flow

Figure 12-29: Typical Ash Outlet Temperature Spatial Distribution

Two calculation procedures were used to determine ash flow rates based on the measured tube bundle heat flows and the ash temperatures. The first procedure used the measured inlet ash temperature from the A-boiler cyclone dipleg and the MBHE ACV seal leg temperatures. The second method used an average solids inlet temperature measured by the thermocouples above the MBHE tube bundles and the average solids outlet temperature measured by thermocouples below the tube bundles. These averaged temperatures are shown in Figure 12-30.

Figure 12-30: Averaged Ash Temperatures

The problem with using an average of the six outlet temperatures below each set of bundles is that each temperature assumes an average ash flow past each thermocouple as opposed to possible variations in ash flows. The cyclone dipleg inlet and ACV seal leg temperatures had a more uniform ash flow past the measurement. However, there was only one thermocouple for these last measurements so it was unknown if there were significant variations in temperature at those locations.

The MBHE was designed with an ash distributor below the tube bundles. PPL laboratory test results (described in Section 7.4) demonstrated that the distributor gave relatively uniform ash flow distributions. The Grant Town MBHE did not have such uniform ash flow distributions as evidenced by the solids outlet temperature distributions shown in Figure 12-26 through Figure 12-29. Most probably, this was due to over conservatism in the design of the distributor plate orifices. There was concern about potential rubble and agglomerates accumulation in the bottom of the MBHE when the initial orifice size was selected. As a result, the orifices were enlarged beyond those used in the laboratory to allow any potential oversized material to pass through the orifices.

Smaller orifice restrictions were installed in the ash distributor after 60 cumulative days of MBHE operation. The top distributor plate restriction was reduced from the original size opening. In addition, the restriction extended below the top plate to further restrict the flow by reducing the discharge angle to the next level of ash distributor. This improved the solids flow distribution through the second distributor plate. The effect of the revised orifice restrictors is shown in Figure 12-27 for the MBHE right side. Prior to the orifice change on day 60, the MBHE center

backside and mid-backside thermocouples showed higher temperatures than the other sections of the MBHE. After inserting the restrictors, those backside temperatures decreased to near normal. However, the end backside temperatures increased to above normal temperatures. This indicates that inserts should be installed only at selected locations (vs. at all locations as was done on day 60) to even out the ash distribution.

The orifice restrictions did not have a significant effect on the right side heat pickup and in fact produced a slightly lower heat flow. This effect was probably not due to the orifices but to the accumulation of agglomerates on the top tube bundles that restricted ash flow. The effect of the distributor plate was more pronounced on the lower tube bundles.

The ash temperature distribution at the bottom of the MBHE for the test condition at the cumulative time of 104 days on 04/21/10 is shown in Figure 12-29. The temperature distribution was still not uniform after installing the orifices. This indicates that the reduced size orifices need to be targeted at areas with excessive solids flow rather than reducing all of the orifice sizes. Figure 12-31 shows the recommended locations to install selected orifice inserts in those areas of above average solids flow to reduce the local temperature variation.

Figure 12-31: MBHE Ash Flow Distributor Recommended Orifice Distribution

12.9. 1.9 Grant Town MBHE Test Results Heat Flow

Figure 12-32 and Figure 12-33 show the left and right MBHE tube bundle heat pickup based upon the solids flow measurement using the cyclone inlet and ACV seal solids outlet temperature. Figure 12-34 and Figure 12-35 shows the left and right MBHE heat pickup based upon the solids flow measurement using the MBHE ash inlet and outlet thermocouples. Each figure shows the total measured heat flow, the heat flow for each individual tube bundle, and the total predicted heat flow. The heat flow for the two left tube bundle figures and the two right tube bundle figures was unchanged, but the calculated ash flow was different – in some cases up to 15%. The heat flow was determined from the measured tube bundle water flow rates and the water inlet – outlet thermocouples.

The initial MBHE design calculations indicated that the MBHE should have a total heat capacity of 10 MMBtu/hr with an ash flow of 40,000 lb/hr. The predicted total heat flow for just the left or right side of the MBHE is half of the total prediction or 5 MMBtu/hr. The total predicted heat flow is represented in Figure 12-32 and Figure 12-33 by a dashed red line. The solid red line represents a best fit of the measured MBHE data for the total heat flow. The initial design heat flows were not achieved because ash transport pressure drops from the MBHE to the boiler combustor were higher than anticipated. This limited the total ash flow through the MBHE and the resulting heat flow. The results show that the measured MBHE heat pickup was about 16% less than expectations for the given solids flow rate.

This slight reduction in heat transfer performance was likely due to four factors:

- The ash flow distribution within the tube bundle was non-uniform resulting in poor heat flow. The temperature imbalance through the MBHE was evidence of non-uniform solids flow. It was likely that the solids flow rate through the MBHE during most of the operation was too low for the flow orifices installed in the MBHE distributing plate to have much effect.
- Agglomerative fouling (due to calcium hydration) of the ash at the inlet to the tube bundles probably contributed to the non-uniform solids flow. Inspection of the tube bundles indicated that agglomerates were present in significant amounts on the top of the tube bundles. Several attempts were made to break up the agglomerates and remove them.
- The return bends of the tube bundles presented a 30% plan area that could be bypassed by the solids as shown in Figure 12-20. Predicted performance was based only on the spiral-finned area of the tube bundle.
- A $\frac{3}{4}$ " perforated screen at the ash inlet pipe was installed to prevent boiler rubble and agglomerates from entering the MBHE. The $\frac{3}{4}$ " perforations could pass particles that would accumulate between the spaced fins of the tube bundles.

The Grant Town MBHE results showed that the measured heat pickup increased linearly with ash flow rate. This was attributed to two effects. First, there was a slight influence of ash velocity on the heat transfer coefficient for the tubes and fins based upon previous cold flow and MTF test results (see Figure 10-42). Second, the ash outlet temperature also increased with ash flow, giving an increased log mean temperature difference between the ash and the cooling water.

Figure 12-32: Comparison of Left Bundle Heat Pickup vs. Ash Flow Calculated from Cyclone and ACV Temperatures

Figure 12-33: Comparison of Right Bundle Heat Pickup vs. Ash Flow Calculated from Cyclone and ACV Temperatures

Figure 12-34: Comparison of Left Bundle Heat Pickup vs. Ash Flow Calculated from MBHE Inlet and Outlet Thermocouples

Figure 12-35: Comparison of Right Bundle Heat Pickup vs. Ash Flow Calculated from MBHE Inlet and Outlet Thermocouples

The Grant Town test data was also examined for evidence of any long-term fouling or heat transfer degradation. Best linear fits to the heat pickup data of Figure 12-32 to Figure 12-35 were used to determine heat flow vs. ash flow for these tests. The equations for each of the tube bundles were then used to calculate the measured to calculated heat ratios for each test period. This data was then plotted (for each bundle and for the total MBHE) against the cumulative operating time for the tube bundles as shown in Figure 12-36 to Figure 12-39. Also shown in these figures by a grey square is the ash flow through that bundle. The measured to calculated heat ratio would be expected to decrease with time to a value less than one if there was any fouling

Key MBHE modifications are also superimposed on Figure 12-36 to Figure 12-39 showing the effects of ash distributor orifices on the tube bundles, relocation of the ash inlet pipe from the cyclone loop seal to the top of the MBHE, and the installation of baffles in the return bend regions of the tube bundles. The most significant effect was the installation of the orifices in the distributor. The deviations from the calculated values are less for the orifice than other effects.

Review of the heat transfer performance in these figures show that the normalized heat transfer performance did not significantly change from the start of MBHE operation until the end. Agglomerated ash hydrates were a problem during operation, resulting in plugging of the ash inlet, tube bundle, ash distributor grid, and ACV. These agglomerates were friable and easily removed by draining the MBHE and vacuuming the agglomerates out. However, these figures do not show a decreasing trend with time on the heat pickup ratio compared to the normal scatter of the data. This indicates that there was not any long-term degradation of the MBHE heat transfer.

The large access door was removed from the front of the MBHE in order to fully access the tube bundles and to clear out any embedded agglomerates. Visual inspection by the plant personnel after the door removal did not reveal any significant adherence of agglomerates on the tubes or fins. The tube bundles were thoroughly air lanced to clean the surfaces. At the same time, plate baffles were installed at four elevations in the tube bundle return bend regions to reduce ash bypassing and the resulting outlet temperature variations.

Figure 12-36: Left Tube Bundle Measured/Predicted Heat Transfer vs. Cumulative Operating Time Using Cyclone/ACV Seal Temperatures

Figure 12-37: Right Tube Bundle Measured/Predicted Heat Transfer vs. Cumulative Operating Time Using Cyclone/ACV Seal Temperatures

Figure 12-38: Left Tube Bundle Measured/Predicted Heat Transfer vs. Cumulative Operating Time Using MBHE Inlet/Outlet Thermocouples

Figure 12-39: Right Tube Bundle Measured/Predicted Heat Transfer vs. Cumulative Operating Time Using MBHE Inlet/Outlet Thermocouples

12.10. Ash Return Transport Pressure Drop

Ash return transport pipe pressure drop during the flow of ash from the MBHE to the combustor is shown in Figure 12-40 and Figure 12-41. The transport pressure drop did not change significantly with time until the ash inlet pipe was revised and the ash flow rates were then increased. The impact of ash flow rate on pressure drop is shown in Figure 12-41. As expected, there was a general increase of pressure drop with ash flow. The ash transport velocity ranged from 30 to 50 fps. The ACV pulse air flow rates ranged from 0 to 5 lb/h. The pulse air solids/air ratio was ~1000 at the maximum ash flow. The pulse air was increased significantly for two tests without a significant increase in ash flow.

The pressure drop of the total ash transport system is shown in Figure 12-41 by the red triangles and the red trend line. The total ash transport pressure drop was higher than expected because of the tortuous path the system had to follow towards the combustor, including bends, vertical rises, and reducers. At the design flow rate of 40,000 lb/hr, the overall transport system would have a pressure drop of 55 inwg, which is beyond the capacity of the PA Fan. The PA Fan had an available head of 50 inwg. However, a commercial MBHE would normally only have a horizontal or slightly downward declined transport line. The blue trend line in Figure 12-41 shows the data for the horizontal only portion of the ash transport line. Based on this trend line, a horizontal transport system at full 40,000 lb/hr capacity would only have a pressure drop of 27 inwg, well within the capability of the PA Fan.

Figure 12-40: MBHE Combustor Ash Return Pipe Pressure Drop vs. Time

Figure 12-41: MBHE Combustor Ash Return Pipe Pressure Drop vs. Total Ash Flow

12.11. Summary

The moving bed heat exchanger (MBHE) performance was measured at commercial CFB operating conditions at the AMBIT Grant Town Power Plant. Results were measured over 121 days of cumulative MBHE operation. During the test period, there were several periods where the MBHE was out of operation due either to power plant outages or to MBHE problems.

Results indicate that the heat transfer performance did not deteriorate with time. However, the overall performance was slightly less than predicted. This was most probably due to ash bypassing the finned region of the tube bundles in the return bend region and because of agglomeration build-up in the tube bundle during the initial start-up. Analysis of the agglomerates indicated that they were caused by excessive moisture from the CFB flue gas reacting with the calcium oxide in the ash. In addition to these results, there were a number of lessons learned about MBHE operation in a commercial power plant.

The results also demonstrated that maldistribution of the solids through the tube bundle had a significant influence on performance. The results demonstrated that the solids flow distribution baffle could be changed to improve the distribution at fixed flows but was susceptible to plugging. Pneumatic ash control valves (ACV) at the bottom of the MBHE demonstrated that ash flow could be controlled for an extended period of time.

Heat transfer prediction methods for a multiple tube moving bed heat exchanger were verified to be consistent with operating time and ash flow. Also, the depth effect of the tube bundle does not appear to be significant.

The ash exit temperature profile at the bottom of the MBHE had more variation than expected. A passive ash-distributing grid at the bottom of the MBHE was prone to pluggage from agglomerates formed in the ash inlet pipe and top of the tube bundle. The distributing holes of the grid were enlarged from prior laboratory predictions to permit rubble to pass. It was uncertain whether this was the cause of the excess temperature variation or that accumulations of rubble was the cause. Orifices were installed to reduce the hole size and resulted in an improved distribution on the right side. The distribution on the right side showed that the orifices did change the temperature distribution, and further adjustments may improve the temperature distribution. Unfortunately the left side ACV was plugged during the reduced orifice test. Furthermore, the solids flow through the MBHE may have been too low for the orifices to be effective.

The ash control valves, (ACV), at the bottom of the MBHE operated without significant problems except for three cases. First, an emergency shut down caused the ACV seal leg to shrink during cool down, compacting the ash in the leg and preventing future flow. Second, the solenoid valves controlling the airflow to the ACV failed and were replaced with long-term performance valves. Third the pulsed control air method of operation produced spikes in the transport pipe pressure drop that in some cases plugged the transport pipe. The effect of the spikes was mitigated by changing the pulse sequence.

The total transport pressure drop necessary to return MBHE solids back to the combustor was higher than designed. This was because the existing plant layout required a difficult transport pipe layout to return the solids back to the combustor, including bends, reducers, and vertical pipe sections. Ideally, a commercial MBHE design would use only horizontal or slightly downward angled return piping. The pressure drop through the horizontal portion of the transport line was very low and supported the design philosophy of using only PA Fan air for the transport air.

Recommendations

The heat transfer performance calculation techniques utilized for this test series should be used to calculate performance of future MBHE applications. The MBHE design needs to incorporate baffles or tube bundle arrangements that minimize the ability of ash to bypass the finned portion of the tube bundle.

The use of orifices in the distributing grid at the bottom the MBHE should be applied with an initial start up followed by selective orifice installation based on exit ash temperature profiles.

The ACV provides a means of controlling ash flow and distribution with no moving parts in the ash flow stream. The ACV air actuating solenoid valves need to be selected for long-term high cycle operation. The metal seal legs need to be heat traced to prevent metal contraction during cool down for hot standby conditions.

13. CONCLUSIONS AND RECOMMENDATIONS

13.1. Conclusions

The objective of the CMB Proof of Concept project was to identify the technical, design, and performance challenges that need to be met to make a commercial CMB system. The project objectives were initially achieved in Phase I through a series of experiments that were conducted in ALSTOM Power's Multi-use Test Facility. A number of technical challenges were identified that were the basis for continuing this work in Phase II. Overall Phase II project results confirmed high heat transfer rates in the combustor and MBHE, along with a methodology for predicting scale up performance. Combustion performance was very good and bed ash agglomeration was controllable. Emissions were close to state of the art levels and can be further optimized to meet the DOE's Advanced Combustion Systems targets. Mechanical systems were developed for the MBHE and the MBHE was then tested at a commercial CFB site for 121 days. The following section summarizes the progress that was made against the specific project objectives that were set by ALSTOM Power, in concert with the U.S. DOE, for the CMB Proof of Concept project.

- Objective: Improve the understanding of gas-to-solids heat transfer and then apply it to scale-up predictions for commercial sized units
- Achievement: The Phase I modelling work identified issues with grid independence, number of particle trajectories, and scale up methodology. All three issues were satisfactorily addressed in Phase II.

A grid independence study compared two different CFD models with the total number of grids ranging from 310,00 up to 2,800,000 cells. Both CFD models showed essentially the same particle concentration profiles at all grid levels, indicating that the models were 'grid independent'. This was a significant finding; since the larger models took weeks to get results even when using a computer cluster of ten state-of-the art CPUs.

A similar study investigated the number of particle tracks that were required for the CFD simulations. The particle tracks determine the position in the flow field that the particles exchange momentum and energy with the gas flow. The same two CFD models were run with the number of particle tracks varying from 240 to 240,000 tracks. The study again showed that results were essentially the same regardless of the number of particle tracks. In all cases, the model results showed that the particle distribution was fairly well dispersed.

A scale up methodology was developed based upon a simplified set of scaling relationships that were proposed by Professor Leon Glicksman. The simplified technique required matching the Froude number, the solids to gas ratio, the ratio of the gas superficial velocity to particle terminal velocity, the ratio of linear bed dimensions, the dimensionless solid flow rate, the particle sphericity, and the dimensionless particle size distribution. The scaling relationship indicated that the MTF could be operated under scaled conditions to model a small 25Mw commercial CMB combustor. A CFD model was then made of a 25Mw_e CMB combustor and of the MTF operating in scaled mode. The model results were very close and validated the use of a scaled MTF test for gathering CFD validation data.

Cold flow validation tests were also conducted on the MSE Meili Labasys 100 measuring system to check its feasibility for solids concentration and velocity measurements. Tests were conducted over a range of bauxite particle sizes, solid flow rates, and gas superficial

velocities. The results confirmed that the MSE Meili system was a feasible tool for characterizing gas/solids flow during the CMB hot gas scaling test campaign.

A further verification of the CFD modelling technique was made by operating the MTF with an intentionally disrupted solids flow field, then trying to simulate the results in the model. A deflector plate was installed under the MTF solids injector to artificially induce a poor solid distribution. The CFD correctly modelled the overall trends observed in the MTF for the deflected solids case.

- Objective: Develop recommendations for the tubing in the hottest parts of the MBHE

Achievement: A stress model was developed to accurately determine the design limits of spiral-finned tubing, specifically the limitation imposed on fin tip to fin base temperature difference. The analyses demonstrated that the stresses in the tube and fin can be predicted, including cyclic response, but the lack of information about the strength of the fin-to-tube weld hampered quantitative life estimates. A stress test rig was therefore built to experimentally determine the strength of the fin-to-tube weld under realistic loading.

The finned tube thermal cycling test rig automatically moved a test sample with either air or water-cooling into and out of a heated furnace on an automatic basis. The first sample experienced over 2000 cycles without any indication that the thermal cycling impaired heat transfer. Exposure testing on the second sample ended after 3600 cycles with a fin tip to tube temperature differential of 360°F, again with no signs of fin separation from the tube.

The finite element analysis stress model was validated with test results from the thermal cycling test rig. The model temperature predictions closely matched the measured temperatures for both the fin tip and the tube. Based on the validated model, tip-to-base temperature differences of up to 400°F are predicted to be permissible for cyclic operation where long-term creep failure is not a concern.

- Objective: Evaluate alternate mechanical designs and develop designs for critical components to be included in the commercial conceptual design.
- Achievement: Mechanical designs were developed for various aspects of the MBHE design, including ash control valves, a solids distributor, a distribution orifice plate, and a vent system. All of these designs were tested in cold flow models, followed by hot testing in both the MTF pilot plant and in the Grant Town CFB MBHE.

The MBHE ash control valve was developed to control the solids flow rate and also designed to pass oversized material through the valve. The pneumatic ash flow control valve has no moving parts and uses low-pressure control air from the PA fan to activate the valve and to transport the solids to the combustor. The ash flow control valves can transport up to 400,000 lb/hr/ft² of ash with an ash to air ratio of 2000 or greater. The clearances of the valve were designed to pass a rubble piece of the same size as the valve inlet.

An active grid flow distributor was developed to provide uniform solids distribution control and the ability to pass rubble without plugging. The MTF active grid consists of 16 pyramidal hoppers in a 4x4 array with a flow control shoe under each hopper. Solids flow is controlled by actuating the flow control shoe air for 10-15 seconds and then stopping the air flow. The

air flow is altered among the various shoes of the grid in a prescribed pattern so that each shoe in succession is actuated. The total flow is controlled by the dwell time between activation.

A MBHE distribution orifice plant was developed to ensure uniform solids flow throughout the MBHE. The multiple orifice plate design is simple and requires a 35-40° slope with less elevation requirement, resulting in a 75% height reduction as compared to using mass flow hoppers with 70° angles at the bottom of the MBHE.

A vent system was also developed to direct transport air from the active solids distribution grid to the vent at the top of the heat exchanger.

- Objective: Evaluate alternative techniques to enhance process performance
- Achievement: Techniques to enhance process performance were evaluated for agglomeration mitigation, CO emissions control, alternate methods for sulfur capture, along with an evaluation of CMB fuel flexibility.

A review was completed of the causes of rubble accumulation in the MBHE discharge during the Phase I testing, plus an analysis of agglomeration experience in commercial BFB boilers. The analysis showed that a properly designed MBHE could be operated with reliable solids flow and with an absence of solids agglomeration. Most of the early MBHE operation problems were addressed through proper refractory curing and vent design.

The CMB pilot plant tests have shown that a layer of ash can build up on the sintered bauxite used as the solids circulating material. With time, the ash accumulation will impact the size and aerodynamic properties of the bauxite and the particles will need to be withdrawn from the system and replaced. Since the bauxite is quite expensive, an effort was made to identify alternative candidates for the CMB heat transfer media. Eleven alternative candidates were identified that were all commonly available worldwide and relatively inexpensive. A laboratory-screening test was developed to study the thermal stability of the candidate materials in a simulated CMB environment with respect to melting, softening, and tendency to agglomerate. The samples were exposed in an oven at temperatures between 2000°F – 2200°F in the presence of coal ash, lime, and a mix of coal ash components. Although each of the samples showed varying degrees of ash wetting, none of them exhibited behaviour that eliminated them from consideration as alternate heat transfer media. Further evaluation would be needed to identify a preferred alternate material for the heat transfer media.

The original CMB concept was based upon concentrating all of the heat transfer surface in the MBHE, thus eliminating the need for any combustor or backpass surface. This approach would result in the combustion gas being cooled to about 700°F leaving the combustor. MTF pilot plant tests showed that CO emissions increased as the combustor outlet temperature decreased below 1300°F. Chemkin and Fluent models showed that char particles continued to react and produce CO at temperatures as low as 800°F. However, the models also predicted that CO conversion was essentially quenched as temperatures dropped below 1200°F.

Various options were considered for mitigation of CO emissions. CMB designs were developed that significantly reduced the combustor height, while correspondingly increasing

the combustor gas outlet temperature. This approach would reduce CO emissions, but the cost savings from the reduced combustor height and smaller MBHE were overwhelmed by the added cost of a high temperature cyclone, additional refractory, and a backpass. In-furnace and backend additives were also considered. These methods would reduce CO emissions somewhat, but the additive costs were quite high. A backend CO oxidation catalyst was also considered. Only one catalyst vendor was willing to propose a system that would survive in this environment, due to concerns of catalyst erosion and deactivation from sulfur poisoning. The catalyst would be able to make significant reductions in the CO emissions, but the equipment cost was quite high.

The CMB combustor meets SO₂ emission requirements through a combination of in-furnace sulfur capture and sulfur capture in the backend NID™ system. The NID system captures the majority of the SO₂ because the high combustion temperatures limit in-furnace capture. Depending on the coal sulfur content, the limited in-furnace sulfur capture could also have a significant negative impact on the CMB boiler efficiency and overall economics. Sorbent consumption was minimized during the MTF pilot plant tests by operating the NID system with higher humidity levels than normal commercial practice. A decision was made for the CMB commercial evaluation to limit NID operation to commercially demonstrated humidity levels despite the positive MTF results. This approach would require significantly higher sorbent feed rates. Alternative sulfur capture methods were therefore assessed for the commercial plant evaluation.

A wide range of alternate approaches to sulfur capture were considered, including both commercially ready technologies and some under development. They included wet and dry scrubbers, steam reactivation to enhance in-furnace sulfur capture, and a cyclone based alternative to the NID system. Capital costs were developed for each system. The lowest cost system was the cyclone NID alternative, followed by steam reactivation. The wet scrubber was clearly the most expensive design.

An assessment of fuel flexibility showed that the CMB is capable of handling a wide range of fuels similar to conventional CFBs, although some limitations may result due to the higher temperatures in the lower furnace. The best fuels for CMB are high reactivity, low sulfur fuels, including many subbituminous coal, high volatile bituminous coals, lignites, brown coals, and biomass. These attributes aid carbon burnout, CO emissions, low sulfur capture equipment requirements, and are attractive for ultra high efficiency and potential CO₂ control schemes. Low reactivity, high sulfur content, very high ash, and very low melting ash fuels may require design compromises that increase costs.

- Objective: Conduct MTF pilot plant tests to improve overall CMB performance with several fuels and to gather scaled heat transfer data for CFD validation.
- Achievement: Four MTF test campaigns were conducted during Phase II. The first test campaign used Pittsburgh #8 coal and a range of different limestone sizes. The main objectives were to evaluate MBHE performance with the improved solids distributor, evaluate process improvements to combustion and environmental performance, and to evaluate gas-to-solids heat transfer with poor solids distribution. The second test campaign continued the objectives of the first campaign, but fired a West Virginia bituminous coal and a petroleum coke. The objective of the third campaign was to gather gas-to-solids heat transfer in a scaled MTF test while also gathering validation data to calibrate a CFD model of the CMB process. The objective of the final test campaign was to collect moving bed heat transfer data when the MTF was operating in CFB mode rather than CMB mode.

- Objective: Evaluate gas-to-solids heat transfer in a scaled MTF test and gather heat transfer validation data for the CFD process model. Apply the model to make scaleup predictions of CMB performance in a commercial scale combustor.
- Achievement: Six scaled heat transfer tests were conducted with solids-to-gas ratios ranging from 0.5 to 4. Detailed temperature, solids concentration, and solids velocity profiles were collected for each test. The results showed that the Nusselt Number was generally lower than predicted for single particles due to particle-to-particle interactions. Particles in close proximity had a tendency to follow in each other's wake due to a drag reduction phenomena. Particles following a lead particle were exposed to gas cooled by the lead particle, and therefore did not have the full gas-particle temperature difference available to the lead particle. A multi-particle model developed from earlier warm air tests gave a good fit to the whole range of MTF data.

The CFD model did a reasonable job modelling the MTF gas and solids flows when the heat transfer coefficient was made to be about 0.1 - 0.2 times the default Fluent single particle heat transfer coefficient. The calibrated CFD model was then used to model a commercial scale CMB combustor. The commercial model shows that the solids are flowing downward through the combustor in a gently expanding cone pattern. The gas has a higher velocity in the center of the combustor but has to travel through the falling solids to exit at the top of three cyclone outlet ducts. This flow pattern results in very good heat transfer even with the use of a low heat transfer coefficient. The results are greatly improved as compared to the previous results from the original Phase I commercial CFD models, which showed varying degrees of gas and solids flow separation.

- Objective: Evaluate solids-to-tube heat transfer and solids distribution in the moving bed heat exchanger
- Achievement: Cold flow model heat transfer tests were conducted in a MBHE slice model to determine the effect of tube bundle depth, particle composition, transverse tube spacing, and ash velocity on MBHE heat transfer performance. There was little heat transfer impact with bed depth at solid velocities below 100 ft/hr and only a small decrease in heat transfer at the bottom of the tube bundle at velocities between 100-300 ft/hr.

There was little difference in heat transfer performance between using bauxite particles and CFB ash, despite the wide differences in particle size distribution and chemical composition. The tube spacing tests showed that there was little transverse mixing of the ash as it passed through the MBHE. The tube transverse spacing should be arranged such that ash flowing downward will touch a fin or tube surface – otherwise, the ash will just pass straight through the bundle without touching any heat transfer surface. MBHE heat transfer was measured at ash velocities from 15 to 300 ft/sec and showed a slight increase with velocity.

The MTF MBHE test results showed that the solids-to-tube heat transfer of the tightened pitch tube bundles exceeded the performance of the previous inline tube bundles used in Phase I by up to 60%. It is possible that the tube bundle performance could have been even higher than observed since the tube bundle showed signs of significant solids flow mal-distribution even when the best heat transfer rates were obtained. One promising result was that the tube bundle fins were largely rock-free, with only a few instances of rocks lodged in

the fins. The staggered tube bundle performance could now be predicted for both hot and cold conditions as well as for variations in tube bundle configuration.

- Objective: Evaluate factors to control ash agglomeration in the bubbling bed
- Achievement: Results indicate that bed ash agglomeration is controllable over the range of CMB commercial temperatures. Two bituminous coals and a petroleum coke were fired for 268 hours of operation at 2000°F or above. No agglomerates were formed in the bed and the bed remained well fluidized throughout the operation. These results indicate that the CMB fluidized bed is quite robust and can tolerate temporary temperature excursions or temperature maldistributions.

The bauxite particles did develop an ash coating on its surface that slowly accumulated during operation. The ash coating was primarily a mixture of calcium, aluminum, silica, and iron oxides in varying proportions. The coating rate was retarded by the mechanical abrasion and attrition as the particles passed through the transport system. Although the growth rate was slowing, test duration was not long enough to conclude if the particle ash coating would eventually reach a steady state level, where any further growth was offset by attrition. There was some evidence from prior work that some of the coating would spall off when the coating became thicker.

If not controlled, the ash buildup on the particle surface can have a significant impact on gas-to-solids heat transfer and bed pressure drop. The larger particle size will increase the combustor gas outlet temperature due to the reduced particle surface area at a given solids circulation rate. Furthermore, bed inventory will need to be controlled to maintain a constant bed pressure drop as the ash coating increases. The bauxite particle size will thus need to be managed to some maximum size level to maintain proper boiler performance. Particle size can be controlled through several techniques, including attrition, modifying local bed conditions to reduce the coating growth rate, and solids replenishment. An alternative approach is to replace the bauxite with one of the less expensive heat transfer particles evaluated in this study. These particles will require additional testing in a hot MTF test to confirm that they are suitable alternatives for the heat transfer media.

- Objective: Evaluate CMB process performance, including carbon burnout, sulfur capture in the furnace and in the backend FDA system, and other gaseous emissions, including CO, NO_x, and N₂O
- Achievement: Results indicate good combustion performance and gaseous emissions that could be controlled to current state-of-the-art CFB levels and can be further optimized to meet the DOE Advanced Combustion System targets. Tests were conducted with two different coals, four different sorbents with six different size distributions, and with ammonia and steam injection.

The combustion performance was as good as CFB performance with the same coals. The combustion efficiency for the two bituminous coals ranged from 96.1 to 99.6% over both the Phase I and II MTF test campaigns. The combustion efficiency of the petroleum coke ranged from 97.1 to 97.6%. One test condition was run at reduced load without any ash recycle from the cyclone. The combustion performance during this test was slightly better than the performance at full load with ash recycle. Further work needs to be done to assess this

condition as operation without a cyclone represents a considerable cost savings for a commercial CMB boiler.

NO_x emissions were reduced to below 100 ppm without the addition of ammonia by low excess air and deeper staging. NO_x levels with the Pittsburgh #8 coal were as low as 68 ppm or 0.08 lb NO_x/MBtu without ammonia injection and averaged about 100 ppm. NO_x emissions were somewhat higher with both the West Virginia bituminous coal and pet coke, averaging about 150 ppm for each fuel. NO_x reductions of up to 40% were achieved when aqueous ammonia was injected. The reduction level was affected by the injection location and the local temperature and oxidizing conditions. No ammonia slip was detected during this testing. No further reduction in NO_x was observed with the addition of steam.

N₂O emissions were not measured during Phase II testing, but they were extremely low during all Phase I tests because of the high temperatures in the lower furnace. Emissions were always less than 5 ppm.

CO emissions were generally higher than observed during CFB operation. CO is apparently generated in the upper furnace where the gas temperatures are low. Emissions tended to increase as the temperature in the upper furnace decreased. CO emissions typically ranged from 180 to 400 ppm during these tests. CO emissions tended to increase slightly with the addition of ammonia, while there was little apparent impact from the steam injection.

The sulfur capture results from the Phase I program showed that sulfur capture close to 100% could be achieved with fine limestones and high NID humidity levels. However, the sulfur capture performance in the NID system was limited by how much calcium oxide was carried over to the backend equipment. An objective of the Phase II test campaign was therefore to try to improve the overall sulfur capture performance by increasing the carryover of calcium to the NID system. This was first accomplished by reducing the size of the initial sorbent from a D₅₀ of 50 microns for the EcoCall 50/50 limestone down to a D₅₀ of 9 microns for the Marblewhite 325 limestone. Further attempts to increase the sorbent carryover were also made by intentionally disrupting the cyclone to decrease its collection efficiency.

All of the tests were conducted with a Ca/S mole ratio of 2. The in-furnace sulfur capture with the Pittsburgh #8 coal was roughly 40% to 50% for most of the tests. The overall sulfur capture was 60% to 70% for the early tests and then increased to over 90% when the NID relative humidity was increased to 80%. The sulfur capture with the finer Marblewhite limestone reached 98% with a relative humidity of 80%. The in-furnace sulfur capture with the West Virginia coal was 30% to 40% and averaged 40% with the petroleum coke. The total sulfur capture was about 70% with the West Virginia coal, largely because it used the coarser limestone and calcium oxide wasn't reaching the NID system. The total sulfur capture with the petroleum coke averaged 99% with a NID relative humidity of 80%. These tests were conducted with the finer Marblewhite limestone, which allowed more of the sorbent to pass through the cyclone to the NID system.

- Objective: Develop a commercial CMB boiler concept and cost evaluation
- Achievement: Two CMB boiler designs and cost comparison studies were completed. The first was a preliminary 300Mw_e CMB design, followed by a more detailed 2x150Mw_e CMB design. The preliminary 300Mw_e study included the evaluation of a wide range of CMB arrangements, with the intent of identifying a configuration that would then be used for a

more detailed costs study. Costs were then developed for the 300Mw_e CMB and for a 300Mw_e CFB. The study showed that the CMB boiler material costs were between 19-31% lower than the CFB design. The CMB auxiliary power was about 0.6% less than for the CFB and the O&M costs were about the same for both designs.

The more detailed 2x150Mw_e CMB design study was for steam conditions of 2130psig/1009°F/1009°F. Five CMB designs were considered with combustor gas outlet temperatures varying from 630°F to 1300°F. As the combustor gas outlet temperature increased, the duty in the MBHE decreased and the size of the backpass increased. At the same time, the height of both the MBHE and combustor decreased. Seven different sulfur capture technologies were also evaluated for each CMB configuration. Costs were developed for each CMB design and then compared against a 2x150Mw_e CFB design, both without and with a backend NID system for sulfur control.

The comparison study showed that the CMB combustor had significantly lower weights than the CFB design, with the largest weight savings in pressure parts and also in refractory for the low temperature CMB cases. The CMB boiler design with a 630°F combustor outlet temperature weighed 32.8% less than the CFB design. The weight savings in the CMB with the 1300°F combustor outlet temperature was reduced to 21.1% because of the much larger requirement for refractory.

The cost study showed that the CMB with the 630°F combustor outlet temperature had boiler island material costs that were 22.7% less than the CFB design. The CMB material savings for the 1300°F gas outlet temperature were only 15.1% less than the CFB design, largely because of the large increase in refractory weight. None of the CMB designs evaluated in this study met the 30% materials costs savings target against the CFB design.

Total boiler island costs were developed for both the 2x150Mw_e CMB and CFB designs, including all materials, services, and erection costs for the boiler island and gas cleanup systems. The original CMB design (with NID system operating at high humidity levels) cost \$17.4 million less than the CFB design with NID system. The lowest cost CMB design used a CycNID system for sulfur capture and cost \$29.0 million less than the CFB design with NID system.

Total evaluated or life cycle costs were also developed for the CMB and CFB designs for a moderate and a high sulfur fuel. The evaluated costs included boiler and gas cleanup system capital costs and life cycle costs for auxiliary power and all operating costs. The evaluated costs were expressed as a delta cost compared to the base CFB with NID design. They did not include any CO mitigation costs, which could add another \$3-5 million cost to the CMB cases with low combustor gas outlet temperatures.

The results showed that several CMB configurations looked promising compared to the CFB with NID design. The CMB with steam reactivation and CycNID showed a \$21 million evaluated cost savings compared to the CFB with NID design, while the original CMB design with NID system at high humidity showed savings of \$5.3 to \$10.5 million with the high and moderate sulfur fuels, respectively.

The CMB cost study showed that the CMB as currently designed would not meet the near-term 30% cost reduction target relative to a conventional CFB even if the high CO emissions could be resolved with no additional cost. There were some scenarios where the total evaluated costs of the CMB design were favourable compared with the CFB with NID

design. However, these scenarios all relied upon environmental control technologies that were also still under development. Further development to address both CMB and advanced environmental control technical issues would be costly and outside the scope of this CMB program. Although CMB's long range potential was for high temperature Rankine cycles, it's near-term use as a lower cost CFB for conventional steam cycles was considered to be a necessary stepping stone in its development path. The total evaluated cost of a CMB was not significantly lower than other CFB designs under development at ALSTOM Power that were much closer to commercial readiness. Given the technical risks associated with CMB, these potential savings did not make a compelling argument for continued development of CMB as a low cost CFB replacement.

- Objective: Conduct extended duration MBHE field tests to evaluate long-term heat transfer performance and MBHE operational reliability
- Achievement: ALSTOM Power plans to continue development of the MBHE technology as a lower cost replacement for a CFB FBHE. The MBHE eliminates the use of high pressure fluidizing air and results in significant auxiliary power savings in addition to capital cost reductions. From the DOE's perspective, the MBHE is an enabling technology for both chemical looping and O₂-fired CFBs. Continued development of this technology will reduce the risk of future demonstrations of both chemical looping and O₂-fired CFBs. ALSTOM Power therefore requested and received the DOE's agreement to redirect the CMB program to continue development of the MBHE only. The focus of this work was directed towards a long-term test of the MBHE at a commercial CFB plant.

Five different CFB plants were visited in an attempt to find a suitable host site to test the slipstream MBHE. The first four CFB sites had various arrangement issues that interfered with the proposed MBHE installation. The final CFB site was a 2x40Mw_e CFB in Grant Town, West Virginia that was owned by American Bituminous Power Partners. The plant had sufficient room for the MBHE installation, plus the CFB was small enough that the MBHE could potentially provide some process benefits to the plant.

The MTF MBHE was reconfigured from a 32" x 32" plan area to a 16" by 84" slice plan area to better evaluate the effect of solids distribution on MBHE heat transfer performance. This modification increased the width of the MBHE to almost half of the width of a commercial 250Mw MBHE.

The MBHE installation at the Grant Town CFB was completed in August 2008. Additional modifications were required, but could not be completed until the September 2009 plant outage. Once the modifications were completed, the MBHE performance was measured at commercial CFB operating conditions over 121 days of cumulative MBHE operation. During the test period, there were several periods where the MBHE was out of operation due either to power plant outages or to MBHE problems. The MBHE operation was limited to less than 50% of the solids design flow rate throughout most of the testing due to various mechanical issues. One test was conducted on just the left side of the MBHE with solids flow up to 18,000 lb/hr or 90% of the design solids flow rate.

Operation was finally suspended in May 2010 following a failure of the cyclone sealpot return chute. Grant Town Plant Operations decided to isolate the MBHE to take the load of the MBHE ash supply line off of the loop seal by unbolting the downstream side of the

MBHE inlet isolation valve. Subsequent inspection showed that the MBHE box inlet was free and able to expand. It is highly unlikely that the MBHE contributed to the cyclone sealpot return chute failure as there was no evidence that the MBHE ash supply line was not able to move thermally with the boiler and loop seal. However, Plant Operations still decided to leave the MBHE ash supply line disconnected from the loop seal until they had completed an investigation of the failure. At the time of this report, ALSTOM Power was still waiting for a decision from Grant Town on whether MBHE operation could be resumed.

The MBHE test results indicated that the heat transfer performance did not deteriorate with time. However, the overall performance was slightly less than predicted. This was most probably due to ash bypassing the finned region of the tube bundles in the return bend region and because of agglomeration build-up in the tube bundle during the initial start-up. Analysis of the agglomerates indicated that they were caused by excessive moisture from the CFB flue gas reacting with the calcium oxide in the ash.

The results also demonstrated that maldistribution of the solids through the tube bundle had a significant influence on performance. The results demonstrated that the solids flow distribution baffle could be changed to improve the distribution at fixed flows but was susceptible to plugging. Pneumatic ash control valves (ACV) at the bottom of the MBHE demonstrated that ash flow could be controlled for an extended period of time.

Heat transfer prediction methods for a moving bed heat exchanger were verified to be consistent with operating time and ash flow. Also, the depth effect of the tube bundle did not appear to be significant.

The ash exit temperature profile at the bottom of the MBHE had more variation than expected. A passive ash-distributing grid at the bottom of the MBHE was prone to pluggage from agglomerates formed in the ash inlet pipe and top of the tube bundle. Orifices were installed to reduce the hole size and resulted in an improved distribution on the right side. The distribution on the right side showed that the orifices did change the temperature distribution, and further adjustments may improve the temperature distribution.

The total transport pressure drop necessary to return MBHE solids back to the combustor was higher than designed. This was because the existing plant layout required a difficult transport pipe layout to return the solids back to the combustor, including bends, reducers, and vertical pipe sections. Ideally, a commercial MBHE design would use only horizontal or slightly downward angled return piping. The pressure drop through the horizontal portion of the transport line was very low and supported the design philosophy of using only PA Fan air for the transport air.

The ACV provides a means of controlling ash flow and distribution with no moving parts in the ash flow stream. The ACV air actuating solenoid valves need to be selected for long-term high cycle operation. The metal seal legs need to be heat traced to prevent metal contraction during cool down for hot standby conditions.

13.2. Recommendations

Although CMB's long-range potential was for high temperature Rankine cycles, ALSTOM Power considered CMB's mid-term use as a lower cost CFB for conventional steam cycles to be a

necessary stepping-stone in its development path. Since the total evaluated cost of the CMB was not significantly less than a conventional CFB for conventional steam cycles, ALSTOM Power decided to refocus its CMB development effort to accelerate the commercialization of the MBHE. The following recommendations are therefore made for further MBHE development:

- Previous work on MBHE tube-fin integrity limited the tip-to-base temperature difference to 250°F due to concerns of fatigue failure (e.g. separation of the fin from the tube). Subsequent work showed that tip-to-base temperatures of up to 400°F are permissible for cyclic operation where long-term creep failure is not a concern. Further work is needed to establish a temperature difference bound to ensure that creep-fatigue interaction does not result in premature failure of the fin or the tube.
- Additional MBHE testing should be continued at the Grant Town CFB after the plant evaluation of the cyclone sealpot return chute failure is completed. Testing needs to be completed at full solids flow through the MBHE to confirm performance across the full load range and to also assess the MBHE's impact on the overall CFB combustor performance.
- The heat transfer performance calculation techniques utilized for the Grant Town test series should be used to calculate performance of future MBHE applications. The MBHE design needs to incorporate baffles or tube bundle arrangements that minimize the ability of ash to bypass the finned portion of the tube bundle.
- The use of orifices in the distributing grid at the bottom the MBHE should be applied with an initial start up followed by selective orifice installation based on exit ash temperature profiles.
- The ACV provides a means of controlling ash flow and distribution with no moving parts in the ash flow stream. The ACV air actuating solenoid valves need to be selected for long-term high cycle operation. The metal seal legs need to be heat traced to prevent metal contraction during cool down for hot standby conditions.

14. REFERENCES

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15. ACRONYMS AND SYMBOLS

ACRONYMS

AB	Allen Bradley
ACV	Ash Control Valve
AMBIT	American Biuminous Power Partners
ASTM	American Society for Testing and Materials
BFB	Bubbling Fluidized Bed
Ca/S	Calcium to Sulfur
CFB	Circulating Fluidized Bed
CFD	Computational Fluid Dynamics
CMB	Circulating Moving Bed
CPU	Central Processing Unit
CycNID	Cyclone NID
DCS	Digital Control System
DOE	Department of Energy
DP	Differential Pressure
EDS	Energy Dispersive Spectroscopy
FA	Fluidizing Air
FBC	Fluidized Bed Combustion
FBHE	Fluid Bed Heat Exchanger
ID	Inside Diameter
LDA	Laser Doppler Anemometry
LSHTL	Large Scale Heat Transfer Test Loop
LTRH	Low Temperature Reheater
LTSH	Low Temperature Superheater
LVDT	Linear Voltage Transmitting Device
MBHE	Moving Bed Heat Exchanger
MDC	Metropolitan District Commission
MIT	Massachusetts Institute of Technology
MTF	Multi-Use Test Facility
NETL	National Energy Technology Laboratory
OD	Outside Diameter
PA	Primary Air
PC	Pulverized Coal
P&ID	Piping and Instrumentation Diagram
PPL	Power Plant Laboratories
SA	Secondary Air
SCR	Selective Catalytic Reduction
SEM	Scanning Electron Microscope
SNCR	Selective Non-Catalytic Reduction
S/G	Solids to Gas Ratio
TOPS	Thermal Optimization and Performance System
TVA	Tennessee Valley Authority

SYMBOLS

A_c	Combustor Cross Sectional Area
A_s	Particle Surface Area
C_{Drain}	Carbon Loss in Drain
C_{FBHE}	Carbon Loss in FBHE Drain
C_{Flyash}	Carbon Loss in Flyash
C_{Hopper}	Carbon Loss in Bauxit Separation Hoppers
C_{loss}	Carbon Heat Loss
C_o	Heat Transfer Constant
C_1	Ash Loading Constant
CO	Carbon Monoxide in Flue Gas
C_{p_g}	Specific Heat Of Gas + Ash
C_{p_s}	Specific Heat Of Solids
D_{50}	Mass Median Diameter
D_{max}	Maximum Particle Size
D_p	Particle Diameter
g	Gravitational Constant
G_s	Solids Flow Rate
h	Surface Heat Transfer Coefficient On The Circumferential Fins
h	Gas To Solid Heat Transfer Coefficient
HHV	Coal Higher Heating Value
k	Thermal Conductivity Of Gas
L	Bed Length
LMTD	Log Mean Temperature Difference
LTRH	Low Temperature Reheater
LTSH	Low Temperature Superheater
m_{gas}	Gas Viscosity
$N_{\text{O Tubes}}$	Total Number Of Tube Bundle Tubes
Nu	Nusselt Number
Q	Heat Transfer From Gas To Particles
Q_{gas}	Gas Heat Flow
Pr	Prandtl Number
Re	Reynolds Number
r_{ts}	Solids Cloud Density
S_T	Transverse Pitch
S_L	Longitudinal Pitch
T_{gi}	Gas Inlet Temperature
T_{go}	Gas Outlet Temperature
T_{si}	Solids Inlet Temperature
T_{so}	Solids Outlet Temperature
U_{mf}	Minimum Fluidization Velocity
U_o	Gas Superficial Velocity
U_T	Particle Terminal Velocity
U_{term}	Particle Terminal Velocity
V_{gas}	Gas Superficial Velocity
V_{mig}	Migration Velocity Of Particles
V_{term}	Terminal Velocity Of Particles
Vel_{max}	Maximum Bulk Velocity Of Ash Flowing Past The Tubes
Vol	Volume Of Combustor
W_{ash}	Weight Of Ash Above The Bottom Of The Tube Bundle
W_{Coal}	Coal Feed Rate

W_h	Ash Loading On The Tube Bundle
W_{g+a}	Gas Mass Flow + Recirculated Ash Mass Flow
W_p	Mass Flow Of Particles
W_s	Solids Mass Flow
η_{comb}	Combustion Efficiency
ρ_g	Gas Density
ρ_p	Particle Density
ρ_s	Particle Density
ρ_{ts}	Test Section Solids Density
ϕ	Particle Sphericity
τ	Thermal Time Constant

APPENDIX

15.1. Grant Town MBHE Instrument/Control List

Allen Bradley PLC - Instrument List IP 192.168.0.2

Thermocouples		Labview					
Tag	Description	Units	Index	Instrument			
TE100	Solids Inlet A - left	deg F	0	Type K - 316 SS sheath - 30" length			
TE101	Solids Inlet B - left mid	deg F	1	Type K - 316 SS sheath - 30" length			
TE102	Solids Inlet C - left mid	deg F	2	Type K - 316 SS sheath - 30" length			
TE103	Solids Inlet D - left	deg F	3	Type K - 316 SS sheath - 30" length			
TE104	Solids Outlet A front	deg F	4	Type K - 316 SS sheath - 30" length			
TE105	Solids Outlet A back	deg F	5	Type K - 316 SS sheath - 30" length			
TE106	Solids Outlet B front	deg F	6	Type K - 316 SS sheath - 30" length			
TE107	Solids Outlet B back	deg F	7	Type K - 316 SS sheath - 30" length			
TE108	Solids Outlet C front	deg F	8	Type K - 316 SS sheath - 30" length			
TE109	Solids Outlet C back	deg F	9	Type K - 316 SS sheath - 30" length			
TE110	Solids Outlet D front	deg F	10	Type K - 316 SS sheath - 30" length			
TE111	Solids Outlet D back	deg F	11	Type K - 316 SS sheath - 30" length			
TE112	Solids Outlet E front	deg F	12	Type K - 316 SS sheath - 30" length			
TE113	Solids Outlet E back	deg F	13	Type K - 316 SS sheath - 30" length			
TE114	Solids Outlet F front	deg F	14	Type K - 316 SS sheath - 30" length			
TE115	Solids Outlet F back	deg F	15	Type K - 316 SS sheath - 30" length			
TE116	Cooling water in #1	deg F	16	Type K - 316 SS sheath - 8" length			
TE117	Cooling water in #2	deg F	17	Type K - 316 SS sheath - 8" length			
TE118	Cooling Water Out Bundle 1 Left	deg F	18	Type K - 316 SS sheath - 8" length			
TE119	Cooling Water Out Bundle 1 Right	deg F	19	Type K - 316 SS sheath - 8" length			
TE120	Cooling Water Out Bundle 2 Left	deg F	20	Type K - 316 SS sheath - 8" length			
TE121	Cooling Water Out Bundle 2 Right	deg F	21	Type K - 316 SS sheath - 8" length			
TE122	Cooling Water Out Bundle 3 Left	deg F	22	Type K - 316 SS sheath - 8" length			
TE123	Cooling Water Out Bundle 3 Right	deg F	23	Type K - 316 SS sheath - 8" length			
TE124	Cooling Water Out Bundle 4 Left	deg F	24	Type K - 316 SS sheath - 8" length			
TE125	Cooling Water Out Bundle 4 Right	deg F	25	Type K - 316 SS sheath - 8" length			
TE126	Soilds Outlet Pipe Left	deg F	26	Type K - 22 ga SS braid - spot weld			
TE127	Soilds Transport Pipe Left	deg F	27	Type K - 22 ga SS braid - spot weld			
TE128	Soilds Outlet Pipe Right	deg F	28	Type K - 22 ga SS braid - spot weld			
TE129	Soilds Transport Pipe Right	deg F	29	Type K - 22 ga SS braid - spot weld			
TE130	Solids Transport Venturi Air Temp	deg F	30	Type K - 316 SS sheath - 8" length			
TE131	Pulsed Air Orifice Temp-Tc	deg F	31	Spare			
TE132		deg F	32	Spare			
TE133		deg F	33	Spare			
TE134		deg F	34	Spare			
TE135		deg F	35	Spare			

Analog In		Labview					
Tag	Description	Units	Index	all @ 150 psi and 60 deg F	Pipe	Orifice (in)	Beta
PDT200	Bundle 1 Water Flow Orifice dp-Left	in H2O	36	105.825 in H2O @ 60 gpm	2.067	1.25	0.605 Top
PDT201	Bundle 1 Water Flow Orifice dp-Right	in H2O	37	105.825 in H2O @ 60 gpm	2.067	1.25	0.605 Top
PDT202	Bundle 2 Water Flow Orifice dp-Left	in H2O	38	73.309 in H2O @ 50 gpm	2.067	1.25	0.605 Mid
PDT203	Bundle 2 Water Flow Orifice dp-Right	in H2O	39	73.309 in H2O @ 50 gpm	2.067	1.25	0.605 Mid
PDT204	Bundle 3 & 4 Water Flow Orifice dp-Left	in H2O	40	144.314 in H2O @ 70gpm	2.067	1.25	0.605 Bottom
PDT205	Bundle 3 & 4 Water Flow Orifice dp-Right	in H2O	41	144.314 in H2O @ 70gpm	2.067	1.25	0.605 Bottom
FT206	Pulsed Air Orifice Flow	lb/hr	42	0 - 80 lb/hr - mass pro plate			
PDT207	Pulsed Air Orifice dp	in H2O	43	0 - 21 in H2O - triloop : 4-20ma			
PT208	Pulsed Air Orifice Ps	in H2O	44	0 - 120 psig - triloop : 4-20ma			
TE209	Pulsed Air Orifice Temp-Fixed	deg F	45	70 deg F (fixed) - triloop : 4-20ma			
PDT210	Solids Return Air Venturi dp	in H2O	46	0 - 20 in H2O			
PT211	Solids Return Air Venturi Ps	in H2O	47	0 - 80 in H2O			
PT212	Solids Return Pickup Pressure	in H2O	48	0 - 80 in H2O			
V213			49				
V214			50				
SP215_V	Solids Flow Setpoints		51	from Plant - inc/dec or setpoint			

Allen Bradley PLC - Instrument List
IP 192.168.0.2

Digital Out - Relay Module (110V)			
Tag	Description	Units	Labview Index
SV300	Pulsed Solenoid Open/Close - Left	Digital	52 110V signal - Solids Flow Control
SV301	Pulsed Solenoid Open/Close - Right	Digital	53 110V signal - Solids Flow Control
AL302	Alarm	Digital	54
V303			55
V304			56
V305			57
V306			58
V307			59

Digital In (24V)			
Tag	Description	Units	Labview Index
SV400	Inlet Valve - Open/Close	Digital	60 110V signal from DCS
SV401	Outlet Valve - Open/Close	Digital	61 110V signal from DCS
V402			62
V403			63
V404			64
V405			65
V406			66
V407			67

Calculated Values			
Tag	Description	Units	Labview Index
FT500	Bundle 1 Water Flow - Left	lb/hr	
FT501	Bundle 1 Water Flow - Right	lb/hr	
FT502	Bundle 2 Water Flow - Left	lb/hr	
FT503	Bundle 2 Water Flow - Right	lb/hr	
FT504	Bundle 3 & 4 Water Flow - Left	lb/hr	
FT505	Bundle 3 & 4 Water Flow - Right	lb/hr	
FT506	Pulsed Air Flow - Orifice	lb/hr	
FT507	Solids Return Air Flow - Venturi	lb/hr	
Q600	Bundle 1 Heat Pickup - Left	Btu/hr	
Q601	Bundle 1 Heat Pickup - Right	Btu/hr	
Q602	Bundle 2 Heat Pickup - Left	Btu/hr	
Q603	Bundle 2 Heat Pickup - Right	Btu/hr	
Q604	Bundle 3 Heat Pickup - Left	Btu/hr	
Q605	Bundle 3 Heat Pickup - Right	Btu/hr	
Q606	Bundle 4 Heat Pickup - Left	Btu/hr	
Q607	Bundle 4 Heat Pickup - Right	Btu/hr	
F700	Solids Mass Flow	lb/hr	

15.2. MBHE Development Log Key

03/08/07							Visit to Grant Town for slip stream test		
05/02/07							Visit to Grant Town for slip stream test - Met with contractor - Eddy Shuck		
06/08/07							Meet Contract Draftsman for GT MBHE Installation		
08/17/07							PO issued for Fall GT MBHE Tie-ins		
08/31/07							Grant Town Fall Outage MBHE Tie Ins installed		
11/11/07							Authorization release for GT MBHE Fabrication		
11/28/07							Authorization release for GT MBHE Fabrication		
02/05/08							Signed Host Site Agreement between AMBIT Grant Town and Alstom Power		
02/12/08							MBHE Passes TDGR Gate Review		
04/07/08							MBHE Fabrication complete		
05/27/08							PO issued for GT MBHE Installation		
07/22/08							MBHE Shipped to Grant Town		
07/28/08							MBHE in Position at Grant Town		
09/09/08							Complete Installation Construction		
09/09/08	1400						Set Slide Seal Air Purge, Set ACV Air		
09/10/08	1330						Start Cooling Water fill		
09/10/08	1630						Cooling Water - Water Hammer - Shut down		

MBHE Data Event Key										
Date	Time	LACV dwell	LACV cycle	RACV dwell	RACV cycle	Flow Ratio	Flow Index	Comment	Pulse Integrated Flow	Ash Flow
09/11/08								Slide Seal Expansion Joint Air Leakage Test		
09/12/08								Replaced Mesh Seal with Ceramic Seal		
09/12/08								MBHE-->Combustor Transport Air Test		
09/13/08								ACV Solenoids not closing		
10/28/08								Replaced ACV Solenoids with Ball Valves		
11/17/08								Installed Cyclone-Filter on ACV air supply		
11/19/08								Repiped Cooling Water Return - Still have water hammer		
12/16/08								Pe-piped Cooling Water to #2 FW Heater Discharge - no water hammer		
12/18/08								Ash Inlet Manual Slidegate Stuck Closed - Stripped stem nut - MSV01		
01/15/09								Connected Purge air to MSV01 - No effect		
01/28/09								Hydraulic Jack on MSV01 - 40 tons - No effect		
02/26/09								Removed MSV01 - Replaced with Spool		
03/02/09								Installed purge air to ports on MSV02		
03/05/09								Filled MBHE with cold ash - Ran ash through ACV - Transported to combustor		
03/10/09								Filled MBHE with ash		
03/10/09	1342							Bled & Zero Water DP cells		
03/10/09	1537							MSV02 open -- Start MBHE Ash Flow		
	1856	5	53	5	43	0.104	505305043			
	1921							Hi solids temp trip		
	1930	4	73	5	43	0.078	407305043			
	2004	5	53	5	43	0.104	505305043			
	2030	3	53	4	43	0.073	305304043			
	2214	4	33	4	33	0.121	403304033			
	2237	3	33	3	33	0.091	303303033			
3/11/09	137	3	33	3	33	0.091	303303033			
	440							Adjust ACV Heel and Sole Air flow		
	722							Effect of PA OF ring air flow		
	830	3	30	3	30	0.100	303003030			
	1050	3	27	3	27	0.111	302703027			
	1419	3	60	3	60	0.050	306003060			
	1424							Hi solids temp trip		
	1428							Hi solids temp trip		
	1935	3	60	3	45	0.057	306003045			
	2113	3	60	3	50	0.055	306003050			
3/12/09	200	3	60	3	60	0.050	306003060			
	645							Close MSV02 Purge Air -- Slide Seal Air Closed		
	650	3	48	3	60	0.056	304803060			
	859	3	39	3	60	0.061	303903060			
	926	3	39	3	48	0.069	303903048			
	951	3	39	3	38	0.078	303903038			
	1052	3	29	3	29	0.103	302903029			
	1307	3	24	3	24	0.125	302403024			
	1316							solids inlet TC 103 Dropping -- Emptying MBHE		
	1318	3	29	3	29	0.103	302903029			
	1328	3	35	3	35	0.086	303503035			
	1244	3	28	3	28	0.107	302803028			
	1405							Inc cyclone loop seal ash takeoff fluidizing air		
	1428	3	24	3	24	0.125	302403024			
	1429							solids inlet TC 103 Dropping -- Emptying MBHE		
	1436							Inlet TC leveled out		
	1518							solids inlet TC 103 Dropping -- Emptying MBHE		
	1524							Restored loopseal ash takeoff air - increased bottom air flow		
	1535	3	30	3	30	0.100	303003030	Stratification on inlet tc		
	1545	3	32	3	32	0.094	303203032			
3/13/09	600							Unit stable		
	924							MSV 02 air purge on, increased leakage from slide seal		
	1015							solids inlet TC 103 Dropping -- Emptying MBHE		
	1054							MSV 02 air purge off		
	1106							MBHE refilled		
03/14/09	1051	3	32	3	32	0.094	303203032			
	1054	3	21	3	21	0.143	302103021			
	1106	3	14	3	14	0.214	301403014			
	1112							DAS 1 sec		

MBHE Data Event Key										
Date	Time	LACV dwell	LACV cycle	RACV dwell	RACV cycle	Flow Ratio	Flow Index	Comment	Pulse Integrated Flow	Ash Flow
	1117	3	10	3	10	0.300	301003010			
	1150	3	40	3	40	0.075	304003040			
	1158							Das 1 min		
	1349	3	32	3	32	0.094	303203032			
	1456	3	10	3	10	0.300	301003010			
	1550	3	32	3	32	0.094	303203032			
03/15/09								Decreased Right ACV Pulse air 66--->41 lb/h		
	925							Left ACV Air 40 lb/h , Right ACV Air 31 lb/h		
	942	3	10	3	10	0.300	301003010			
	1054							MBHE Draining		
	1100	3	12	3	12	0.250	301203012			
	1106	3	15	3	15	0.200	301503015			
	1336							Close PA air supply valve 1 turn		
	1412							DAS 1 sec		
	1415	2	10	2	10	0.200	201002010			
	1513	2	8	2	8	0.250	200802008			
	1533							MBHE Trip - high transport Press		
	1549							Restart MBHE		
	1543	2	10	2	10	0.200	201002010			
	1625	3	11	3	11	0.273	301103011			
	1751							MBHE Draining		
	1757	3	12	3	12	0.250	301203012			
	1819	3	15	3	15	0.200	301503015	Left ACV Air 40 lb/h , Right ACV Air 31 lb/h	7.1	6500
03/16/09	718	3	15	3	15	0.200	301503015			
03/16/09	720	3	12	3	12	0.250	301203012			
03/16/09	814							MBHE Slide seal ash leakage is significant		
	848	3	6	3	6	0.500	300603006			
	850							Overpressure trip		
	850	3	8	3	8	0.375	300803008			
	854	4	9	4	9	0.444	400904009			
	909							Overpressure trip		
03/16/09	915							Closed MSV 02 -- Stop Ash Flow		
	918	3	15	3	15	0.200	301503015			
	930	3	12	3	12	0.250	301203012			
	943							MBHE Empty		
	945							MSV05 closed		
	950							MSV 07 opened		
	953							PA valve closed		
	1000							MSV purge air at 1.5 psi		
	1002							DAS 1 min		
03/25/09								Replaced ash slide seal		
								PT212 Connected to MBHE Rupture disk Nozzle		
								PT213 Added - connected to Ash Inlet Pipe Vent Nozzle		
								Installed purge and bleed port on slide seal body		
								Installed 2 - 4" access ports on MBHE @ ash distributor		
								Vacuum ash from ash distributor		
								Video inspection of ash distributor and top of MBHE		
03/27/09	1500							MSV02 stuck in closed position		
03/28/09								Increased house air pressure to MSV02 still stuck closed		
03/29/09								MSV04 stuck open, cannot isolate transport pipe from combustor		
04/06/09	1406							PT213 connected to center vent on MSV02 - ash inlet pipe vent open		
	1539							Ash inlet vent closed with viewport		
	1542							PT213 connected to viewport		
	1552							Purge air tests on MSV02 - still stuck		
	1609							9 psi purge air on MSV02		
04/07/09	818							opening center cavity port on MSV02		
04/08/09								Rodded, air lanced and vacuumed center cavity - 2"ball valve installed in cavity vent		
	1510							Purge air vents open on MSV02		
04/08/09	1532							MSV02 opened		
	1550							Labview bug. -- restarted Labview		

MBHE Data Event Key										
Date	Time	LACV dwell	LACV cycle	RACV dwell	RACV cycle	Flow Ratio	Flow Idex	Comment	Pulse Integrated Flow	Ash Flow
	1559							MSV02 purge air at 2psi, bottom vent open		
	1604							start pulse air - high solids outlet alarm - reset alarm - restart		
	1606	2	33	2	33	0.061	203302033	start ACV		
	1629							Solids outlet temp alarm 1000F		
	1644							MSV02 purge air set to 1 psi		
	1708							MSV02 purge air off		
	1950							Slide seal purge air off		
	2013							MBHE full		
04/09/09	7:33	2	33	2	33	0.061	203302033			
	9:18	2	24	2	24	0.083	202402024			
	1241	2	18	2	18	0.111	201802018			
	1433							Slide seal leaking		
	1433	2	24	2	24	0.083	202402024			
	1625							Das 1 min		
04/10/09	839							Das 1 s		
	840	2	24	0	0	0.083	202400000	Left pulse air total = 32 lb/h		
	853							cycled right side 2 cycles		
	914							cycled right side 1 cycle		
	932							heel air off 2 cycles Wsole = 15 lb/h		
	933							sole air off 2 cycles Wheel = 26 lb/h		
	935							Heel + sole = 32 lb/h		
	1113							Slide sheal purge air on at 0.2 psi = 9" on MBHE top		
04/14/09		2	24	0	0	0.083	202400000	Left pulse air total = 32 lb/h	2.67	2900
04/15/09	1222							A boiler startup after outage, MSV02 opened --		
								Transport pipe plugged, ACV Left on during outage		
04/16/09								MSV 04 closed - will attempt to clear tranport pipe		
04/19/09	900							Transport pipe cleared - air flow/ash flow initiated		
	1000							ash flow initiated left side only		
04/20/09	0							Left and Right solids flow		
	1048	2	30	2	30	0.067	203002030			
	1130							MSV02 purge air at 10 psi to 3 ports		
	1410							MSV02 purge air at 0 psi		
	1539							Slide seal pressure increased from 2" to 9"		
	2000							Lost Labview		
04/21/09	400							A boiler trip		
	900							restart A boiler		
	1432	2	30	2	30	0.067	203002030			
	1423							Inc slide seal pressure ~ 10"		
	1603	2	30	0	0	0.067	203000000			
	1608							DAS 1s		
	1617							Left ACV Sole flow 18-22 lb/h		
	1622							Left ACV Heel flow 22-25 lb/h		
	1624							Left ACV Total flow 32 lb/h		
	1630							Dec Left ACV Heel flow 24-->21 lb/h		
	1640							Dec Left ACV Heel flow 23-->18-21 lb/h		
	1720	5	57	0	0	0.088	505700000			
	1804	2.5	27	0	0	0.093	252700000			
	2004	0	0	2.5	27	0.093	2527			
	2018							Right ACV Sole Flow 22 lb/h		
	2021							Right ACV Heel Flow 14 lb/h		
	2032							Right ACV Heel Flow Inc 18--> 25 lb/h		
	2035							Right ACV Total Flow Inc 44 lb/h		
	2047							Right ACV Total Flow Inc 40 lb/h		
	2052							DAS 1 m		
04/22/09	817	0	0	2	27	0.074	2027	MBHE drained		
	819	0	0	2	35	0.057	2035			
	1158	0	0	2	40	0.050	2040			
	1324							Right Ws+Wh = 40 lb/h		
	1325							DAS 1s		
	1330							Right ACV Heel air Wh = 22--> 18 lb/h		
	1333							Right ACV Total air Wh = 34 lb/h		
	1342							Right ACV Total air Wh = -->18 lb/h		

MBHE Data Event Key										
Date	Time	LACV dwell	LACV cycle	RACV dwell	RACV cycle	Flow Ratio	Flow Index	Comment	Pulse Integrated Flow	Ash Flow
	1344							Right ACV Total air Wh = 30 lb/h		
	1717							Right ACV Heel air Wh = 16 lb/h		
	1719							Right ACV Total air Wh = 24 lb/h		
	1826	0	0	2	30	0.067	2030			
04/23/09	740	0	0	2	20	0.100	2020			
	1305	0	0	2	14	0.143	2014			
	1300							Reset load on Ash inlet pipe constant load hangers - 1/2" lighter load		
	1337	0	0	2	16	0.125	2016	Inlet Temps dropping MBHE draining		
	1400							Aboiler vibration showing on Slide seal - effect of spring change		
	1600	0	0	2	15	0.133	2015	Right ACV Total air Wh = 24 lb/h	3.2	2000
04/24/09	1400							tightened leaking outlet union on Bundle R1 cooling water pipe		
04/25/09	700							Das 1 sec		
	720							Das 1 m		
04/30/09								DOE Review		
05/01/09	925			2	15	0.133	2015			
	930			2	12	0.167	2012			
	1520	2	12	2	12	0.167	201202012			
05/02/09	130	2	12	2	14	0.154	201202014	right side hi temp trip		
05/04/09	325	2	14	2	14	0.143	201402014			
	1600							trip on high solids outlet temp - reset - MBHE empty		
	1718	2	16	2	16	0.125	201602016	MBHE still empty		
05/05/09	930	2	20	2	20	0.100	202002020			
05/06/09	1115	2	18	2	14	0.125	201802014			
	1720	2	18	2	19	0.1081	201802019			
05/07/09	1122	2	22	2	22	0.0909	202202022	MBHE empty		
05/08/09	831	0	0	0	0		0			
	1430	2	27	2	27	0.0741	202702027			
	1700							Plant PLC failure MSV02 closed. Stuck Closed		
05/10/09								MSV02 opened - no ash flow - Hi MBHE pressure		
05/11/09								A-boiler in operation - no ash flow		
05/11/09	1700	2	27	2	27		202702027	High MBHE Pressure		
05/12/09	842	2	35	2	35		203502035	start ash flow		
05/13/09	1130							slide seal purge air off - 1 hr - reset at 0.3 psi		
05/13/09	1630	2	35	2	32					
05/14/09	838	2	35	3	31					
05/17/09	2040							A boiler off		
05/17/09	2330							Transport air off		
05/19/09	200							A boiler on		
05/19/09	640	2	35	3	31			Transport air on - Ash flow on		
05/21/09	843	2	35	5	31					
05/26/09	930	3	35	7	31					
05/26/09	2330							Ash inlet TC - TE102 Right Mid C failed		
06/05/09	910	3	35	10	31					
06/16/09	1027							No Right ash flow		
06/16/09	1028							1 sec labview logging rate		
06/16/09	1030	3	35	0	31			left pulse air 23-26 lb/h		
06/16/09	1041	0	35	3	31			No Right pulse air		
06/16/09	1042							Open plenum drains on Right ACV		
06/16/09	1108							Right Pulse air valve cycling		
06/16/09	1122							Increased Right Sole Flow 88 lb/h		
06/16/09	1142							Restored Right sole flow		
06/16/09	1153							Stopped AB ACV Flow		
06/16/09	1300							Closed Ash inlet MSV02		
06/16/09	1342									
								Took 2 ash samples from bottom inspection nozzles		
06/16/09	1347	3	35	6	31			Start Ash Flow - No right flow		
06/16/09	1816	3	35	5	31					
06/17/09	657							Stop ash flow - vacuuming ash		
06/18/09								DAS revision c:\alabview 6182009		
06/18/09	1226							DAS set pt for low transport air flow trip		
06/18/09	1406							Distributor vacuumed - 2" orifices installed		
06/18/09	1406							Replaced defective solenoids on ACV Pulse air-buttoned up MBHE - pushed slide seal rope back in groove		
06/18/09	1439							PA air flow to combustor @ 4000 lb/h		

MBHE Data Event Key										
Date	Time	LACV dwell	LACV cycle	RACV dwell	RACV cycle	Flow Ratio	Flow Index	Comment	Pulse Integrated Flow	Ash Flow
06/18/09	1512							Opened Ash inlet valve		
06/18/09	1535	3	35	3.1	31					
06/18/09	1543									
06/19/09	833							Start ACV - no pulse air - no AB pulse air indication		
06/19/09	839							Set AB to Manual control system ok - Recording time = 1s		
06/19/09	844	3	35	3.1	31			Dec Right ACV pulse flow from 49 to 38 - Left flow at 28		
06/19/09	903							Right Sole flow decreased from 38 to 34		
06/19/09	908	3	35	2.5	31					
06/19/09	914	4.1	35	2.5	31					
06/19/09	916							Stop ash flow - Hi pressure fan alarm - Plugged transport pipe		
06/19/09	931	2	35	2.1	31			Reset transport pressure trip from 68 to 60		
06/19/09	955							Blew ash from transport pipe through drain		
06/19/09	1008							PA fan on - no flow pipe still plugged		
06/20/09								MBHE Placed in service Fri pm - Right side overran and blew out slide seal - Ash inlet valve closed		
06/24/09								S Friend installed updated labview revision		
07/07/09	800							Slide seal repaired - box pressure = 13.9 inwg		
07/07/09	913							Ash inlet valve opened		
07/07/09	1220	2	35	1	31					
07/07/09	1630							no flow - High Solids temp alarm reset - start ash flow transport pipe trip - transport velocity too low - AB trip 60 " PA fan at 63"		
07/08/09	1200							transport pipe cleared - flow at 5500 lb/h		
07/09/09	850	1	59	1	60			Reset water flows -		
07/09/09	1122							MBHE in service - no labview data		
07/10/09	845							MBHE in service - no labview data		
07/10/09	1500							MBHE in service - no labview data		
07/13/09	2010							Labview started - Labview was off		
07/14/09	844							MBHE was shut down during past weekend - A boiler problems - Ash inlet valve stuck - personnel not available to open valve		
07/15/09	1030							MBHE on Line		
07/15/09	1240							MBHE tripped on high transport dp		
07/15/09	1652							MBHE continues to flow on right side after pulse air signals are off		
07/15/09	2010							asked Snodegrass to close Right Pulse air isolation ball valves - Stopped right ash flow		
07/16/09	900									
07/28/09	1534	2	44	2	45			Data review shows that ball valves stopped ash flow		
07/28/09	1715	2	44	1	55			ACV pulse air at 90 lb/h - Asked to have it lowered to 25-30 lb/h		
07/29/09	1100	2	44	1	55			Changed transport dp trip to 55"		
07/30/09	0							Ash flow 1700		
07/30/09	140							ash surge to 4000 lb/h		
07/30/09	730							Ash surge to 12000 lb/h - MBHE emptying - Ash transport dp spike - tripped		
07/31/09								A boiler tube leak		
08/03/09	830	1.2	58	1.2	58			Picture from A boiler cyclone shows limestone powder - A boiler back on line - asked operator not to put MBHE in service for 2 days.		
08/03/09	1240							started ash flow		
08/03/09	1440							ash surge to 8000 lb/h		
08/03/09	2100							ash flow 4000 lb/h		
08/04/09	1150	2.5	70	1.2	73			lost data transmission		
08/07/09								grant town having phone problems MBHE operating cycles changed		
08/10/09	620							Walked control room through procedure to put labview on line		
08/11/09	1510	2.5	70	1.2	73			ACV online - MBHE started up		
08/12/09	940	2.5	70	1.2	70					
08/12/09	943	2.5	70	0	70			Asked control rm to set right dwell to 0 then send operator down to left ACV and open pulse air inlet plenum for 15 min, the close and let mbhe operated for 1 hr		

MBHE Data Event Key										
Date	Time	LACV dwell	LACV cycle	RACV dwell	RACV cycle	Flow Ratio	Flow Index	Comment	Pulse Integrated Flow	Ash Flow
08/12/09	1125							Ash transport dp=7.7"		
08/12/09	1326	0	70	1.2	73			requested operators blow out right ACV plenum		
08/17/09	1018	0	7	1.2	73					
08/17/09	530							ash flow stopped		
08/17/09	640							ash flow started		
08/17/09	540							left ash flows increased		
08/17/09	1130							closed ash inlet valve		
08/18/09	1230							ACV pulse air off		
08/18/09	1254							MBHE isolated for vacuuming		
08/18/09	1549							AB constipated - historic screen filed selected		
08/20/09								MBHE vacuumed out, ash distribution orifices removed, access port added below bottom distributor plate, air lanced top of tube bundle through vent pipe, rigid see snake used to inspect top of tube bundle, top and bottom of left side of distributor, see snake too short to see inlet of ACV seal legs.		
08/20/09	1544							ash inlet valve open		
08/20/09	1614							start solids flow		
08/20/09	1617							stop AB solids flow		
08/20/09	1625							PA transport air at 5000 lb/h		
08/20/09	1645							left acv pulse air globe valves opened 2 turns		
08/20/09	1649	3	90	1.2	91			Start solids flow - left acv 85 lb/h		
08/20/09	1701							No left ACV flow		
08/20/09	1702	10	90	1.2	91					
08/20/09	1715							blowdown left ACV plenums		
08/20/09	1724							Stop ACV		
08/20/09	1730							left ACV pulse air max		
08/20/09	1731							start ACV		
08/20/09	1735	10	90	1.2	5					
08/20/09	1743							Stop solids flow		
08/20/09	1748							All pulse air to left ACV sole		
08/20/09	1749							start ACV		
08/20/09	1800							Stop ACV		
08/20/09	1805							All pulse air to left ACV heel		
08/20/09	1806							start ACV		
08/20/09	1819							closed Left ACV heel ball valve - opened Right ACV valve - reduced setting		
08/20/09	1823	20	45	1.2	5					
08/21/09								Transport DP alarm in and out - Transport pipe plugged - changed left ACV pulse air to same as right side		
08/24/09	1020							Transport pipe unplugged		
08/25/09	1000	0	90	1.2	91			Start ash flow		
08/25/09	1241							Transport pipe plugged - A boiler down to 1 cooler		
08/31/09								Requested MBHE start for 1 hr then stop ACV		
09/08/09	1930							A boiler down		
09/09/09	850							requested ash inlet valve be closed		
09/11/09								Requested ACV operate with ash inlet valve closed		
09/15/09								A - boiler outage will start thrus 9/17 end 9/27		
09/17/09	930							stop ACV		
09/28/09								MBHE Modifications 1-13		
10/05/09	1054							modify DCS logic to provide AB with pressure difference from Ps transport venturi - Average bed pressure		
10/06/09	750							connected venturi Ps - new tap on horizontal pipe before vertical rise to DP213		
10/06/09	1150							Cooling water on		
10/06/09	1637							Ash inlet valve stuck closed		
10/06/09	1701							Ash inlet valve open		
10/06/09	1727							AB ACV pulse air on		
10/06/09	1803	2	58	2	61			Right and left pulse air at 25-26 lb/h		
10/06/09	1812	4	58	4	61			ash flow started		

MBHE Data Event Key										
Date	Time	LACV dwell	LACV cycle	RACV dwell	RACV cycle	Flow Ratio	Flow Index	Comment	Pulse Integrated Flow	Ash Flow
10/07/09	653							labview logging at 1 sec		
10/07/09	740							labview logging at 1 min		
10/07/09	741	4	29	4	30			left pulse flow 26pph right 30pph		
10/07/09	819							condensate dripping out of ash inlet valve bottom packing		
10/07/09	1231							labview logging at 1 sec		
10/07/09	1255							labview logging at 1 min		
10/07/09	1342	4	21	4	19					
10/07/09	1631							labview logging at 1 sec		
10/07/09	1925							labview logging at 1 min		
10/07/09	1929							increased transport PA flow 2 turns 4500 -> 5300pph		
10/07/09	1935	4	12	4	19					
10/07/09	1953							labview logging at 1 sec		
10/07/09	2007							labview logging at 1 min		
10/08/09	800	4	50	4	55					
10/08/09	906							switched transport air to HP blower		
10/08/09	919	4	11	4	11					
10/08/09	1045	4	8	4	8					
10/08/09	1117							labview logging at 1 sec		
10/08/09	1122							labview logging at 1 min		
10/08/09	1600	4	5	4	6					
10/08/09	1809							left ACV not flowing		
10/08/09	1818	4	8	4	9					
10/08/09	1833							opened left ACV pulse air to 88 lb/h		
10/08/09	1918							hammered left ACV seal leg - flow started		
10/08/09	1930							decreased left ACV pulse air to 30 lb/h		
10/08/09	1950	4	50	4	55			Switch from HP blower to PA transport		
10/08/09	2000	4	15	4	16					
10/09/09	718	4	11	4	10					
10/10/09	920							Ash leak at transport pipe isolation valve inlet to combustor - MBHE shut down - solids drained		
10/12/09	1950							Transport air off - combustor inlet knifegate closed		
10/14/09								Gasket leak on knifegate inlet		
10/16/09								Ordered replacement gasket + Stainless studs		
10/19/09								Stuff shipped		
10/26/09								Gasket repair complete		
10/27/09	1300							Ash inlet valve opened		
10/28/09								Plant personnel busy - no ash flow		
10/29/09		4	33	6	31			Ash leak in Transport reducer near combustor inlet		
10/31/09		6	33	6	31			Ash leak repaired with uniram refractory		
11/04/09		9	33	9	31					
11/05/09		12	33	12	31					
11/06/09		15	33	15	31					
11/07/09	1140							Transport reducer ash leak MBHE out of service - entire circumference of reducer coated with uniram		
11/08/09	1150	18	33	18	31			MBHE back in service		
11/09/09	100							A-boiler tube leak MBHE isolated full of ash - Ash inlet valve closed - combustor transport inlet valve closed		
11/10/09								A-boiler in service		
11/13/09	1030	18	33	18	31			MBHE in service		
11/16/09	910	12	33	12	31					
11/16/09	1430							No right flow Operators varying ACV flow settings in attempt to restart flow		
11/17/09	1600	16	33	0	6			left flow only		
11/18/09	1600							Increased ACV pulse flow to 90 lb/h - right side ash flow sporadic		
11/19/09	1030							lost right side ash flow - significant air bypassing right side solenoid valve		
11/19/09	1300							Lost AB data storage		
11/20/09	820							Restarted AB		
11/20/09								Started receiving AB data - sent replacement solenoid valves to Grant Town		
11/23/09	1035	16	31	16	31			cycle change to eliminate aliasing with ACV/topps data transmission		
11/24/09	900							MBHE shut down for cleaning and reducer replacement		

MBHE Data Event Key							
Date	Time	LACV dwell	LACV cycle	RACV dwell	RACV cycle	Comment	Pulse Integrated Flow Ash Flow
12/21/09						MBHE cleanout and reducer replacement started	
12/28/09						Completed reducer replacement and cleanout. Little or no ash material within tube bundles, most removed by air lance, some by mechanical on top of tube bundles.	
01/04/10						Right side ash outlet plugged. MBHE must be reinstalled before cleanout ports on ACV can be installed.	
01/06/10						GT requested second set of cleanout ports on ACV above expansion joint. Installed baffle plates in return bend regions of tube bundles.	
01/07/10						Reinstalled door on MBHE	
01/11/10						Ash inlet valve opened, no ash flow. Ash inlet pipe plugged with hydrated ash above ash inlet valve, boiler outage required to remove spool and clean out ash inlet pipe.	
04/06/10						Ash inlet spool removed to clean out ash inlet pipe	
04/07/10						Ash inlet pipe cleaned and spool replaced. Boiler restarted - no ash inlet to MBHE for 2 days.	
04/10/10						MBHE restarted.	
04/12/10		6	16	6	16	MBHE running at 6-8Klb for weekend	
04/12/10	1000	6	12	6	12	Caused Transport DP trip	
04/12/10	1130	6	14	6	14	8-10klb ash flow	
04/12/10	1300	6	16	6	16		
04/13/10	925	12	32	21	31	changed DP trip point from 25 to 35 "wg	
04/13/10	1240	18	32	18	31		
04/13/10	1600					Periodic flow stoppages 3 on left and 1 on right ACV	
04/14/10	130					Right side ash flow stopped	
04/14/10	1600					Left side ash flow stopped - transport pipe plugged	
04/16/10	2000	16	32	16	32	Transport pipe unplugged	
04/19/10		16	32	16	32	Weekend flow - Pulse air flow ~55 lb/h	
04/19/10	1100	6.5	16	6.5	16	Decreased Pulse air flow - Left ~28, Right ~33 lb/h	
04/20/10	910	10	16	10	16	Transport velocity 4500-5000lb/h	
04/21/10	200	10	16	10	16	Right ACV started erratic flow ~30 min no flow	
04/21/10	1400	10	16	10	16	uniform flow	
04/21/10	1630	20	32	20	32		
04/22/10						A - boiler shut down	
04/26/10		12	32	12	32	Replacing right solenoid valve	
04/27/10						Replaced both ACV solenoid valves - sprayed oil into air line to solenoid	
04/28/10	1000					Adjusted ACV air flows left~32, right~36 lb/h	
04/28/10	1420					Started ash flow	
04/28/10	1430					Transport pipe plugged	
04/28/10	1930	12	32	12	32	Started ash flow	
04/29/10	950	18	32	18	32	ACV left and right stopped flow	
04/29/10	1000	12	32	12	32	Flow started	
04/29/10	2200	12	32	18	32		
04/30/10	830	18	32	18	32		
04/30/10	1400	18	32	18	32	trip on high DP	
04/30/10	1410	18	32	18	32	Trip reset to 35 "wg	
05/03/10	930	18	32	18	32	Periodic flow stoppages all weekend	
05/03/10	933	24	32	24	32	set to 24 sec dwell then reduced one sec/h until 18 then ran all week at 18	
05/07/10	1500	12	32	12	32		
05/10/10	955	12	21	12	21		
05/10/10	1330					ACV Supply air lubricator installed	
05/10/10	1900	12	21	12	21	DP trip at 30"wg	

MBHE Data Event Key							
Date	Time	LACV dwell	LACV cycle	RACV dwell	RACV cycle	Comment	Pulse Integrated Flow
05/10/10	2333	12	21	12	21	DP trip at 30"wg	
05/11/10	830	12	21	12	21	DP trip reset at 35"wg	
05/11/10	1100					Right side not operating consistently	
05/12/10	840	18	21	6	21		
5/14/110	115	10	21	18	21	Line plugging changed ACV settings	
5/14/110	1700	12	21	21	21		
05/15/10	945					A boiler off line - Ash leak on A boiler cyclone loop seal return chute at seal box	
05/17/10						Phone conference with GT re leak, asked GT to check spring haners to confirm hangers were unlocked from last outage, they were, then ok to uncouple MBHE from A boiler	
05/17/10						Plant disconnected Ash inlet pipe at inlet slide plate valve discharge flange, uncoupled MBHE from boiler	
05/25/10						Phone conference with GT . Alstom believes generic problem with Pyro units is cause of leak Alstom asked GT to measure expansion of loop seal during outage with MBHE disconnected.	