POWER SYSTEMS DEVELOPMENT FACILITY

TOPICAL REPORT

GASIFICATION TEST CAMPAIGN TC25

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ABSTRACT

In support of technology development to utilize coal for efficient, affordable, and environmentally clean power generation, the Power Systems Development Facility (PSDF), located in Wilsonville, Alabama, routinely demonstrates gasification technologies using various types of coals. The PSDF is an engineering scale demonstration of key features of advanced coal-fired power systems, including a KBR Transport Gasifier, a hot gas particulate control device, advanced syngas cleanup systems, and high-pressure solids handling systems.

This report summarizes the results of TC25, the second test campaign using a high moisture lignite coal from the Red Hills mine in Mississippi as the feedstock in the modified Transport Gasifier configuration. TC25 was conducted from July 4, 2008, through August 12, 2008.

During TC25, the PSDF gasification process operated for 742 hours in air-blown gasification mode. Operation with the Mississippi lignite was significantly improved in TC25 compared to the previous test (TC22) with this fuel due to the addition of a fluid bed coal dryer. The new dryer was installed to dry coals with very high moisture contents for reliable coal feeding.

The TC25 test campaign demonstrated steady operation with high carbon conversion and optimized performance of the coal handling and gasifier systems. Operation during TC25 provided the opportunity for further testing of instrumentation enhancements, hot gas filter materials, and advanced syngas cleanup technologies. The PSDF site was also made available for testing of the National Energy Technology Laboratory's fuel cell module and Media Process Technology's hydrogen selective membrane with syngas from the Transport Gasifier.

ACKNOWLEDGEMENT

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1.0 EXECUTIVE SUMMARY

Test campaign TC25 demonstrated the Power Systems Development Facility (PSDF) gasification process with high moisture lignite from the Red Hills Mine in Ackerman, Mississippi. TC25, occurring from July 4 to August 12, 2008, for 742 hours of air-blown gasification, was a continuation of Mississippi lignite testing first begun in March 2007 in test campaign TC22. Although gasification operation in TC22 was successful, operation of the coal feed systems was problematic due to a high as-fed lignite moisture content.

To improve coal feed operation with high moisture fuels, a fluid bed dryer was installed in the coal preparation structure for commissioning and initial operation in TC25. Further evaluation and characterization of gasification operation with the high moisture lignite after processing in the fluid bed dryer and the original coal mill system was the major test objective. Other test objectives included:

- Operating the fluid bed dryer system
- Testing the developmental coal feeder
- Testing instrumentation enhancements
- Evaluating hot gas filter materials
- Evaluating catalytic filter element concepts and carbon dioxide (CO_2) capture solvents
- Supporting commissioning by Media and Process Technology (MPT) of its hydrogen membrane
- Assisting with the installation and operation of the Fuel Cell Module developed at the National Energy and Technology Laboratory (NETL)

1.1 PSDF Overview

PSDF Mission. The Power Systems Development Facility (PSDF) is a state-of-the-art test center sponsored by the U.S. Department of Energy (DOE) and dedicated to the advancement of clean coal technology. Located in Wilsonville, Alabama, the PSDF is a highly flexible test center where researchers can economically evaluate innovative power system components on a semi-commercial scale.

Development of advanced power systems at the PSDF focuses specifically on identifying ways to reduce capital cost, enhance equipment reliability, and increase efficiency while meeting strict environmental standards. Current testing involves pressurized feed systems, coal gasifier optimization using a variety of fuels, sensor development, hot gas particulate removal, and syngas cleanup. The reliable operation of the integrated components allows the PSDF to effectively support additional testing of advanced gas cleanup, including low-cost CO_2 capture technologies.

Project Participants. The PSDF is operated by Southern Company Services. Other project participants during TC25 testing include the Electric Power Research Institute, KBR (formerly Kellogg Brown & Root), the Lignite Energy Council, and Peabody Energy.

1.2 Process Description

The PSDF gasification process, shown in Figure 1-1, features key components of an integrated gasification combined cycle (IGCC) plant. These include high pressure solids feed systems; a KBR Transport Gasifier; syngas coolers; a hot gas filter vessel, the particulate control device (PCD); continuous ash depressurization systems developed at the PSDF for ash cooling and removal; a novel piloted syngas burner; a slipstream syngas cleanup unit to test various pollutant control technologies; and a recycle syngas compressor.

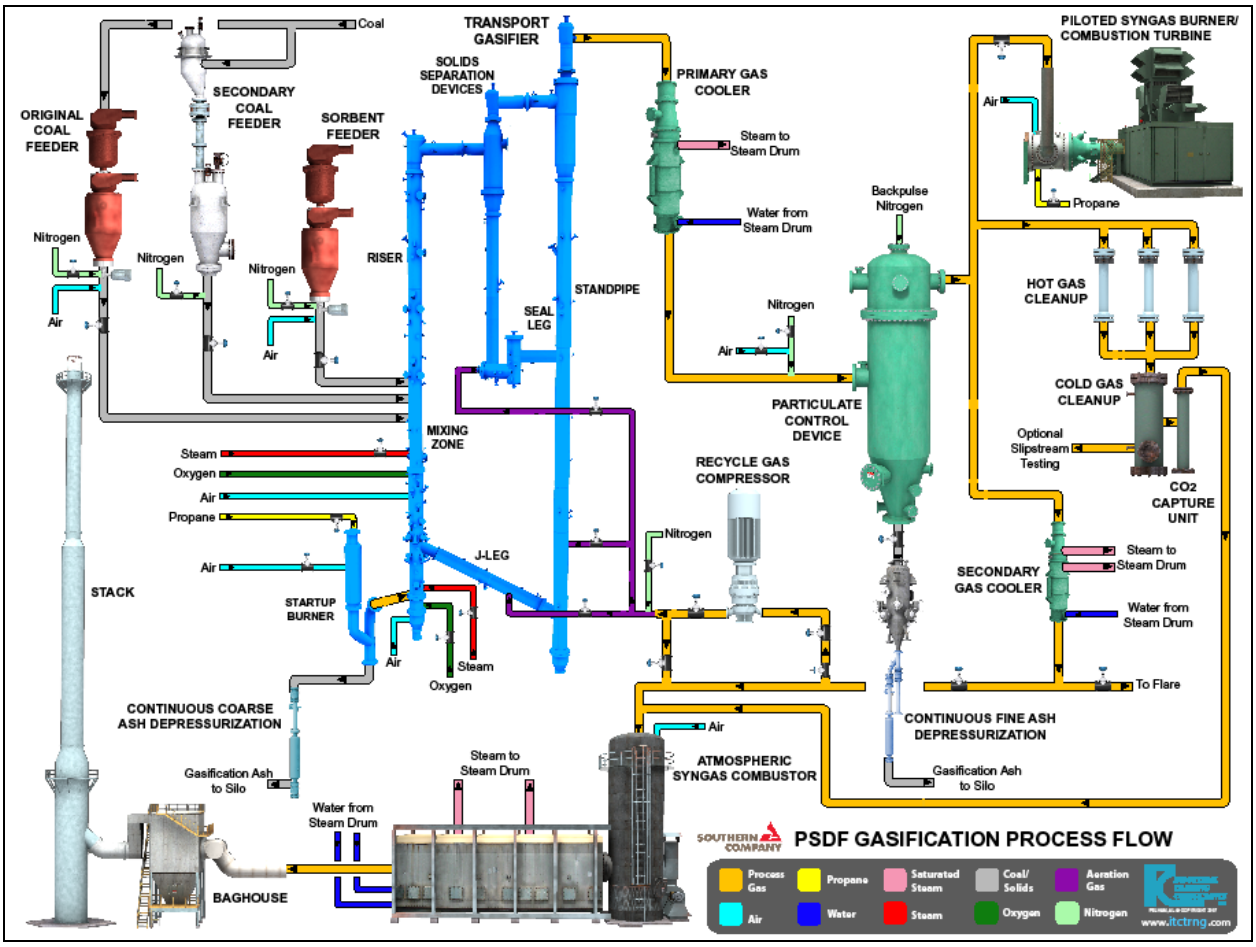

Figure 1-1. PSDF Gasification Process Flow Diagram.

Coal Preparation and Feeding. The coal used as the gasifier feedstock is processed on site. The material is crushed, dried in the fluid bed dryer (high moisture fuels only), and pulverized to a nominal particle diameter between 250 and 400 microns. Coal may be fed to the gasifier using two systems, the original coal feed system and a secondary coal feed system. The original coal feed system is a lock hopper, horizontal pocket feeder design with a "rotofeed" dispenser. It consists of two pressure vessels, with the coal pressurized in the upper lock vessel and then gravity fed into a dispense vessel, which is always pressurized. The material is fed out of the dispense vessel by the rotofeed dispenser, which is driven by a variable speed electric motor and delivers the material into the discharge line where it is conveyed by air or nitrogen into the

gasifier. The secondary coal feeder is a developmental test unit designed to evaluate different feeder mechanisms. Types of mechanisms that can be tested with this system include auger-style, fluid bed, and a higher pressure rotary feeder. Coal is fed at a nominal rate of 4,000 lb/hr.

Sorbent Feeding. A sorbent feeder is available to feed material into the gasifier for in-situ sulfur capture or to address ash chemistry issues. For sulfur capture, either limestone or dolomite is fed after being crushed and pulverized to a nominal particle diameter of 10 to 100 microns. The sorbent feeder utilizes the same design as the original coal feeder, but for a lower feed rate of nominally 100 lb/hr.

Start-up Burner. The start-up burner is a direct propane-fired burner operated to heat the gasifier to about 1,200°F. The burner is typically started at a system pressure of 60 psig, and can operate at pressures up to 135 psig.

Transport Gasifier. The Transport Gasifier (a pressurized, advanced circulating fluidized bed reactor) consists of a mixing zone, riser, solids separation unit, seal leg, standpipe, and J-leg. The gasifier is equally capable of using air or oxygen as the gasification oxidant. A mixture of or oxygen with steam is fed into the mixing zone at different elevations and orientations to evenly distribute heat generated from the partial combustion of the circulating solids. The oxygen from the air or pure oxygen feed is completely consumed in this section of the gasifier. The coal and sorbent are fed at a higher elevation in the mixing zone where the atmosphere is reducing, or oxygen-free.

Gasifier Solids Removal. As the coal devolatilizes and chemical reactions occur to generate syngas, the gas and solids move up the riser and enter the solids separation unit. This unit contains two solids separation devices, which use cyclonic action to remove particles. Between the first and second solids separation devices is the seal leg, which prevents backflow of gas. The solids collected by the solids separation unit are recycled back to the gasifier mixing zone through the standpipe and J-leg. The gasifier solids inventory is controlled by removing gasification ash through the continuous coarse ash depressurization (CCAD) system, which cools and depressurizes the solids. The nominal gasifier operating temperature is 1,800°F, and the gasifier system is designed to have a maximum operating pressure of 294 psig with a thermal capacity of about 41 MBtu/hr.

Syngas Filtration. The syngas exits the Transport Gasifier, passes through the primary gas cooler where the gas temperature is reduced to about 750°F, and enters the PCD for final particulate removal. The metal or ceramic filter elements used in the PCD remove essentially all the particulate from the gas stream. The PCD utilizes a tube sheet holding up to 91 filter elements, which are attached to one of two plenums. Process gas flows into the PCD through a tangential entrance, around a shroud, and through the filter elements into the plenums. Failsafe devices are located downstream of the filter elements to stop solids leakage by plugging in the event of element failures. High pressure nitrogen back-pulsing, typically lasting 0.2 seconds, is used to clean the filters periodically to remove the accumulated gasification ash and control the pressure drop across the tube sheet. The solids fall to the bottom of the PCD and are cooled and removed through the continuous fine ash depressurization (CFAD) system.

After exiting the PCD, a small portion of the syngas, up to 100 lb/hr, can be directed to an advanced syngas cleanup system downstream of the PCD. The syngas cleanup system is a specialized, flexible unit, capable of operating at a range of temperatures, pressures, and flow rates. It provides a means to test various pollutant control technologies, including removal of sulfur and mercury compounds, and CO_2 capture. The syngas cleanup slipstream can also be used to test other power technologies such as fuel cells.

Piloted Syngas Burner. A portion of the syngas can also be directed to the piloted syngas burner (PSB), a gas turbine combustor designed to burn coal-derived syngas. After syngas combustion in the burner, the flue gas passes through a 4 MW turbine before exiting the turbine stack. An associated generator can supply power from the turbine to the electric transmission grid.

Recycle Syngas. The main stream of syngas is then cooled in a secondary gas cooler, which reduces the temperature to about 450°F. Some of this gas may be compressed and sent to the gasifier for aeration to aid in solids circulation. The recycle gas compressor is a vertically mounted centrifugal compressor which operates at high temperature, nominally 500 to 600°F, and was designed for a throughput of about 2,000 to 3,000 lb/hr.

Final Syngas Processing. The remaining syngas is reduced to near atmospheric pressure through a pressure control valve. The gas is then sent to the atmospheric syngas combustor which burns the syngas components. The flue gas from the atmospheric syngas combustor flows to a heat recovery boiler, through a baghouse, and then is discharged out a stack. A flare is available to combust the syngas in the event of a system trip when the atmospheric syngas combustor is offline.

NOTE: A brief description of the PSDF gasification testing history can be found in Appendix A.

1.3 Major Test Campaign Objectives

Gasification Operations. The Transport Gasifier operated smoothly over a wide range of conditions, including:

- Maximum gasifier temperature: 1,700 to 1,820°F
- Gasifier pressure: 180 to 260 psig
- Coal feed rate: 3,500 to 4,900 lb/hr
- Air-to-coal mass ratio: 2.8 to 3.4 lb/lb
- Gasifier riser velocity: 18 to 28 ft/s

The gasifier carbon conversion was high, at over 97 percent for all conditions tested. In addition, the standpipe level operated at the highest level to date and the aeration flows were reduced to the lowest levels operated to date.

Coal Preparation. The fluid bed dryer operated well at average feed rates of about 13,000 lb/hr, reducing the moisture content from about 41 weight percent to about 19 weight percent. The dryer produced material with a mass median diameter (MMD) of about 250 microns. Although the mean particle size of the material was acceptable, the percentage of oversize particles, (those greater than 1,180 microns) was higher than desired. The dryer product contained 42 weight

percent of oversize particles and about 7 weight percent of fine particles smaller than 45 microns. The lignite material was pulverized in the existing coal mill pulverizers to reduce the amount of oversize material to about 17 weight percent. The additional pulverizing did not significantly impact the amount of fine material. The amount of fine particles increased by 1 percent up to 18 weight percent, which was within the acceptable limit.

Coal Feed Systems. The feed systems combined fed about 2,800 tons of lignite to the gasifier. Modifications to the discharge line on the original coal feed system significantly improved the reliability of the system by reducing the occurrence of material pluggages in the line. Only two discharge line pluggages occurred during 740 operating hours, as compared to more than 90 pluggages occurring in 543 hours of operation in TC22. The developmental coal feed system also operated well for over 650 hours at feed rates of about 2,500 lb/hr, although the feed rate was not as steady as desired. Control logic modifications stabilized the feed rate some, and additional modifications to the logic and feed system were planned for completion during the post-TC25 outage to improve the feed rate controllability further.

Sensor Development. Real-time particulate monitoring was provided at the PCD outlet by both the Process Metrix Process Particle Counter (PPC) and the PCME DustAlert-90 monitor. The modifications made to the PPC system prior to TC25 appeared to improve instrument sensitivity, and the instrument demonstrated improved accuracy in TC25. Although some initial problems were encountered with the syngas flow control that delayed startup of the system for a few days, once the system was on-line, it operated for over 500 hours without any required maintenance.

Fifty-one HR-160 thermowells and five ceramic thermowells were tested in the gasifier during TC25, as these materials have shown superior resistance to erosion in previous gasification tests. No thermocouple measurements were lost during the test campaign, and eight thermocouples that were inspected following the test campaign were in excellent condition.

Particulate Control Device. Continued evaluation of hot gas filter materials included iron aluminide sintered powder elements and two types of HR-160 sintered fiber elements (two fiber sizes). Three of the iron aluminide elements were filled with water gas shift catalysts for testing. After damage of these elements resulting from catalyst exposure to oxygen, these elements were removed and replaced.

Post-TC25 inspections showed pitting of the HR-160 elements, particularly with the finer fiber design. Flow test results for the iron aluminide filter elements showed the same trend seen in the past, with the flow resistance increasing with syngas exposure time. For both the clean and dirty iron aluminide elements, the rate of increase in pressure drop also appeared to be increasing with syngas exposure.

Advanced Syngas Cleanup. Since enhancing the syngas CO_2 and H_2 concentrations is desirable for applications such as CO_2 capture and fuel cell operation, testing of water-gas shift systems and catalysts continued in TC25. The advanced syngas cleanup unit was used in TC25 to test two water-gas shift catalysts installed in the same type of filter elements used in the PCD. The syngas cleanup slipstream was also used to support outside researchers from the DOE National Energy Technology Laboratory and from Media Process Technology. A Johnson Matthey

mercury sorbent was successfully tested during TC25 to separate mercury from syngas at high temperatures.

During TC25, about 40 absorption and 40 regeneration tests were performed with a bench scale Parr reactor to capture syngas CO_2. During testing, parameters such as temperature, pressure, gas flow rate, solvent concentration, and CO_2 concentrations were varied to determine the optimum operating conditions. Preliminary analysis showed promise for high CO_2 capture efficiencies.

MPT Hydrogen Membrane. MPT successfully tested its carbon molecular sieve membrane on coal-derived syngas. The performance of the membrane in terms of selectivity and permeation rate for hydrogen was stable and unaffected by syngas contaminants. Very high permeate hydrogen concentration was achieved, and nearly half of the hydrogen in the syngas was separated in a single pass.

NETL Fuel Cell. Testing of the NETL solid oxide fuel cell module, begun in TC24, was continued. Fuel cell operation on syngas lasted for a total of 200 hours. The experience gained and data collected through the combination of the exposure and electrochemical tests under real syngas conditions yielded valuable information to assist in future SOFC technology development fueled on coal-derived gas. During this period, a NETL modified gas chromatograph with inductively coupled plasma/mass spectrometer was also successfully commissioned for accurately measuring syngas trace metals at sub-ppm levels.

1.4 Report Structure

The following report presents the operational data and results of gasification technology development at the PSDF during TC24, compiled in the sections listed below.

Section 2 Coal Preparation and Feed — Discusses operation of the fluid bed dryer, the original coal feed system and the developmental feed systems. Presents coal moisture values and particle sizes and their effect on coal feed system performance.

Section 3 Transport Gasifier —Includes the major gasifier operating parameters and the gasifier performance as indicated by solids and gas analyses.

Section 4 Sensor Development — Discusses performance of real-time particulate monitors and gasifier instrumentation and operations of gas analysis systems.

Section 5 Particulate Control Device — Describes the hot gas filter particulate characteristics, PCD performance, and filter element testing.

Section 6 Advanced Syngas Cleanup — Describes testing of water-gas shift catalysts, CO_2 capture, the NETL-developed fuel cell module, the MPT hydrogen membrane, and a high temperature mercury sorbent from Johnson Matthey.

Section 7 Support Equipment — Describes operations of the start-up burner and recycle gas compressor.

Section 8 Conclusions — Summarizes the major conclusions and lessons learned from TC25 operation.

Appendix A — Provides a brief history of gasification operation at the PSDF.

Appendix B — Presents the steady state operating periods and the major system operating conditions for each period.

Appendix C — Shows material and energy balances.

Appendix D — Lists the acronyms and engineering units used in this report.

2.0 COAL PREPARATION AND FEED

A new fluid bed coal dryer was installed and commissioned to improve the drying capacity of the coal preparation system when processing high moisture fuels and thus expand the PSDF gasification process fuel envelope. This system offers economic advantages in a commercial application due to the use of waste heat to increase the process gas temperature. During TC25, the fluid bed dryer successfully reduced the Mississippi lignite moisture content low enough for reliable coal feeding. The lignite was processed first in a crusher, then in the fluid bed dryer, and finally in a pulverizer before being conveyed to the coal feed systems. During TC25, the original coal feed system and a developmental coal feed system, the PDAC (pressure decoupled advanced coal) feed system, operated for a combined 742 hours.

2.1 Coal Characteristics

The Mississippi Red Hills Mine lignite coal used during TC25 is a low quality fuel with an as-received heating value of about 5,000 Btu/lb and a moisture content of 42 weight percent. Processing in the fluid bed dryer and coal pulverizer reduced the lignite moisture content to about 17 percent, which increased the heating value to about 8,000 Btu/lb. Table 2-1 shows the coal characteristics after drying and pulverizing for the lignite sampled from the coal feeders. Hydrogen in the coal is reported separately from hydrogen in moisture.

Table 2-1. As-Fed Lignite Coal Characteristics.

	Average Value	Minimum Value	Maximum Value	Standard Deviation
Moisture, wt %	17.1	13.7	19.8	1.3
Carbon, wt %	46.0	40.5	50.0	1.6
Hydrogen, wt %	3.5	3.2	4.1	0.1
Nitrogen, wt %	1.0	0.9	1.0	0.04
Oxygen, wt %	17.1	13.7	21.4	1.0
Sulfur, wt %	0.6	0.4	0.8	0.1
Ash, wt %	14.8	12.0	22.0	1.7
Volatiles, wt %	37.1	33.9	40.6	1.5
Fixed Carbon, wt %	31.0	27.7	33.7	0.9
Heating Value, As Received, Btu/lb	8,000	7,230	8,380	190
CaO, wt % in Ash	15.6	11.4	19.0	2.3
SiO_2, wt % in Ash	44.2	40.2	52.3	3.6
Al_2O_3, wt % in Ash	20.2	18.0	22.8	1.1
MgO, wt % in Ash	3.2	2.6	3.8	0.3
Na_2O, wt % in Ash	0.4	0.3	0.7	0.1

2.2 Coal Preparation

The coal preparation system originally installed at the PSDF was designed to process bituminous coal, which contains about 10 weight percent moisture. During initial testing of lower rank fuels to develop fuel flexibility with the Transport Gasification process, operation of the preparation and feed systems was problematic. Several system modifications (first implemented in 2000) improved operation and allowed for effective processing of tested subbituminous and lignite coals. However, when the high moisture Mississippi lignite was first tested in TC22 (March 2007), the coal preparation system could not sufficiently dry the coal because of its higher relative amount of intrinsically bound moisture. During TC22, the as-fed coal moisture content averaged 28 weight percent, and this relatively high moisture content caused numerous coal feed operational problems. To improve coal feeding with high moisture coals, a fluid bed dryer was installed to decrease the moisture content to 20 weight percent or less.

2.2.1 Fluid Bed Coal Dryer System Description

The fluid bed technology employed for the new coal dryer system does not require any internal moving parts, operates with very high efficiency, and uses less water than conventional drying methods. Figure 2-1 is a flow diagram of the PSDF fluid bed coal dryer system.

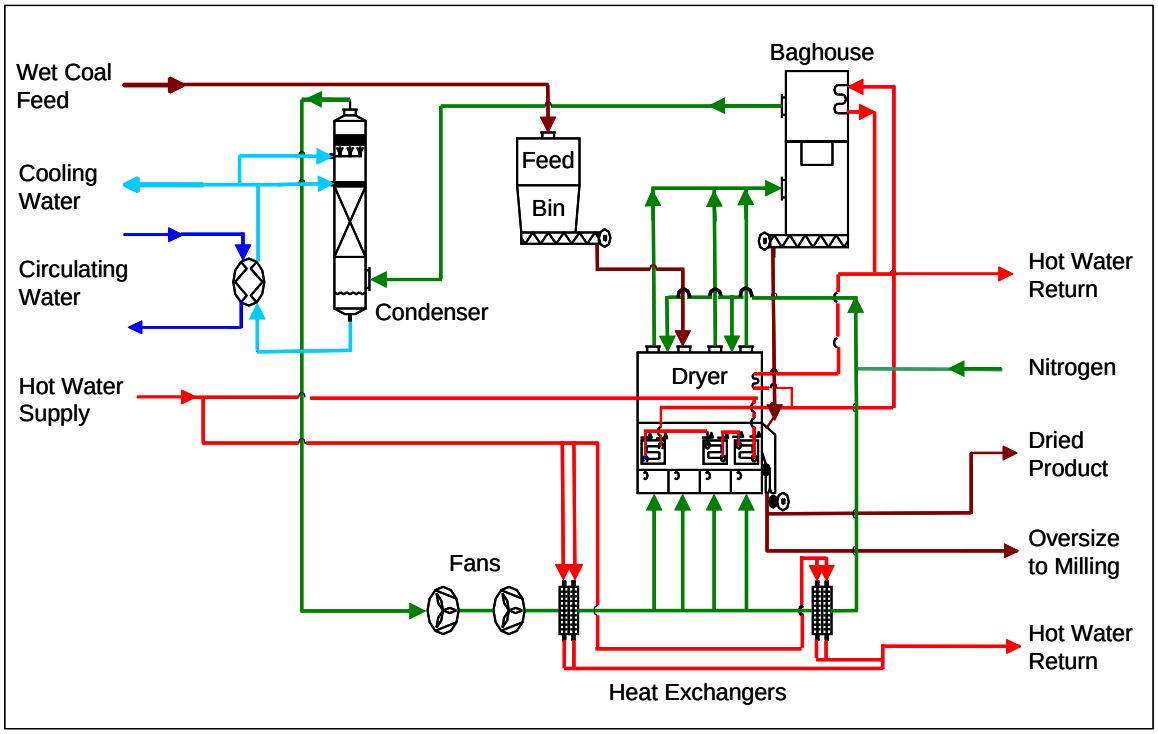

Figure 2-1. Fluid Bed Dryer System Flow Diagram.

After processing through a crusher, the coal is fed into the dryer feed bin and then from the dryer feed bin directly to the dryer by a variable rate feed system. Nitrogen is used for drying and

fluidization and is heated in a finned tube heat exchanger prior to entering the dryer. As the nitrogen and moist coal mix, the moisture transfers from the coal to the nitrogen. Three in-bed heat exchangers promote additional drying. The fluidized coal flows around and through the heat exchangers for efficient utilization of this drying energy.

The nitrogen at the top of the dryer is nearly saturated with water vapor. A slip stream of gas that bypasses the dryer is sent through a second finned tube heat exchanger to reheat the exit vapor preventing condensation of the gases in the exhaust duct or the baghouse. The gases pass through a baghouse, where entrained particulate is extracted from the gas stream and is conveyed to the dryer product outlet where it is mixed with the dried coal.

Exhaust gas from the baghouse enters a direct contact spray condenser, where the evaporated moisture from the lignite is condensed and extracted. A quench-water recirculation pump takes water from the condenser basin and discharges it above a packed bed to cool the process gas and condense the evaporated water. This condensed water goes to the process wastewater stream at the PSDF, but could be recycled in a commercial facility.

The water from the condenser basin is circulated through a heat exchanger to maintain a constant cooling water temperature. A cooling tower is used to provide cooling water to the shell and tube heat exchanger. The source of thermal energy for the dryer heat exchangers is a high temperature water heater operating at 300^{o}F. The operating conditions selected emulate waste heat streams at a commercial facility, further enhancing the efficiency of the operation.

The quenched nitrogen stream exits the condenser and passes through primary and secondary process blowers. Some gas may be exhausted to the atmosphere to control system pressure.

The crushed and dried coal is fed to a pulverizer where it is mechanically ground and contacted with heated process gas (mainly nitrogen) from an electric heater. By design, the pulverizer functions as a flash dryer with the heated process gas also functioning to convey the pulverized coal from the mill. This results in a very short residence time (approximately 1 to 3 seconds) during which only surface or “free” moisture is evaporated. As a result, high moisture coals containing a high relative amount of intrinsically bound moisture, such as lignite, retain much of this bound moisture through this process.

A cyclone separates the process gas (containing coal fines) from the pulverized coal. The separated process gas is then filtered in a baghouse and is dehumidified, heated in the electric heater, and recycled back to the pulverizer. The separated coal is screened, and oversize coal returns to the pulverizer for further milling. The pulverized coal product continues through a screw cooler, and is stored in a silo ready for use as gasifier feedstock.

2.2.2 Commissioning of the Fluid Bed Coal Dryer System

The fluid bed dryer system, manufactured by Schwing Bioset, was designed in early 2007. Construction and installation was completed in March 2008, and the initial commissioning was completed off-line in May 2008. Figure 2-2 is a photograph of the fluid bed dryer system installed at the PSDF.

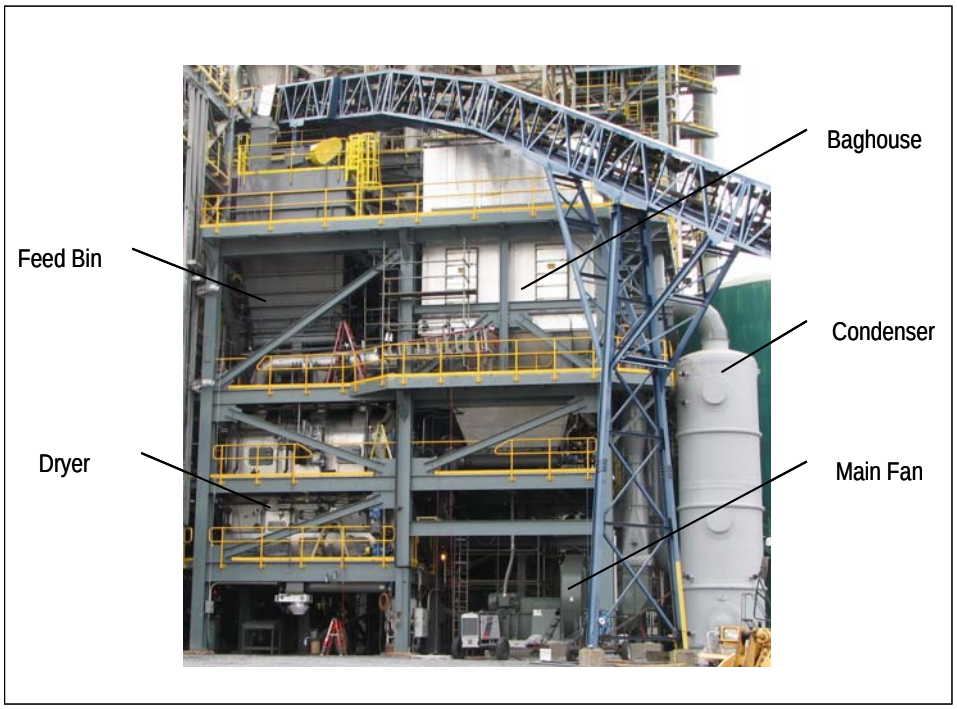

Figure 2-2. Fluid Bed Dryer System Installed at PSDF.

Fluid bed dryer commissioning started mid-March and ended in early May. Several operating challenges, mostly associated with support equipment, were resolved, and a number of parametric tests were successfully completed. The dryer operated on high moisture lignite successfully for 150 hours at the design coal feed rate of 13,600 lb/hr. Samples of the dryer feedstock, product, and oversize were collected during each test to determine the optimal operating conditions for particle size and coal moisture reduction. Heat transfer data was also collected to assess dryer thermal efficiency.

Most of the operational issues experienced were associated with support equipment and with incorrect manufacturers' installation. These issues and their resolutions included:

- The pneumatic feeder discharge valve failed to close (make up the closed limit switch) during the feed cycles, so the supply nitrogen line size was increased, and the control valve which actuates the main valve was replaced. A purge line with a solenoid valve was added to the valve body to prevent solids buildup. Also, the control logic was modified to increase the length of the purge cycle timer at the end of each conveying cycle.
- The pneumatic feeder discharge line plugged near the ash silo when sending material to this vessel. This caused the line to appear plugged near the feeder as well. The conveying and purge cycle times were further increased to ensure that the line was blown clean. Nitrogen boosters were added to the discharge line (most importantly near the ash silo) to improve material transfer. Regulator settings on the clean-out valve and the nitrogen supply pressure were also increased.

- Because one of the baghouse explosion doors was not set correctly as received from the manufacturer, the door leaked. The explosion door hinges were properly set and the door was re-sealed.
- A pressure regulator was added to the baghouse nitrogen backpulse supply to decrease supply pressure from 115 to 85 psig.
- The top and middle flanges on the baghouse had not been installed properly, as the bolts used were off specifications and not completely tightened, and the flanges were misaligned. Both flanges were seal welded and made pressure-tight.
- Because of improper baghouse assembly, a bag fell from the baghouse tubesheet and became lodged in the screw conveyor at the bottom of the baghouse. This caused the screw to over-torque and break the hanger bearing. The pieces of the hanger bearing were conveyed forward to the rotary feeder where they became lodged, causing the rotary feeder drive chain to break. Repairs to the hanger bearing and drive chain were completed. A subsequent inspection of the baghouse showed that 40 bags were installed in the tubesheet incorrectly, so these bags were re-installed.
- Bed fluidization was lost once during a parametric test to decrease temperature of bed material. The temperature of the bed material was increased, manual fluidization dampers on fluidization to the dryer were adjusted, and the oversize (reject) removal frequency was increased to address the problem.

Results of the commissioning tests included:

- Quantification of coal particle size attrition through dryer
- Quantification of oversize removal rate
- Dryer operation at different hot water supply temperatures
- Minimization of nitrogen consumption
- Confirmation of design mass and energy balances
- Development of programmable logic controls
- Identified steps to stop a CO increase
- Demonstrated that cleanout of the fluidization windbox was unnecessary
- Improved the nitrogen makeup and nitrogen vent control

Table 2-2 shows the fluid bed dryer operating performance during commissioning. Testing revealed that while particle attrition in the dryer reduced the coal particle size significantly, the percentage of oversize particles was higher than desired for coal feed system operation. Thus, the material was sent to the pulverizers for final product sizing. Figure 2-3 shows the particle size distribution curves for the raw lignite and the fluid bed dryer product.

Table 2-2. Performance of Fluid Bed Dryer during Commissioning.

	Average Value
Inlet Coal Moisture Content, wt %	45
Outlet Coal Moisture Content, wt %	20
Moisture Content Reduction, %	55
Inlet Coal MMD, micron	1,500
Outlet Coal MMD, micron	275
Outlet Coal Oversize (>1,180 micron) Content, wt %	20
Outlet Coal Fine (<45 micron) Content, wt %	20

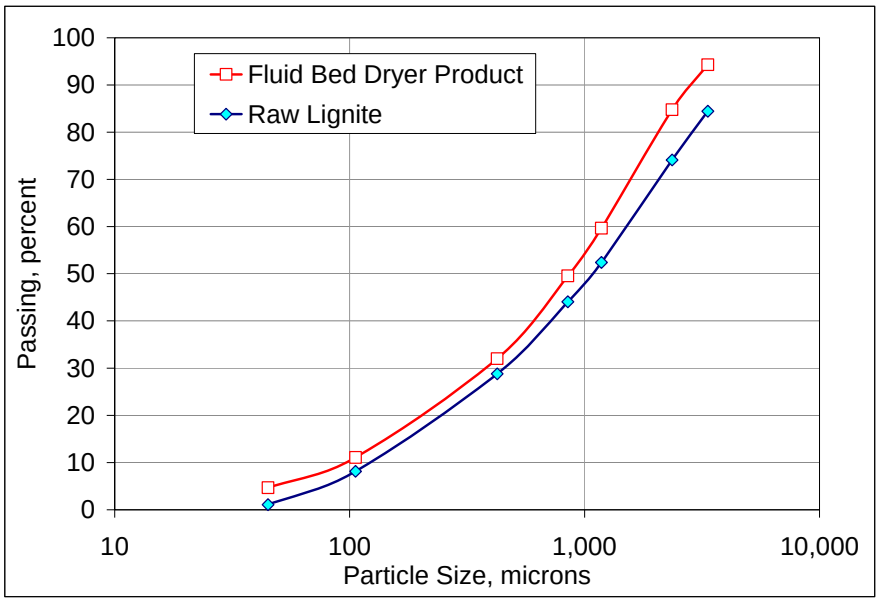

Figure 2-3. Particle Size Distributions of Lignite Processed in Fluid Bed Dryer and Raw Lignite.

Parametric testing consistently demonstrated the positive relationship between bed temperature and coal moisture content, and the operating data established the minimum bed operating temperature. Figure 2-4 plots the moisture content of the fluid bed dryer product versus the bed temperature. As expected, testing also showed that the hot water supply temperature had little impact on the coal moisture content.

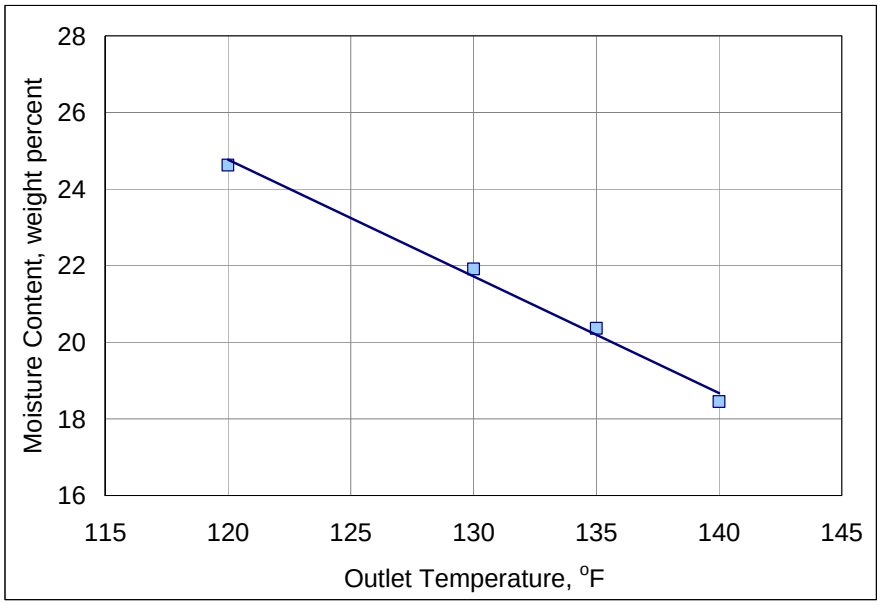

Figure 2-4. Product Moisture Content versus Fluid Bed Dryer Outlet Temperature.

The oversize (reject) removal rate was about 4 to 5 percent of the dryer feed rate. At this rate, dryer fluidization and bed differential pressure remained stable. The O_2 and CO analyzers provided accurate readings. Significant progress was made on control automation, with only three main control loops requiring additional tuning after completion of commissioning activities.

2.2.3 TC25 Operation of the Fluid Bed Coal Dryer System

In preparation for TC25, the fluid bed dryer was started on June 23, 2008, and operated for about 100 hours, processing 365 tons of lignite. The system performed well with no operational problems. The dryer operating temperature was held constant at about 135°F while the lignite flow rate varied from about 4,500 to 13,000 lb/hr to evaluate how the system operated at different flow conditions. The fluid bed dryer reduced the moisture content of the lignite from about 41 weight percent as-received to about 19 weight percent. The particle size of the dryer produced material, at about 900 microns MMD, was consistent with previous testing. The dryer product had 42 weight percent oversize material and about 7 weight percent fines (material less than 45 microns).

The lignite processed in the fluid bed dryer was then pulverized in the coal mill pulverizers to reduce the amount of oversize material. The pulverizers reduced the MMD to about 475 microns and the weight percent of oversize particles to about 21 weight percent. Figure 2-5 compares the particle size distribution curves for the fluid bed dryer product and the pulverized material. The additional pulverizing slightly increased the amount of fine material, which averaged about 18 weight percent. In addition, the pulverizers were operated without heat input from the electric heaters resulting in minimal added moisture reduction (lignite moisture content averaged 18 weight percent). The mill system was shut down after pulverizing about 195 tons of lignite.

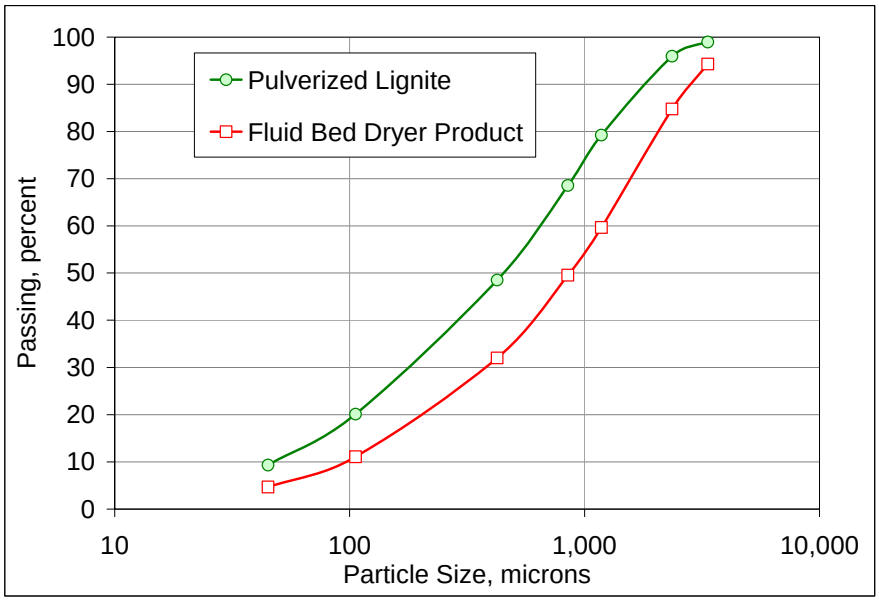

Figure 2-5. Particle Size Distributions of Pulverized Lignite and of Lignite Processed in Fluid Bed Dryer.

During on-line operation in TC25, the fluid bed dryer system operated without major operational issues for 469 hours, processing 2,800 tons of the high moisture lignite. The MMD particle sizes of the processed lignite as sampled from the coal feeders are shown in Figure 2-6. The lignite MMD particle size varied mostly from 300 to 500 microns for most of TC25 with some outliers ranging from 95 to720 microns. The lower MMD around Hour 650 resulted from accumulated fines as the silo was emptied in preparation for a test with a slightly larger particle size.

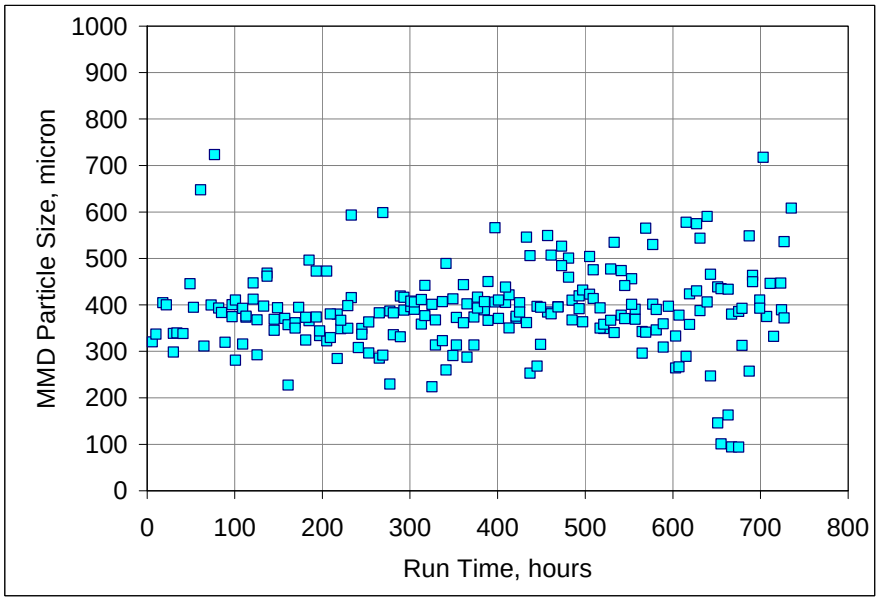

Figure 2-6. Particle Sizes of As-Fed Lignite.

Figure 2-7 shows the percentage of oversize coal (above 1,180 microns), and Figure 2-8 gives the percentage of fine coal (below 45 microns). The oversize particles varied mostly from 15 to 25 percent with outliers ranging between 5 and 33 weight percent of the coal. Fines concentrations varied mostly from 12 to 20 percent with outliers ranging from 9 to 32 percent.

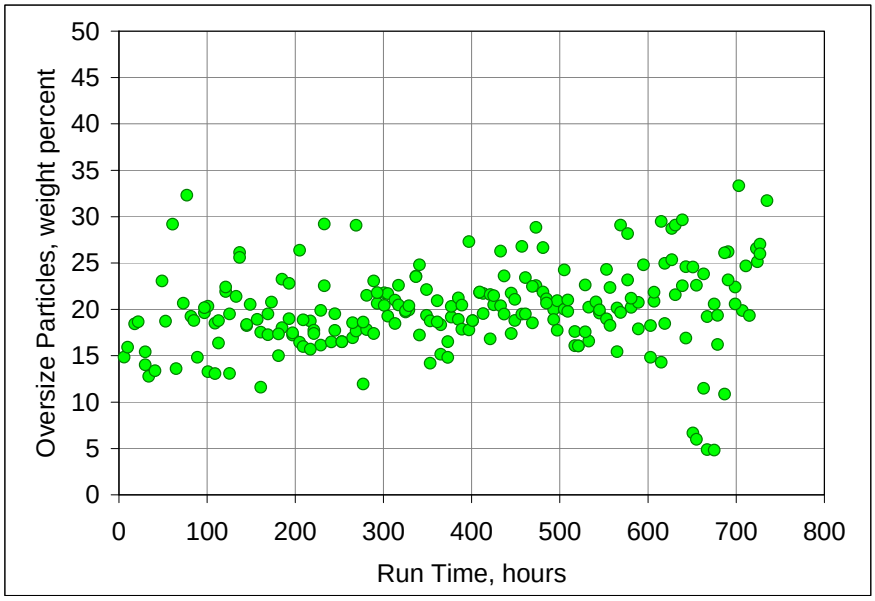

Figure 2-7 Percentage of Oversize Particles in As-Fed Lignite.

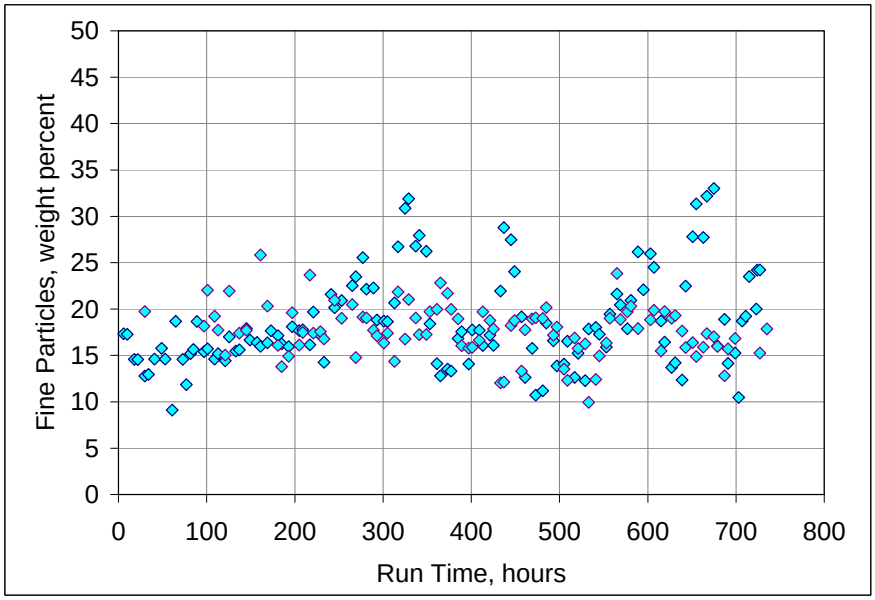

Figure 2-8. Percentage of Fine Particles in As-Fed Lignite.

Figure 2-9 plots compares the processed as-fed moisture content of the Mississippi high moisture lignite during TC22 (before the fluid bed dryer was installed) and during TC25, when the dryer was operating.

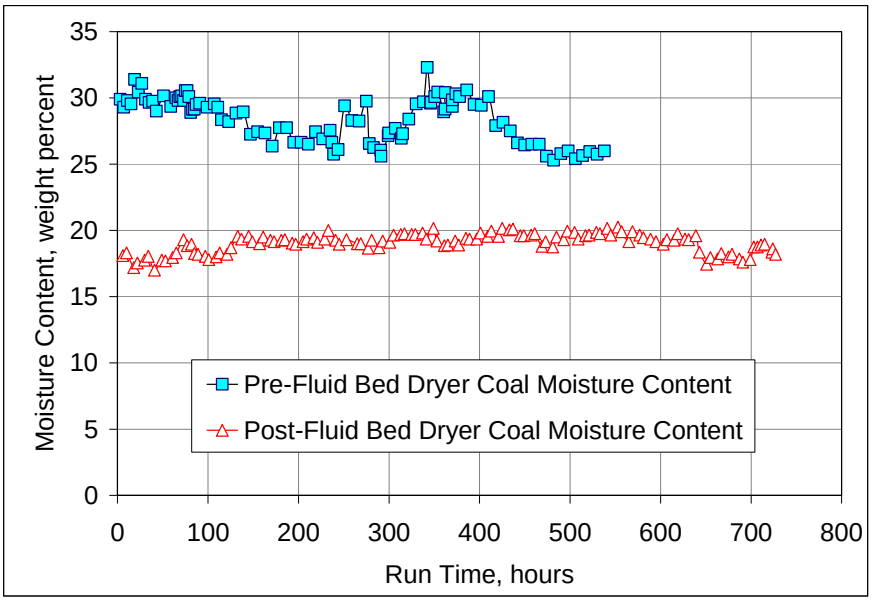

Figure 2-9. Moisture Content of As-Fed Lignite before and after Fluid Bed Dryer Installation.

2.3 Coal Feed Systems

2.3.1 Pressure Decoupled Advanced Coal (PDAC) Feeder

Installed to support high-pressure coal feeder development, the PDAC feeder is a non-mechanical feed control device with no moving parts. The design of PDAC combines some of the successful concepts developed at the PSDF such as continuous ash depressurization systems with traditional designs for flow rate control. The driving force for the solids flow is differential pressure. The nitrogen conveying gas meters the solids flow. Like the original feeder, the PDAC system is a lock hopper-based feeder. PDAC differs in that it uses conveying gas flow to control the solids feed rate.

The PDAC feed system was tested off-line from May 15 to May 23, 2008. Three main operating challenges were encountered:

- Repeated plugging of the lock vessel
- Conveying line differential pressure increase without a corresponding increase in density
- Frequent plugging of one of the off-line feeder vessel discharge lines

To identify needed improvements in the lock vessel operation, the pulverized lignite was tested in the laboratory and the cold flow unit. The testing revealed that the lignite had a high angle of repose and easily packed together when pressurized. Testing also showed the need for pressurization logic modifications to allow a slower pressurization ramp rate to the desired lock vessel pressure setpoint and to permit more frequent fluidization of the vessel during the fill cycle. In addition the lock vessel pressure setpoint was lowered relative to the pressure of the dispense vessel. By operating the lock vessel in this under pressurization state, additional fluidization of the lock vessel was achieved upon attempting to transfer material to the dispense

vessel. Additional data analysis showed that the conveying line pressure differential variations were due to the piping configuration of the off-line testing system. Booster nitrogen was added to address the off-line feeder plugging issue.

The PDAC feed system was restarted on June 4, 2008. After working through some initial issues and modifying the pressurization logic, the PDAC feed system operated well. The system was operated over a range of feed rates and was shut down on June 20, 2008.

The PDAC feed system was successfully operated for 441 hours in TC25. Modifications to the feed system were planned for the post-TC25 outage to improve feed rate control.

2.3.2 Original Coal Feed System

The original coal feeder performed well, operating at feed rates from 1,500 to 4,000 lb/hr and pressures up to 230 psig. Modifications to the discharge line were successful in allowing operations with a transport velocity of 40 to 45 ft/s without any discharge line plugging. Nitrogen was the coal conveying gas for the duration of TC25.

Figure 2-10 illustrates the original coal feeder operating envelope for coal particle size versus coal moisture content. The operating envelope (shown in green) represents reliable coal feeder operation during recent test campaigns (TC20 through TC25). The coal particle size and moisture content ranges from TC22, shown for comparison, were outside the reliable coal feeder operating envelope. Reliable coal feeder operation during TC24 with Utah bituminous coal extended the operating envelope to include very low moisture contents.

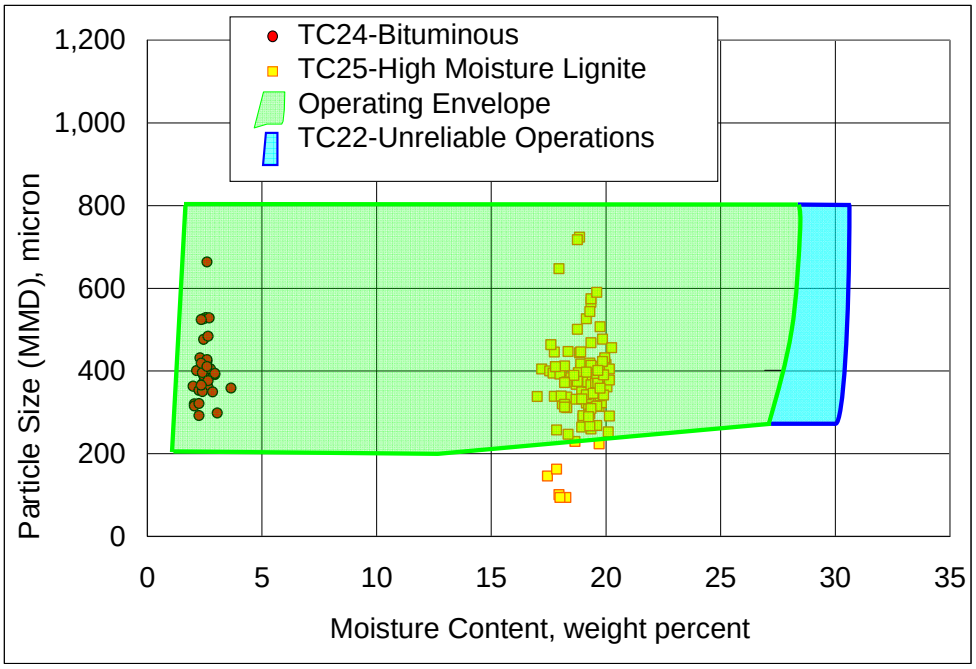

Figure 2-10. Original Coal Feeder Operating Envelope.

3.0 TRANSPORT GASIFIER

Test campaign TC25, a demonstration of the PSDF gasification process with Mississippi lignite, consisted of two main periods of operation, which were designated TC25A and TC25B. TC25A was a short period of operation (38 hours of lignite feed) which was ended due to the failure of experimental filter elements in the PCD. During TC25A, the coal feed and preparation systems operated well, and gasifier operations were stable until lignite feed to the gasifier was interrupted when the nitrogen supply to the process was compromised. During the upset condition caused by the lack of nitrogen, three experimental filter elements in the particulate collection device were damaged. To prevent damage to downstream equipment caused by the presence of particulate, the system was shut down for a short outage to replace the damaged filter elements.

After a one-week outage, gasifier operation with lignite feed resumed. This portion of the test campaign, designated as TC25B, consisted of 704 hours of operation. During TC25B, the gasifier was operated over a range of operating parameters to evaluate the effect on gasifier operations and performance. The carbon conversion was exceptionally high for lignite operation, at over 97 percent for all of the operating conditions tested.

All 42 of the TC25 steady state operating periods occurred during TC25B. These steady state periods are selected based on maintaining operating conditions in the gasifier within defined ranges. Appendix B lists the steady state operating periods and major operating parameters for each period.

3.1 TC25A Gasifier Operations

The system pressure test was completed on June 30, 2008, and was followed by start-up of the auxiliary equipment, preheating the PCD, and lighting the main start-up burner. Modifications to the start-up burner pilot were successful in addressing the initial lighting difficulty experienced in the previous test campaign. The gasifier temperature was increased slowly to reduce the thermal stress on the new refractory in the first gasifier solids separation device.

When the gasifier temperatures reached about 1,100°F, lignite feed began using the original coal feed system. The temperature was ramped up slowly to complete the refractory cure of the first solids separation device. The mixing zone temperature was gradually increased to 1,750°F, then decreased back to 1,400°F to troubleshoot the primary syngas cooler operation.

After 6 hours of coal feed (i.e., Hour 6) the PCD inlet temperature started increasing rapidly due to deterioration of the primary syngas cooler performance. Figure 3-1 plots the syngas cooler differential temperature for the first 24 hours of coal feed.

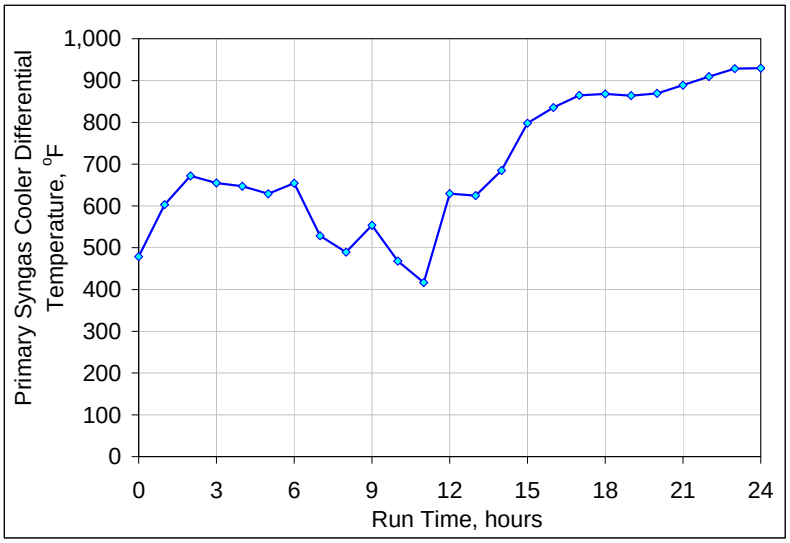

Figure 3-1. Primary Syngas Cooler Differential Temperature.

This rapid deterioration of the primary syngas cooler performance, despite the normal startup conditions in TC25, indicated that the fouling was likely due to residual material from the previous test campaign with bituminous coal when fouling was a serious operational issue. Although the gas cooler had been cleaned during the outage, the cleaning methods may have been inadequate in light of the severity and nature of the fouling. After the syngas cooler fouling became evident in TC25, several corrective actions were taken in order to remove the fouling and improve the gas cooler performance. These included:

- Increasing system pressure to increase residence time
- Increasing steam drum pressure to increase water-side operating temperature
- Reducing the total syngas flow
- Operating in combustion mode rather than gasification mode to destroy any hydrocarbons remaining on the cooler surface

Operation of the gasifier in combustion mode apparently was most effective in restoring the heat transfer capability of the syngas cooler. The differential temperature increased beginning at about Hour 11 as the fouling was removed. The gasifier was then transitioned back to gasification mode of operation.

At around Hour 39, a problem with a nitrogen vendor delivery caused a loss of nitrogen and led to oxygen breakthrough in the PCD. During this time, the oxygen concentration in the PCD was about 3 mole percent, which caused an exothermic reaction with the water-gas shift catalyst in the PCD filter elements. The damage to the elements, which led to high particulate concentrations at the PCD outlet, resulted in a system shutdown to replace the elements.

During TC25A, the maximum gasifier temperature was maintained around 1,800°F and the gasifier outlet temperature was around 200 psig. The Transport Gasifier operated well and the gasifier solids density and particle size decreased rapidly as the ash generated from lignite

gasification displaced the start-up bed material as observed in the previous test campaign with Mississippi lignite (TC22).

Figure 3-2 plots the gasifier solids bulk density during comparable operating times of TC22 and TC25. Figure 3-3 plots the gasifier solids Sauter mean diameter (SMD).

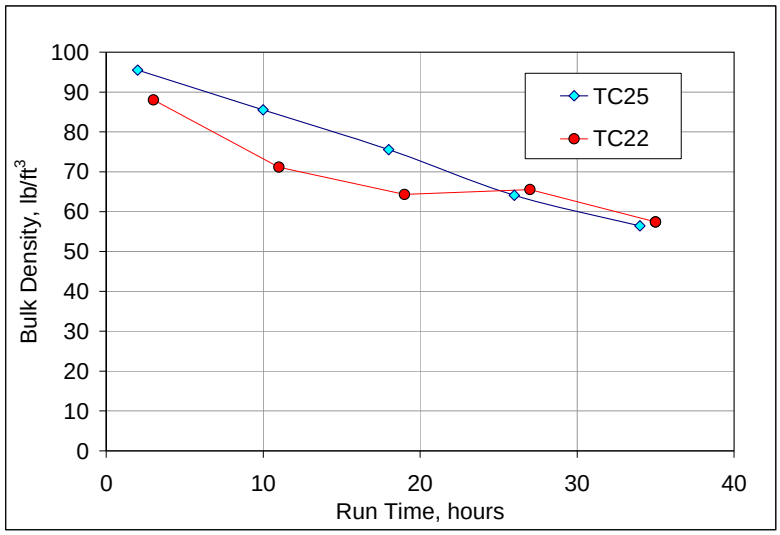

Figure 3-2. TC25A Gasifier Solids Density.

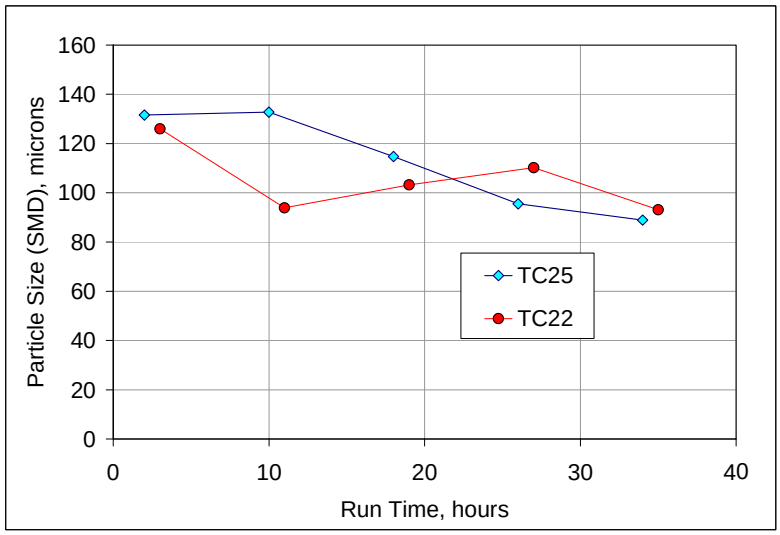

Figure 3-3. TC25A Gasifier Solids Particle Size.

3.2 TC25B Gasifier Operations

3.2.1 Operating Parameters

There were 42 steady state operating periods in TC25B, 32 with recycle syngas utilized for gasifier aeration. Ranges of the gasifier operating parameters are given in Table 3-1.

Table 3-1. TC25B Major Gasifier Operating Parameters.

Coal feed rate	3,500 to 4,900 lb/hr
Maximum gasifier temperature	1,700 to 1,820°F
Gasifier outlet pressure	180 to 260 psig
Steam feed rate	0 to 600 lb/hr
Gasifier riser velocity	18 to 28 ft/s
Standpipe level	100 to 210 inH_2O

Figure 3-4 plots the maximum gasifier temperature, the outlet gasifier temperature and pressure, and the primary gas cooler outlet temperature during the TC25B steady state periods. The primary gas cooler operated as designed, with an outlet temperature of about 800°F during TC25B.

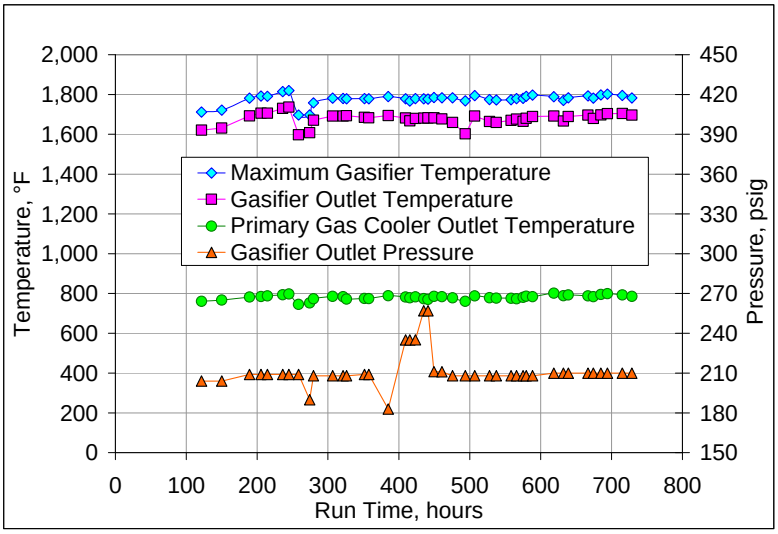

Figure 3-4. Gasifier and Primary Gas Cooler Operating Parameters.

Flow rates to the gasifier are shown in Figure 3-5. Coal feed rates were calculated from the feeder weigh cells, and the air, nitrogen, steam, and recycle gas flow rates were taken from flow indicators. The reported nitrogen flow rates were reduced from the indicated rates by 850 lb/hr to account for ash transport nitrogen. The coal feed rate was held fairly constant at around 4,000 lb/hr. The steam flow rate was minimal, less than 600 lb/hr, for the entire test campaign, and the recycle gas flow rate was held nearly constant around 1,000 lb/hr after the recycle syngas system was started at Hour 300.

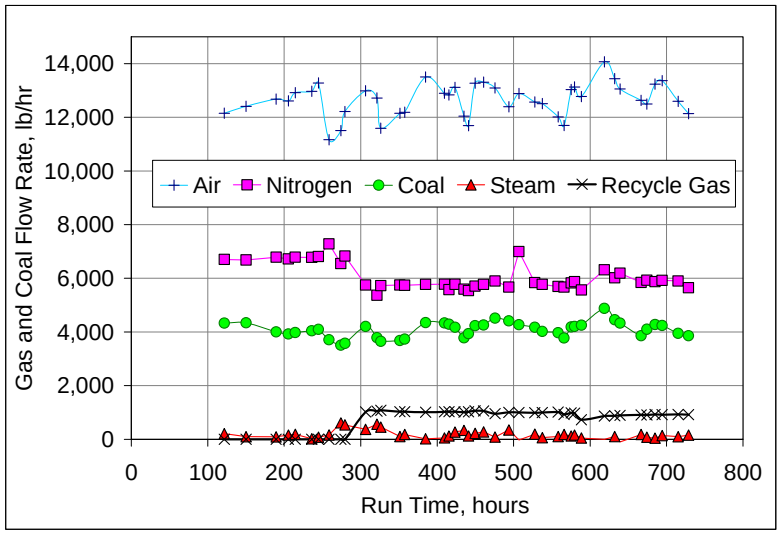

Figure 3-5. Flow Rates to the Gasifier.

Figure 3-6 shows the standpipe levels (measured as differential pressures) and the riser differential pressure. The standpipe level varied from 105 to 210 inH_2O, and the riser differential pressure varied from 30 to 84 inH_2O. Standpipe level variations were varied according to the test plan matrix.

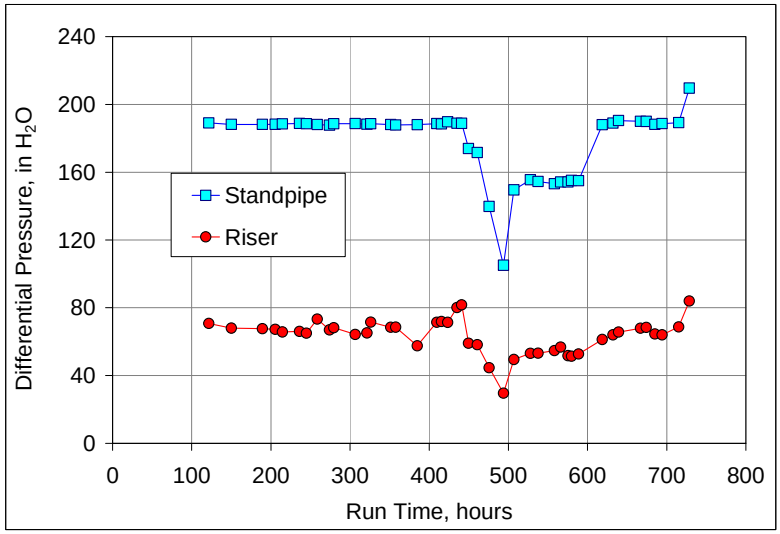

Figure 3-6. Gasifier Standpipe and Riser Differential Pressures.

3.2.2 Gasifier Performance

The gasifier operated well under a range of operating conditions, achieving high carbon conversion of over 97 percent for all operating periods. Table 3-2 gives the syngas lower heating value, carbon conversion, and hot and cold gasification efficiency for the steady state periods.

Table 3-2. Syngas Heating Value, Carbon Conversion, and Gasification Efficiency.

	Average	Minimum Value	Maximum Value
Syngas Lower Heating Value, Btu/SCF	58.3	52.5	63.2
Carbon Conversion, %	98.5	97.4	99.3
Hot Gasification Efficiency, %	88.3	86.6	89.5
Cold Gasification Efficiency, %	57.0	54.4	59.1

3.2.2.1 Solids Analysis

The gasifier solids chemical composition and particle size analyses represent both the circulating gasifier solids sampled from the gasifier standpipe and the PCD solids, which are sampled from the CFAD system.

Solids Chemical Analyses. The solids chemical analyses were used to evaluate the gasifier operations and performance of the solids separation devices. The chemical analyses of the gasifier circulating solids ash are given in Table 3-3, and Table 3-4 gives the chemical analysis of the PCD solids. The main constituents of both the gasifier and PCD solids were silicon dioxide (SiO_2), aluminum dioxide (Al_2O_3), and calcium oxide (CaO), and were similar in concentration to that of the coal ash. The heating value of the gasifier solids was negligible.

Table 3-3. Gasifier Circulating Solids Analysis.

	Average	Standard Deviation	Minimum Value	Maximum Value
SiO_2, wt%	46.5	2.2	43.0	50.1
Al_2O_3, wt%	21.8	1.2	20.3	23.3
Fe_2O_3, wt%	4.8	0.4	4.3	5.5
Other Inerts (P_2O_5, K_2O, Na_2O, BaO, MnO_2, SrO, & TiO_2), wt%	3.4	0.1	3.1	3.5
CaO, wt%	19.0	3.2	13.6	23.8
CaS, wt%	0.2	0.1	<0.1	0.5
MgO, wt%	3.7	0.2	3.4	3.9
Carbon, wt%	2.1	0.1	1.9	2.5

Table 3-4. PCD Solids Analysis.

	Average	Standard Deviation	Minimum Value	Maximum Value
SiO_2, wt%	42.8	2.9	37.3	46.7
Al_2O_3, wt%	19.6	1.0	17.7	21.3
Fe_2O_3, wt%	4.7	0.5	4.0	5.4
Other Inerts (P_2O_5, K_2O, Na_2O, BaO, MnO_2, SrO, & TiO_2), wt%	3.1	0.1	2.9	3.3
CaO, wt%	18.6	2.2	15.7	23.0
CaS, wt%	0.9	0.3	0.2	1.8
MgO, wt%	3.6	0.3	3.2	4.1
Carbon, wt%	7.9	1.6	5.2	13.1
Heating Value, As Received, Btu/lb	1,120	250	720	1,960

Solids Physical Analyses. The particle sizes of the gasifier circulating solids and PCD solids are shown in Figure 3-7. The MMD of the gasifier circulating solids varied from 60 to 170 microns while the SMD of the gasifier circulating solids varied from about 60 to almost 100 microns. The PCD solids MMD and SMD varied from 7 to 14 microns and 4 to 10 microns, respectively.

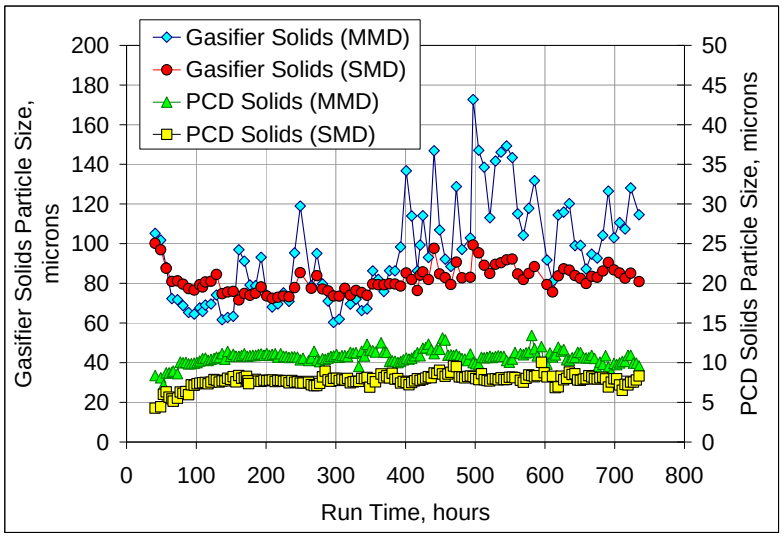

Figure 3-7. Particle Sizes of Gasifier Circulating Solids and PCD Solids.

Bulk densities of the gasifier circulating solids and PCD solids are shown in Figure 3-8. The bulk density of the circulating gasifier solids initially decreased from about 70 lb/ft^3 and then varied from 40 to 50 lb/ft^3. PCD solids bulk density also decreased initially and remained between 20 and 30 lb/ft^3.

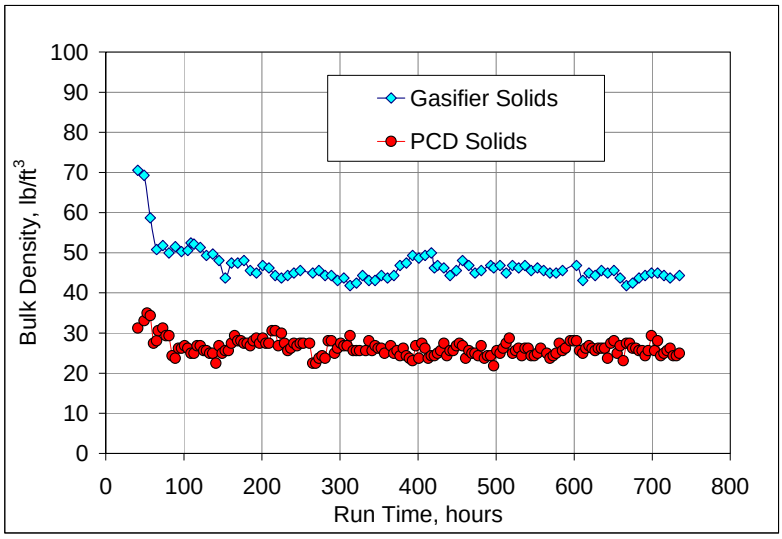

Figure 3-8. Bulk Densities of Gasifier Circulating Solids and PCD Solids.

Gasification Ash Removal. Figure 3-9 shows the removal rates for the fine PCD solids removed by the CFAD system and for the coarse gasifier solids removed by the CCAD system. The PCD solids rates were determined from the PCD inlet in-situ sampling, and the rates for CCAD were determined by a system ash balance. The PCD solids rate varied from about 260 to 480 lb/hr while the solids rates from the CCAD system varied from 150 to 460 lb/hr.

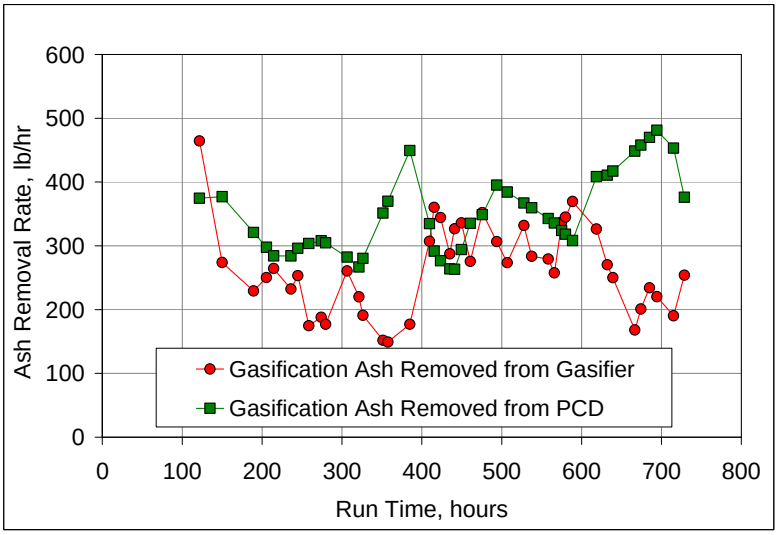

Figure 3-9. Gasification Ash Removal from the Gasifier and PCD.

3.2.2.2 Gas Analysis

Continuous extractive syngas sampling was performed between the primary gas cooler and the PCD inlet, and the syngas constituents were analyzed using continuous analyzers and gas

chromatography. Manual in-situ samples of syngas moisture were made at the PCD outlet during the particulate sampling.

The molar concentrations of the major syngas components for the steady state operating periods are given in Figure 3-10. The H_2, CO, CO_2, and CH_4 concentrations were measured by a gas chromatograph on a moisture free basis and converted to wet gas concentrations using the water concentration. An FTIR analyzer also made periodic measurements of the syngas moisture. The water concentration for steady state periods was estimated based on the PCD outlet sampling, FTIR moisture measurements, and a mathematical model based on the water-gas shift reaction equilibrium.

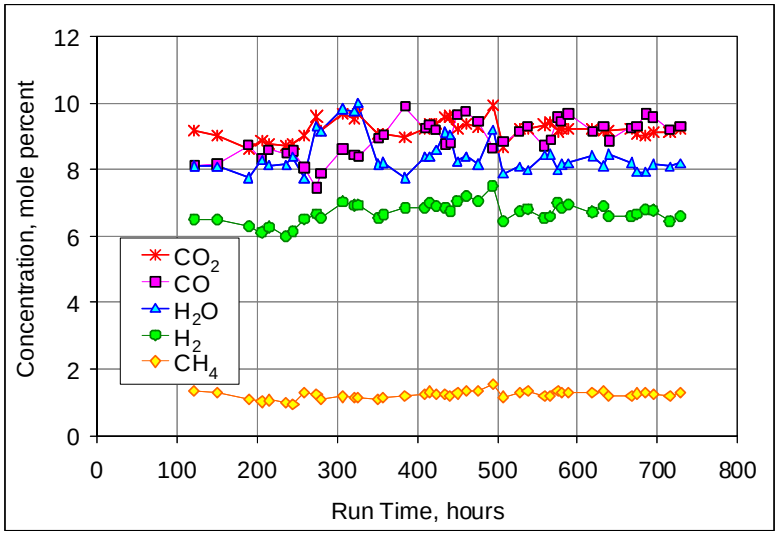

Figure 3-10. Concentrations of Major Syngas Components.

Minor constituents in the syngas include reduced sulfur compounds (mostly in the form of hydrogen sulfide, or H_2S) and reduced nitrogen compounds (mostly in the form of ammonia, NH_3). The H_2S concentration ranged from about 490 to 1,070 ppm on a wet molar basis. Ammonia concentrations were 1,170 to 2,140 ppm on a wet molar basis.

3.2.3 Parametric Testing

A number of tests were performed to establish boundary conditions of Transport Gasifier operations with high moisture lignite. The parametric testing included gasifier temperature, gasifier pressure/riser velocity, steam-to-coal ratio, and coal feed rate effects on gasifier performance. In addition, the effect of standpipe level on gasifier circulation rate was examined. To obtain meaningful analyses, data were analyzed using selected steady state periods which held other variables nearly constant to focus on the variable of interest.

Figure 3-11 gives the effect of gasifier temperature on carbon conversion during lignite operation in TC25 (at a gasifier pressure of 205 psig, air-to-coal mass ratios between 2.7 and 2.9, and coal feed rates ranging from 4,200 to 4,600 lb/hr) and in TC22 (at a gasifier pressure of 185 psig, air-to-coal mass ratios between 2.4 and 2.55, and coal feed rates ranging from 4,850 to 5,200 lb/hr).

As expected, the data shows a slight linear correlation between carbon conversion and temperature. The carbon conversion was higher at a given temperature during TC25 compared to TC22 due to higher efficiencies of the solids separation devices in TC25.

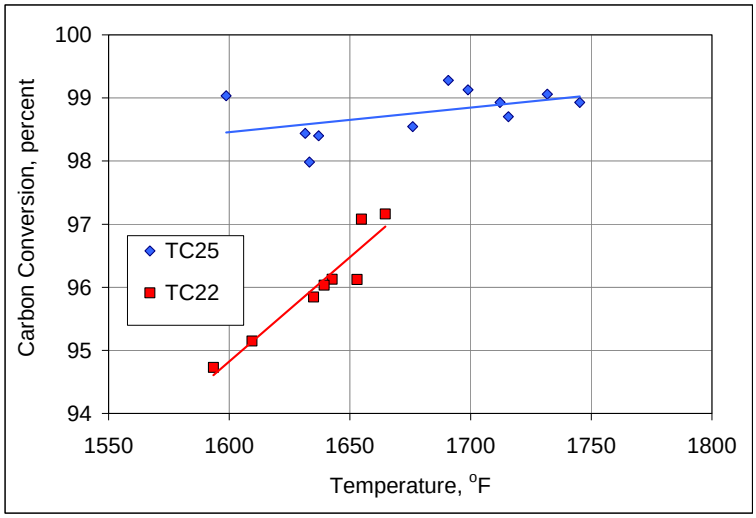

Figure 3-11. Carbon Conversion as a Function of Gasifier Temperature for Lignite Operation during TC22 and TC25.

Figure 3-12 shows the relative solids circulation rate as a function of gasifier standpipe level during lignite operation. As expected, the data indicate a strong linear relationship.

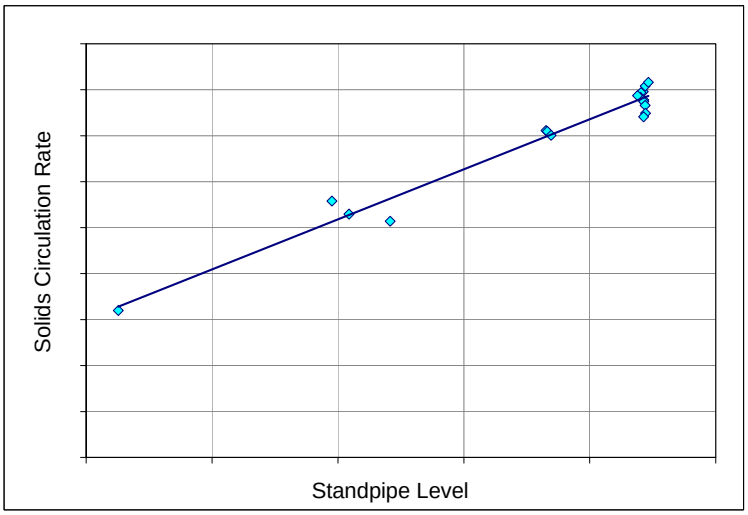

Figure 3-12. Effect of Standpipe Level on Circulation Rate for Lignite Operation.

Air feed to the gasifier is distributed to the lower and upper mixing zones of the gasifier to balance the temperature profile in the gasifier. However, at high gas flow rates to the lower mixing zone solids flow to the coarse ash removal system may be affected. Parametric tests were performed to evaluate the gasifier temperature profile and solids flow to CCAD as the air distribution was varied. Figure 3-13 shows the temperature difference from the maximum

temperature in the lower mixing zone to the maximum temperature in the gasifier as a function of percent air to the lower mixing zone. At low air flow to the lower mixing zone, the temperature difference was the lowest and the absolute temperature was too high to reduce the percentage below 20 percent. The temperature differential reached 480°F when the air flow percentage reached 55 percent; however the solids flow to the coarse ash removal system was unstable. The optimum percentage of air to the lower mixing zone based on these tests is about 35 percent. At this level, the temperature difference is about 100°F, which is optimum, and flow to the ash removal system is stable.

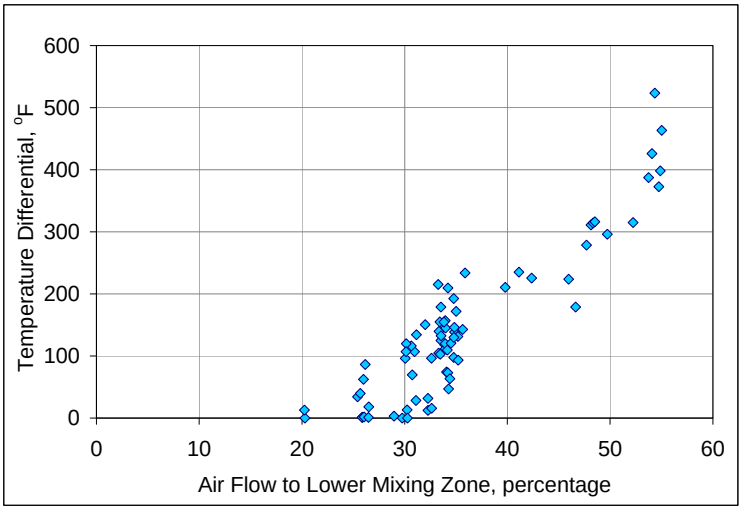

Figure 3-13. Effect of LMZ Air Flow on Temperature Profile.

3.2.4 Gasifier Inspections

Inspections of the Transport Gasifier and the primary and secondary gas coolers were completed. There were no agglomerations found in the gasifier; however, some refractory obstructing the gasifier circulating flow path was found in the slant leg of the seal leg. Figure 3-14 is a photograph of the refractory and the location from which it was removed. The refractory piece shown in the figure was approximately 18 inches by 12 inches. Chemical analysis of a sample from the piece removed confirmed that it was Sureflow 88 refractory so the piece did not originate from the first solids separation device. Visual inspection of the separation device supported that conclusion. The restricting refractory likely originated from the upper portion of the slant leg. Repairs were not necessary, and the obstructing refractory was removed.

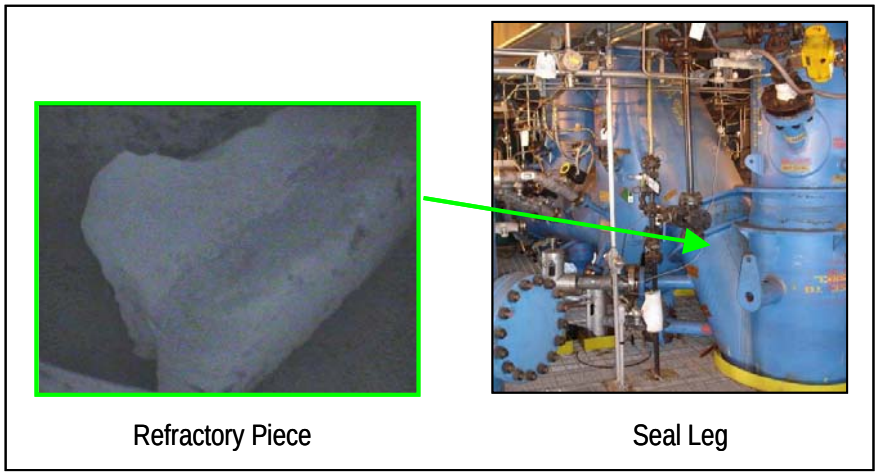

Figure 3-14. Gasifier Seal Leg Obstruction.

Inspection of the primary gas cooler top tubesheet showed minimal wear and solids buildup. Two of the heat exchange tubes contained a small amount of collected solids, which was easily removed with a scraper. Figure 3-15 shows the tubesheet before and after cleaning.

Figure 3-15. Primary Gas Cooler Inlet Tubesheet.

The primary gas cooler shell side at the riser was also inspected. There was no evidence of corrosion or deposition. A metallurgist was consulted and confirmed that there were no significant deposits, signs of oxygen pitting, or significant corrosion and that the tubesheet condition looked acceptable during a visual inspection. Figure 3-16 is a photograph of the primary gas cooler shell side.

Figure 3-16. Primary Gas Cooler Shell Side.

Inspection of the secondary gas cooler showed no tar buildup or evidence of condensate leaks. The outlet tubesheet is shown in Figure 3-17.

Figure 3-17. Secondary Gas Cooler Outlet Tubesheet.

4.0 SENSOR DEVELOPMENT

4.1 Real-Time Particulate Monitors

Two real-time particulate monitors, the PCME DustAlert-90 and the Process Particle Counter (PPC) by Process Metrix, were evaluated in TC25.

4.1.1 Process Metrix Process Particle Counter

The PPC instrument system is a light-scattering single particle counter operating on an extracted syngas sample at the outlet of the PCD. The optics in the counter are optimized to the range of particle sizes from 2.5 to 75 microns, which is the range that would detect leaks that might damage the downstream equipment. Though the use of a single particle counter has the potential to provide measurements at low concentration, the very low particle counts experienced at the PCD outlet during normal operations caused inaccurate measurements.

Several critical improvements were made to this measurement system prior to TC25. The system optics were modified to produce a particle counting volume about 8 times larger than the original. This increased the numbers of particle counts by 8 times, but decreased the ability to measure high particle concentrations.

In addition to modifying the optics to increase particle counts, the software was modified to provide an output of total particle counts to allow tracking low concentrations better in near real time. The normal output of the PPC is a calculated total particle mass concentration in mg/acm. The mass calculation algorithm developed by Process Metrix requires a minimum number of counted particles in a number of size ranges during an integration period to calculate the mass concentration in the syngas. Therefore, in the past, long integration times were required with the loss of near real time data and in many cases there were not sufficient particles to conduct the calculation.

The output of the PPC on July 23, 2008, when collection efficiency was high (low particle count) is shown in Figure 4-1. The graph shows both the particle count and the calculated mass concentration for a 20-minute period. The particle count is a continuous trace with large spikes when the PCD is backpulsed. With the 10-second integration time used during this test, the particle mass concentration is calculated for the backpulse spikes and thus is a non-continuous trace. From this graph, the particle count is high during a backpulse indicating that particle penetration occurs during the backpulse. However, interpretation of mass concentration data is more problematic. The average concentration over the entire 20-minute period shown was 0.08 mg/acm, while the spikes are only slightly higher than this and are of short duration. This result seems to be counterintuitive, but is a function of the calculation method and the very high peak to average concentration values downstream of the PCD, as explained in the following paragraph.

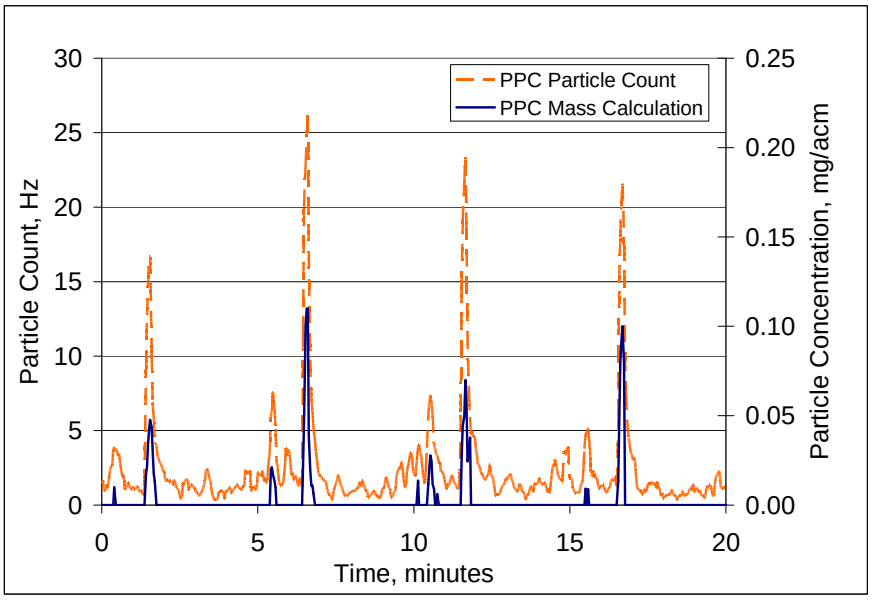

Figure 4-1. PPC Particulate Monitor Output during Low Particulate Loading on July 23, 2008.

When a 20-minute average is conducted with the PPC software over the period shown in the figure, the algorithim does not average the mass values of the 120 individual 10-second integrations, but sums all particles counted in each size range during the 20-minute period and calculates a new mass concentration. The large (massive) particles are being counted during the whole 20-minute period, but the numbers in any 10-second integration are too small to calculate a mass concentration. However, when all these large particles are included with the smaller particles counted during the backpulse spikes the value calculated is larger than expected. The effect of the large particles can been seen in the MMD values calculated. During the backpulse spikes the MMD is typically around 10 microns with few particles larger than 20 microns. For the 20-minute average, the MMD was calculated to be over 20 microns with significant numbers out to almost 50 microns. Obviously, interpretation of PPC data must be considered carefully. Near real time results can be useful for understanding the process, but a knowledge of the functionality of the system is necessary to avoid misinterpretation of the data.

Values of mass concentrations calculated by the PPC are compared with particle measurements made in-situ with the SRI sampling systems in Table 4-1. Except for Run 13, which was measured when the particulate concentration was relatively high because of solids injection in the PCD outlet duct, all the runs were averaged over 4 hours. Despite the long sample time, the in-situ sampling system was not able to resolve a value below the lower concentration measurement limit of 0.1 ppmw. The concentrations measured by the PPC were comparable and ranged from 0.004 to 0.017 ppmw.

During the solids injection test on July 28, 2008, the in-situ sampling system indicated a particulate concentration of 0.32 ppmw compared to 0.37 ppmw calculated by the PPC. This is remarkable agreement and this comparison will be repeated in future test campaigns. The average mass concentration from the PPC in the hour prior to the start of solids injection was 0.005 ppmw, suggesting an increase of 74 times resulting from the injected particulate.

Table 4-1. Comparison of PPC and In-Situ Concentrations.

Date	Outlet Run No.	Start Time	End Time	PPC				In-Situ Loading ppmw
				Loading,		Mass > 10 microns		
				mg/acm	ppmw	% of Total	ppmw	
7/21/08	7	8:45	12:45	0.028	0.004	81.20	0.003	<0.1
7/22/08	8	8:30	12:30	0.040	0.005	88.34	0.005	<0.1
7/23/08	10	9:00	13:00	0.048	0.007	88.71	0.006	<0.1
7/24/08	11	8:45	12:45	0.040	0.005	90.23	0.005	<0.1
7/25/08	12	8:45	12:45	0.028	0.004	87.73	0.003	<0.1
7/28/08	13	9:15	10:00	2.42	0.37	82.05	0.305	0.32(1)
7/29/08	14	8:45	12:45	0.024	0.003	83.59	0.002	<0.1
7/31/08	15	8:30	12:30	0.024	0.003	86.43	0.003	<0.1
8/1/08	16	9:00	13:00	0.020	0.003	83.38	0.002	<0.1
8/4/08	17	8:45	12:45	0.016	0.002	84.11	0.002	<0.1
8/5/08	18	9:00	13:00	0.024	0.003	85.55	0.003	<0.1
8/7/08	19	9:00	13:00	0.048	0.006	81.33	0.005	<0.1
8/08/08	20	8:30	12:30	0.032	0.004	77.54	0.003	<0.1
8/11/08	21	8:30	12:30	0.020	0.003	70.78	0.002	<0.1
8/12/08	22	8:30	12:30	0.012	0.002	75.66	0.001	<0.1

Notes: 1. Dust injection for PPC calibration testing.

Also shown in Table 4-1 are the percent of mass and the mass concentration that the PPC determined to be larger than 10 microns. These are the particles most likely to damage downstream components due to erosion. The amount of mass larger than 10 microns corresponded to 71 to 90 percent of the total mass, corresponding to 0.001 to 0.006 ppmw, well below the particle concentration limit for General Electric (GE) turbines of 0.024 ppmw. This is the first time that it has been possible to make an estimate of the large particle concentration at the PCD outlet, and although unconfirmed by other experiments, these results are reassuring.

The response of the PPC to the solids injection test is shown in Figure 4-2. The steady-state comparison period for the in-situ measurement shown in the table is indicated on the figure. Prior to the 9:04 am start of injection the background concentration can be seen as discrete small spikes with mass calculation possible only after backpulse spikes. Injection of solids produced a dramatic increase in PPC output and there was enough mass for a continuous calculation of concentration.

In addition to improved accuracy in TC25, operation of the PPC was very reliable. Although some initial problems were encountered with the syngas flow control that delayed startup of the system for a few days, once the system was on-line, it ran for over 500 hours without maintenance. This is partially attributable to the Mississippi lignite fuel, which does not produce much condensable tar or other organic compounds to contaminate the optical windows. Operation of the PPC with bituminous coals is expected to remain problematic.

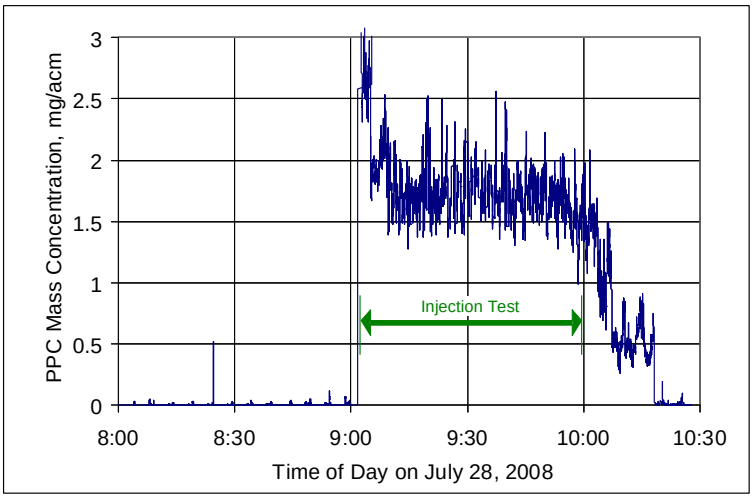

Figure 4-2. PPC Particulate Monitor Output during Solids Injection Test.

4.1.2 PCME DustAlert-90

The PCME DustAlert-90 particulate monitor (referred to as the PCME) was operational throughout TC25. This instrument has proven to be reliable over many runs with little maintenance required, however it has historically not been as sensitive to low mass concentrations as needed.

The output of the PCME during the solids injection test on July 28, 2008, is shown in Figure 4-3. The PCME response looks very much like the PPC instrument shown above. The background particle concentration before injection started is detectible only during the puff of particles that occurs after a backpulse and is indicated by a spike associated with cleaning the two plenums. The average output of the PCME during the solids injection period was 3.84 percent (the PCME makes no attempt to convert its readings to actual particle concentrations). Averaging the hour before the start of the injection test the background level was 0.04 percent. Comparison of these values indicates an increase of 96 times in the PCME output attributable to injection of solids. This is somewhat more than the PPC, but that is probably attributable to how the different instruments handle low concentrations. The PCME average is simply that – an average of the spikes with zeros between. As described above, the PPC keeps all particle counts and recalculates over a time period, making it more sensitive to low concentrations with longer integration times.

The relationship between actual particle mass concentrations and the PCME output is shown in Figure 4-4. Within an individual test program there is generally a reasonable linear agreement, but over all of history the results are quite scattered since a differences in particle size distribution affects the PCME response. Since the instrument measures electrical charge on the particles passing nearby, other immeasurable factors may affect the sensitivity of the instrument. Thus, this instrument is simple to maintain and operate and is useful as a filter failure monitor, but it is does not provide actual concentrations or particle size data as does the PPC.

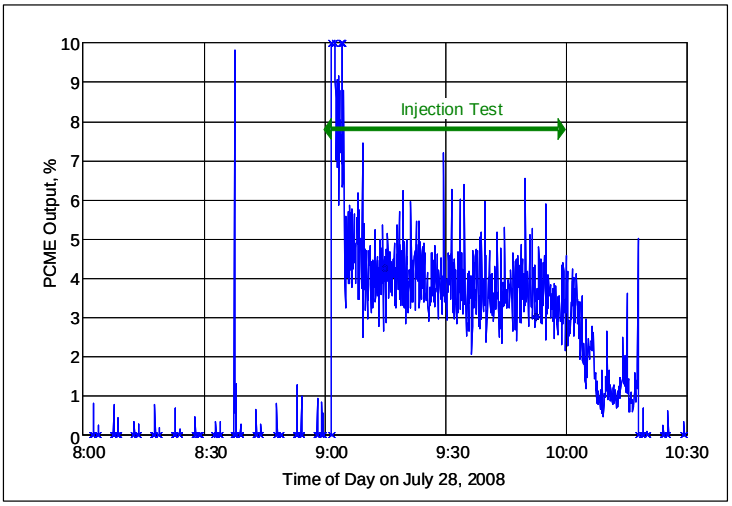

Figure 4-3. PCME Particulate Monitor Output during Solids Injection Test.

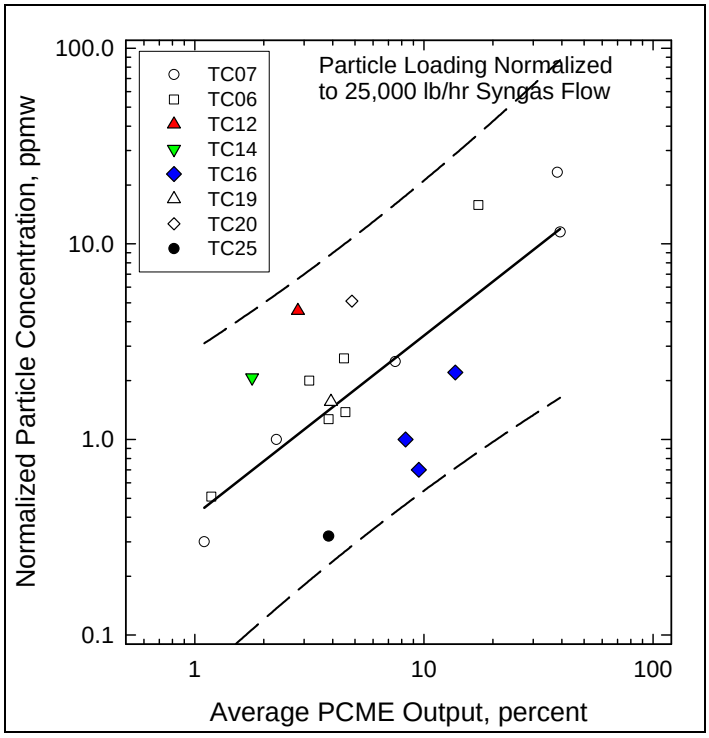

Figure 4-4. Comparison of PCME Particulate Monitor Response to Actual Mass Concentration.

4.2 Thermowell Materials

The primary concern of thermowell performance in the Transport Gasifier is excessive wear of the tip exposed in the gasifier. Mississippi lignite gasification operating parameters provide conditions such as low temperatures and finer gasifier ash that are favorable for long thermowell life. No thermocouple measurements were lost during the test campaign, and eight thermocouples inspected following system shutdown showed no significant metal loss.

Because thermowells with HR-160 material have proved to be more rugged and more economical than ceramic tipped thermowells, the majority of the thermowells tested during TC25 were HR-160 material (51 HR-160 thermowells and 5 ceramic thermowells were installed). Figure 4-5 shows two of the HR-160 thermowells after operation in TC25. These thermowells were typical of the HR-160 material, showing negligible tip wear.

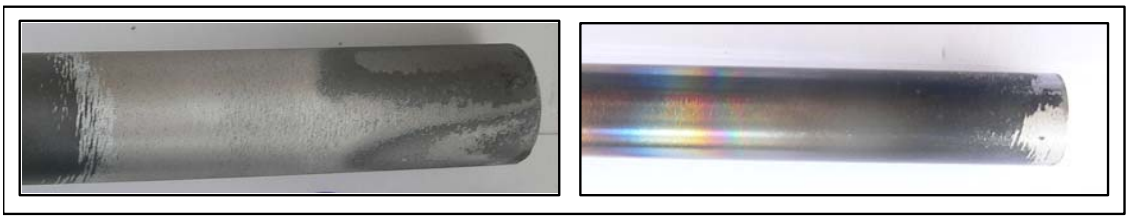

Figure 4-5. Inspection of HR-160 Thermowells after TC25.

4.3 Gas Analysis

Routine analyzers operated during the TC25 test campaign including reflux probes for syngas sampling, and area gas detectors and personal gas monitors needed for compliance with gasification protocol.

A redundant oxygen analyzer was added to the atmospheric syngas combustor. This type of analyzer has been utilized for several other applications at the PSDF (atmospheric fluid bed combustor, stack, and coal mills). This particular instrument is equipped with a high temperature probe, making it well suited for service on the atmospheric syngas combustor.

One of the hydrogen analyzers failed during TC25, and a backup hydrogen analyzer was used for the remainder of the test campaign. The faulty analyzer was not repaired during operation since other vital analyses, namely for hydrogen sulfide, could have been lost or damaged during repairs. Repairs were completed during the outage.

The improvements to the reflux probes reported previously performed well during TC25, and reduced the duration of cleaning periods by 50 percent. More frequent cleaning was required during TC25, however, although this was attributed to an increased flow rate (150 percent of typical flow) which was instituted to decrease analyzer response time.

The first continuous ammonia measurements were achieved during TC25 with the start-up and commissioning of a tunable diode laser (TDL) ammonia analyzer. The instrument performed well for the entire test campaign. To check the accuracy of the TDL analyzer, its results were compared to those of the FTIR ammonia analyzer. Results from the analyzers matched closely

(within 200 ppm) during most of test campaign, although the TDL did occasionally have somewhat higher (up to 600 ppm) readings. These higher readings have been attributed to a much faster response time (real time) and a lack of sampling interference inherent with the FTIR analyzers.

5.0 PARTICULATE CONTROL DEVICE

For TC25, the particulate characteristics and performance of the PCD with Mississippi lignite were quantified by particulate sampling and analyses. Long-term evaluation of filter element and failsafe materials was continued. The filter layout begun in TC24 with a higher efficiency version of the Pall Dynalloy HR-160 sintered metal fiber filter elements in the bottom plenum and with iron aluminide filter elements in the top plenum was continued for TC25. Failsafe testing was not performed during TC25.

5.1 PCD Inlet and Outlet Particulate Measurements

In-situ particulate sampling was performed at the PCD inlet and outlet using the in-situ batch sampling systems described in previous reports. The inlet particulate measurements were used to quantify the particulate loading to the PCD and were used in a later section to calculate transient drag. The outlet measurements indicated the collection performance of the PCD with the installed filter elements.

Particulate mass concentrations measured at the PCD inlet and calculated particulate mass flow rates are given in Table 5-1. The first inlet run was collected at very low coal feed conditions and is not typical of the Mississippi lignite. After that, the carryover to the PCD was in the range of 11,300 to 21,200 ppmw (258 to 496 lb/hr). A lot of the variation in these values was the result of test conditions that evaluated high and low particle loadings, high or low standpipe levels, and high and low gas velocities. The average inlet particle concentration of 15,000 ppmw (350 lb/hr) is probably typical of the Mississippi lignite coal.

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Table 5-1. PCD Inlet and Outlet Particulate Measurements.

	PCD Inlet					PCD Outlet				
Test Date	Run No.	Start Time	End Time	Particle Loading, ppmw	Particle Loading, lb/hr	Run No.	Start Time	End Time	H_2O Vapor, vol %	Particle Loading, ppmw (1)
7/04/08	1	10:15	10:30	6680	94	1	10:00	11:00	13.4	0.17
7/06/08	--	--	--	--	--	2	2:50	3:20	--	3.47 (2)
7/14/08	2	9:30	9:45	15100	310	3	9:15	13:15	8.2	0.34
7/15/08	3	9:00	9:13	15500	309	4	8:45	11:00	8.7	0.65
7/16/08	4	9:15	9:30	14400	318	5	9:00	13:00	NA	0.22
7/17/08	5	9:30	9:45	17700	400	6	9:15	13:15	8.8	0.16
7/18/08	6	8:45	9:00	16100	374	7	8:30	12:30	7.5	0.10
7/21/08	7	9:30	9:45	12000	270	8	8:45	12:45	9.0	<0.10
7/22/08	8	8:45	9:00	12600	300	9	8:30	12:30	8.5	<0.10
7/23/08	9	9:30	9:45	13800	309	10	9:00	13:00	9.6	<0.10
7/24/08	10	9:00	9:15	12300	293	11	8:45	12:45	10.2	<0.10
7/25/08	11	10:00	10:15	11300	264	12	8:45	12:45	9.8	<0.10
7/28/08	12	8:30	8:45	19100	469	13	9:15	10:00	7.9	0.32 (3)
7/29/08	13	9:00	9:15	12200	286	14	8:45	12:45	9.9	<0.10
7/30/08	14	8:45	9:00	11900	258	--	--	--	--	--
7/31/08	15	8:45	9:00	14200	351	15	8:30	12:30	8.5	<0.10
8/01/08	16	9:15	9:30	14900	349	16	9:00	13:00	8.6	<0.10
"	17	13:15	13:30	17000	398	--	--	--	--	--
8/04/08	18	13:30	13:45	15300	336	17	8:45	12:45	9.0	<0.10
8/05/08	19	13:00	13:15	13200	307	18	9:00	13:00	8.1	<0.10
8/07/08	20	10:15	10:30	17300	407	19	9:00	13:00	7.9	<0.10
8/08/08	21	8:45	9:00	16500	410	20	8:30	12:30	7.5	<0.10
8/11/08	22	10:30	10:45	21200	496	21	8:30	12:30	8.2	<0.10
8/12/08	23	8:45	9:00	16700	369	22	8:30	12:30	8.2	<0.10

Notes:
1. A fraction of the mass on all samples is corrosion product not char.
2. Sand circulation only. Startup burner off.
3. Dust injection for particle monitor calibration.

Particulate concentrations measured at the PCD outlet are included in the right half of Table 5-1. The first outlet measurement confirmed that the PCD was leak free and only a small amount of corrosion product was collected on the sample filter. Unfortunately, shortly after that an oxygen breakthrough caused a thermal event that damaged the catalytic filter elements that were installed in the PCD. The measurement on July 6 indicated that a large amount of catalyst material was leaking from the PCD and the unit was shut down for repair.

After restart of the PCD with the catalytic filters removed, the outlet particle loading was elevated for a few days, as is typical after an outage. After a week of operation, the outlet particle concentration dropped below the lower measurement limit of 0.1 ppmw and remained there for the rest of the test campaign. There was some degree of contamination of all the sample filters with corrosion product, but the amount was below the detection limit. The low concentration indicates that the new finer fiber HR-160 Dynalloy filters has high collection efficiency. Microscopic examination of the sample filters suggests that the numbers of large particles (> 10 microns) was reduced relative to the older Dynalloy filter elements. This agrees with the data collected by the PPC system that was discussed in Section 4 of this report that both total and large particle concentrations were within acceptable limits.

5.2 PCD Solids Analysis

Particulate characteristics that can affect PCD pressure drop include particle size distribution, density, porosity, surface area, composition, and flow resistance. These parameters were evaluated for characterization of PCD pressure drop performance.

5.2.1 Particle Size Distributions

A Microtrac X-100 particle size analyzer was used to measure the particle size distributions of the in-situ particulate samples collected at the PCD inlet and the PCD hopper sample used for the laboratory drag measurements.

Hopper Samples. Figure 5-1 compares the differential mass percentage distributions for the in-situ samples with four hopper composite samples used for the TC25 lab drag measurements. (Although the in-situ samples are a more accurate representation of the particulate entering the PCD at a given time, the quantity of material collected is far too small of be useful for drag measurements.) The hopper composites were blended from samples collected during period of similar particulate carbon content. Four well-defined periods of different carbon contents were observed during TC25 ranging from less than 3 percent loss on ignition (LOI) to almost 15 percent LOI. Since previous results have indicated the flow resistance of dustcakes with high carbon contents can be very sensitive to the amount of carbon in the sample, this was an excellent opportunity to look for a similar effect with lower carbon content ashes.

Only minor differences were observed in any of these size distributions. The differences in size distribution should not affect the flow resistance properties of the dustcake significantly. Therefore, the conclusion is that these composite hopper samples are appropriate for simulation of the pressure drop characteristics of the PCD during operation in TC25 and for examining the effect of carbon in the samples in the absence of other factors.

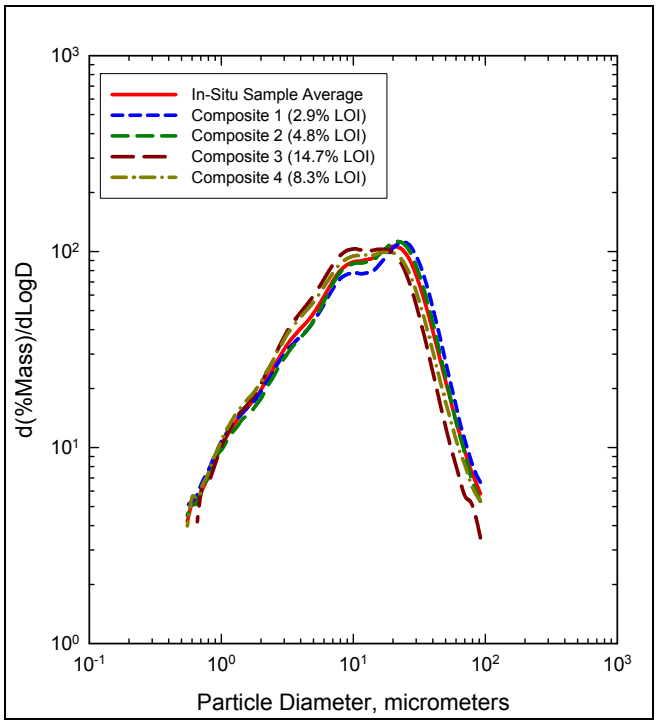

Figure 5-1. Comparison of TC25 Particle Size Distributions.

5.2.2 Dustcake Observations and Thickness Measurements

At the conclusion of TC25, a dirty shutdown of the PCD was performed, successfully preserving the entire (transient-plus-residual) dustcake. As expected, the bottom plenum dustcakes were thicker (averaging 0.11 in.) than the top plenum dustcakes (averaging 0.06 in.) since the bottom plenum HR-160 elements provided less flow resistance than the top plenum FEAL elements.

As noted after TC24, the dustcakes on the metal fiber (HR-160) elements were generally uniform and smooth, while dustcakes on the iron aluminide elements had a lumpy appearance. The degree of lumpiness appeared to increase with increasing age of the element (increasing syngas exposure), suggesting that the variations in dustcake thickness and appearance were influenced by the condition of the element on which the dustcake is collected. Some of this effect may be attributable to changes in the roughness of the element surface brought on by corrosion and wear. Plugging caused by ash penetration and sulfidation could also play a role. These effects will be closely monitored in future testing with the objective of better understanding the effect of filter element condition on the formation and thickness of the dustcake.

5.2.3 Particulate Physical Properties and Chemical Compositions

This section discusses the physical properties and chemical compositions of the in-situ samples collected at the PCD inlet, the PCD hopper sample used for the laboratory drag measurements, and the dustcake samples.

In-situ Samples. Table 5-2 gives the physical properties and chemical compositions of the in-situ samples collected at the PCD inlet and the hopper samples selected for laboratory drag

measurements. All of the samples were collected during operation on the Mississippi lignite, and they generally have consistent physical properties and chemical compositions. The LOI and non-carbonate carbon (NCC) content of the samples are generally low, indicating good carbon conversion.

Table 5-2. Physical Properties and Chemical Composition of In-Situ Samples and Composite Hopper Samples Used for Lab Drag Measurements.

Sample ID	Run No.	Sample Date	Bulk Density g/cc	True Density g/cc	Bulk Porosity %	Surface Area m^2/g	Median Particle Size microns	Loss on Ignition %	$CaCO_3$ Wt %	CaS Wt %	CaO Wt %	Non-Carbonate Carbon Wt %	Inerts (Ash/Sand) Wt %
In-Situ Samples - Mississippi Lignite													
AB25582	1	07/04/08	0.50	2.57	80.5	13	4.3	12.15	0.41	0.65	15.06	10.53	73.35
AB25583	2	07/14/08	0.51	2.73	81.3	11	9.1	1.93	0.66	0.22	19.75	1.56	77.80
AB25584	3	07/15/08	0.50	2.66	81.2	10	13.2	1.56	0.59	0.13	17.78	1.38	80.12
AB25585	4	07/16/08	0.47	3.00	84.3	19	12.2	2.89	0.89	0.38	16.38	2.54	79.80
AB25586	5	07/17/08	0.44	2.61	83.1	35	12.0	5.67	0.86	0.63	16.73	5.16	76.62
AB25587	6	07/18/08	0.45	2.65	83.0	36	11.8	5.98	0.89	0.76	18.21	5.46	74.68
AB25660	7	07/21/08	0.46	2.74	83.2	20	12.3	2.68	0.66	0.34	22.79	2.40	73.82
AB25661	8	07/22/08	0.48	2.78	82.7	17	11.8	2.63	0.82	0.27	24.81	2.30	71.80
AB25836	9	07/23/08	0.40	2.64	84.8	57	10.7	9.10	1.36	1.01	20.47	8.27	68.89
AB25837	10	07/24/08	0.49	2.80	82.5	18	12.9	2.74	0.95	0.40	24.90	2.45	71.29
AB25838	11	07/25/08	0.48	2.79	82.8	16	12.3	2.61	0.09	0.31	24.56	2.42	72.61
AB25880	12	07/28/08	0.42	2.61	83.9	57	10.5	9.05	1.52	1.63	21.77	9.24	65.84
AB25881	13	07/29/08	0.42	2.64	84.1	48	13.3	7.19	1.30	1.10	20.66	6.79	70.15
AB25918	14	07/30/08	0.46	2.71	83.0	19	13.9	3.21	0.80	0.34	20.64	2.85	75.37
AB25952	15	07/31/08	0.44	2.65	83.4	55	12.8	8.22	1.36	1.41	20.31	7.99	68.93
AB25990	16	08/01/08	0.40	2.65	84.9	56	10.3	8.34	1.48	1.39	20.93	8.02	68.18
AB25991	17	08/01/08	0.36	2.44	85.2	90	10.3	17.94	0.70	1.50	18.18	16.71	62.91
AB26084	18	08/04/08	0.42	2.64	84.1	44	11.7	6.50	1.07	0.96	19.46	6.12	72.39
AB26129	19	08/05/08	0.41	2.61	84.3	56	11.9	8.63	1.14	1.32	18.58	8.33	70.62
AB26166	20	08/07/08	0.46	2.65	82.6	40	11.4	5.93	1.30	0.85	18.62	5.50	73.72
AB26204	21	08/08/08	0.39	2.60	85.0	68	11.1	10.29	1.30	1.79	18.58	10.29	68.04
AB26294	22	08/11/08	0.45	2.67	83.1	38	11.6	5.99	1.02	0.76	18.77	5.67	73.77
AB26333	23	08/12/08	0.43	2.64	83.7	39	9.6	6.33	1.11	0.81	17.90	5.99	74.19
Hopper Samples Used for Lab Drag Measurements													
AB26355	Composite	07/25/08	0.47	2.69	82.5	13	12.8	2.93	0.95	0.40	23.69	2.90	72.05
AB26360	Composite	07/30/08	0.39	2.69	85.5	12	12.4	4.80	0.98	0.74	21.36	4.82	72.10
AB26371	Composite	08/01/08	0.38	2.48	84.7	10	9.8	14.70	1.64	1.86	18.24	14.70	63.56
AB26413	Composite	08/08/08	0.39	2.54	84.6	10	10.4	8.31	1.27	1.48	18.86	8.67	69.72

As shown in Figure 5-2. Effect of Carbon Content on Specific Surface Area of Gasification Ash., the specific-surface area and NCC data from TC25 follow the previously established trend based on data from TC22 and previous test campaigns using Powder River Basin (PRB) coal. The bituminous coals (Illinois Basin and Utah) do not follow the same trend and tend to have lower specific-surface areas for the same NCC content. This difference in surface area may be a factor that contributes to the lower drag of the bituminous gasification ash compared to the drag of lignite and PRB gasification ash.

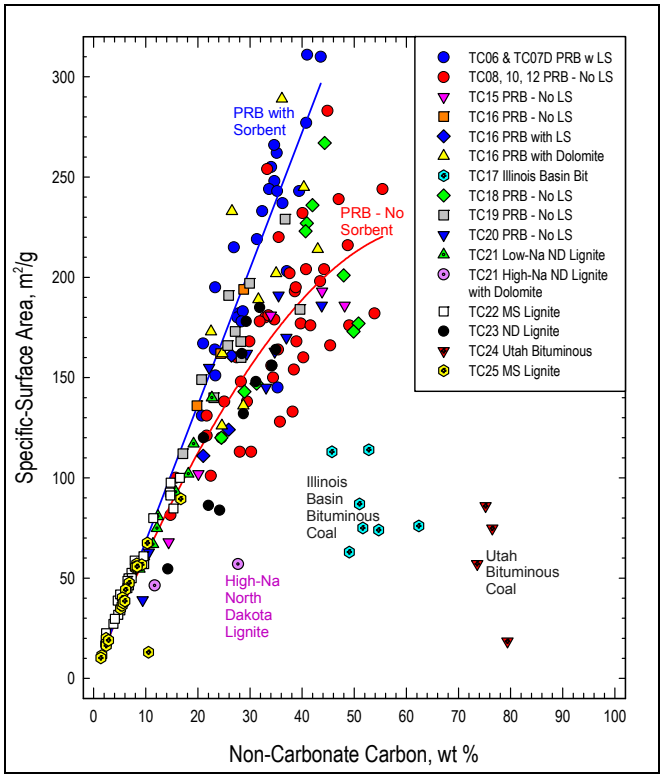

Figure 5-2. Effect of Carbon Content on Specific Surface Area of Gasification Ash.

Composite Hopper Sample. For the lab drag measurements, several different composite hopper samples were prepared to cover a range of NCC contents, since previous studies have shown that the drag varies with NCC content. As shown in Table 5-2, the physical properties and chemistry of the composite hopper sample were similar to those of the in-situ samples, suggesting that the composite hopper samples were representative samples for the lab drag measurements.

Dustcake Samples. Table 5-3 gives the physical properties and chemical composition of the dustcake samples taken after TC25. The samples included a bulk sample of the transient dustcake taken from both plenums and a bulk sample of the residual dustcake taken from both plenums. The physical properties of the dustcake were generally similar to the physical properties of the in-situ samples, except that the dustcake had a finer mean particle size. As seen in previous tests, the residual dustcake typically has a finer particle size distribution than does the transient dustcake, due to the effects of fine-particle enrichment in the residual dustcake.

Table 5-3. Physical Properties and Chemical Composition of Dustcake Samples.

Sample ID	Sample Date	Bulk Density g/cc	True Density g/cc	Bulk Porosity %	Surface Area m^2/g	Mass Median Diameter mm	Loss on Ignition %	$CaCO_3$ Wt %	CaS Wt %	CaO Wt %	Non-Carbonate Carbon Wt %	Inerts (Ash/Sand) Wt %
Transient Cake from Both Plenums												
AB26449	09/10/08	0.46	2.66	82.7	30	10.6	12.13	0.02	0.63	2.71	17.20	79.44
Residual Cake from Both Plenums												
AB26450	09/10/08	0.46	2.67	82.8	35	7.3	9.82	0.02	0.40	2.03	13.74	83.80

In addition to the slight difference in particle size, the chemical composition of the dustcake is markedly different from that of the in-situ samples. The dustcake generally contains a much higher level of non-carbonate carbon. This may suggest that there was a drop-off in carbon conversion just prior to shutdown. In fact, during the 20-minute period prior to tripping the coal feed, the peak pressure drop across the PCD dropped from about 120 to 72 inH_2O, suggesting a decline in solids carryover prior to shutdown. This probably explains the lower carbon conversion in the dustcake samples.

5.2.4 PCD Pressure Drop Performance

5.2.4.1 Transient PCD Drag

The pressure rise within a cleaning cycle of the PCD is a direct measure of the characteristics of the particulate being collected at that time. Under stable operation, the vast majority of this particulate is removed from the filter elements during cleaning, so this is referred to as the transient pressure drop. Since pressure drop is a function of the gas velocity, temperature (gas viscosity), particulate loading, and the flow resistance of the particulate, describing PCD operation in terms of pressure drop makes comparison of different conditions difficult. Instead, a value of normalized drag is calculated, which is pressure drop that is normalized to 1 ft/min face velocity, 1 lb/ft^2 areal particulate loading, and gas viscosity of air at 70°F. The result is a fundamental parameter that describes the flow resistance of the collected dustcake.

For each in-situ sample at the PCD inlet, the PCD transient drag was calculated using the measured particulate concentration along with the pressure drop increase and face velocity during the period of the in-situ test. All of the particulate measured at the PCD inlet is assumed to be collected on the filter elements and to contribute to pressure drop.

The inputs and results of the transient drag calculations are shown in Table 5-4. The calculated transient drag at PCD conditions is listed under the column heading “PCD.” The corresponding value of transient drag normalized to the viscosity of air at room temperature (70°F) is listed under the heading “PCD@RT”. These values are comparable to the lab drag measurements discussed in a later section and are also comparable to other test campaigns that operated at different temperatures. The comparison of these values is shown in a later section. The drag values for TC25 were relatively low, consistent with previous experience with lignite coal operation.

Table 5-4. Transient Drag Determined from PCD Pressure Drop and from Lab Measurements.

Run No.	ΔP/Δt, inwc/min	Δ(AL)/Δt, lb/ft^2/min	FV, ft/min	MMD, μm	NCC, %	Drag, inwc/(lb/ft^2)/(ft/min)		
						PCD	PCD@RT	Lab
1	0.88	0.007	3.96	4.3	10.5	120	76	80
2	0.83	0.024	3.44	9.1	1.6	35	21	27
3	0.57	0.024	3.33	13.2	1.4	24	14	18
4	0.81	0.025	3.68	12.2	2.5	33	20	20
5	1.29	0.031	3.78	12.0	5.2	42	25	22
6	1.18	0.029	3.91	11.8	5.5	41	25	23
7	0.49	0.021	3.73	12.3	2.4	24	14	20
8	0.50	0.023	4.01	11.8	2.3	22	13	21
9	1.09	0.024	3.98	10.7	8.3	46	28	28
10	0.71	0.023	4.01	12.9	2.4	32	19	19
11	0.57	0.020	3.92	12.3	2.4	28	17	20
12	2.06	0.036	4.68	10.5	9.2	57	34	29
13	1.03	0.022	3.44	13.3	6.8	47	28	21
14	0.54	0.020	2.92	13.9	2.9	27	16	18
15	1.28	0.027	4.08	12.8	8.0	47	28	22
16	1.52	0.027	3.91	10.3	8.0	56	34	29
17	2.30	0.031	3.84	10.3	16.7	75	46	36
18	1.12	0.026	3.65	11.7	6.1	43	26	24
19	1.43	0.024	3.91	11.9	8.3	60	36	25
20	1.39	0.031	3.95	11.4	5.5	44	26	24
21	1.77	0.032	4.19	11.1	10.3	56	33	28
22	1.29	0.038	3.94	11.6	5.7	34	20	24
23	1.77	0.028	3.68	9.6	6.0	62	37	29
Avg Lig	1.16	0.027	3.82	11.7	6.0	42	26	24
Lab drag data calculated from linear regression to MMD and NCC of lab drag samples.								

Nomenclature:

ΔP/Δt = rate of pressure drop rise during particulate sampling run, inwc/min.
Δ(AL)/Δt = rate of increase in areal loading during sampling run, lb/min/ft^2.
FV = average PCD face velocity during particulate sampling run, ft/min.
MMD = mass-median diameter of in-situ particulate sample, μm.
NCC = non-carbonate carbon. LOI = Loss On Ignition.
RT = room temperature, 77°F (25°C).

Normalized PCD transient drag is plotted as a function of carbon content in Figure 5-3. As seen in previous test campaigns, transient drag increased with increasing carbon content in the gasification ash. Good agreement is seen between the TC22 and TC25 ashes that were generated from the same Mississippi lignite coal. The relatively poor agreement of the TC20 data points with the average line for all PRB results shows that carbon content is not the only factor affecting drag (particle size).

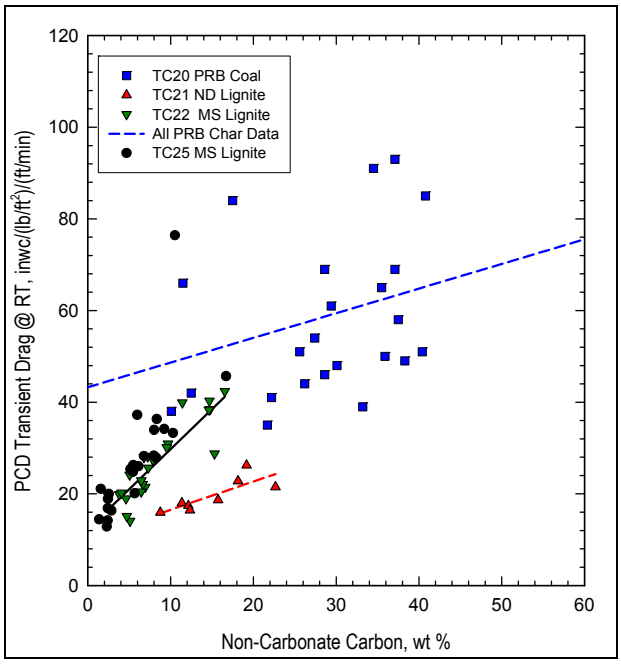

Figure 5-3. PCD Transient Drag versus Carbon Content of In-Situ Samples.

5.2.4.2 Baseline Pressure Drop Analysis

Figure 5-4 shows the baseline PCD pressure drop measurements for TC25 as well as recent test campaigns. While these pressure drop values are normalized for temperature and gas velocity, they are not normalized for particulate concentration as are the drag data discussed in the previous section.

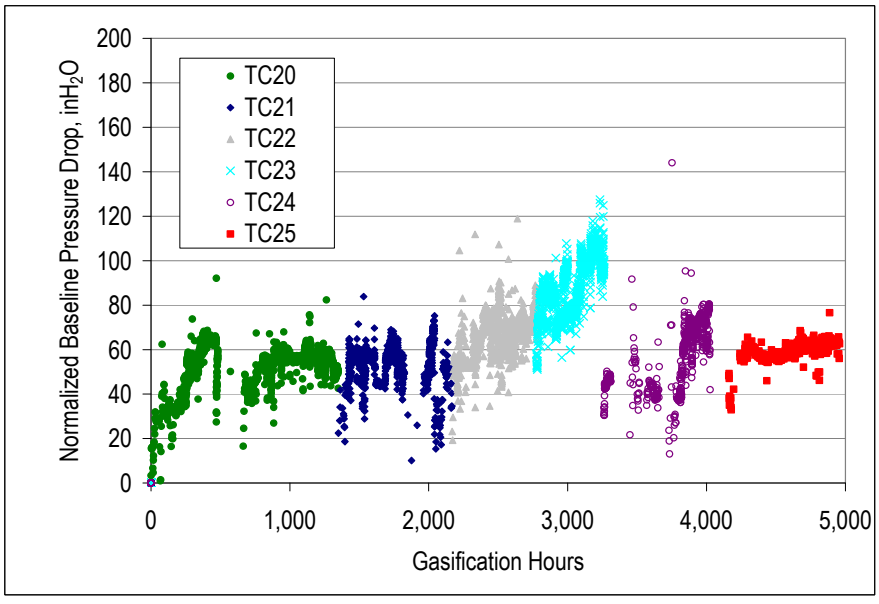

Figure 5-4. Normalized Baseline Pressure Drop from TC20 through TC25.

Both baseline and peak pressure drop were moderate during TC25 especially considering the high particulate mass rate and subsequent elevated areal loading on the filter elements that would have resulted. The moderate pressure drop is consistent with the low values of dustcake drag measured in the lab.

For TC25, the bottom plenum was installed with finer fiber version of the Pall Dynalloy filter elements. These finer fiber elements were expected to provide improved collection efficiency but increased pressure drop. However, neither this effect nor the effect of the old FEAL filter elements installed in the top plenum was obvious from the pressure drop data.

5.2.5 Prediction of PCD Drag and Pressure Drop

5.2.5.1 Lab Drag Measurements

To investigate the characteristics of the TC25 particulate more completely, drag measurements were made in the lab flow resistance test device on the composite hopper samples. This lab apparatus uses a series of cyclones between the particulate generator and the dustcake collection surface to vary the particle size distribution of the dustcake. Figure 5-5 illustrates the results with normalized drag plotted against the MMD of the collected dustcake. The four sets of drag data showed the expected positive correlation between drag with carbon content. All of the samples also showed a negative correlation between drag and particle size.

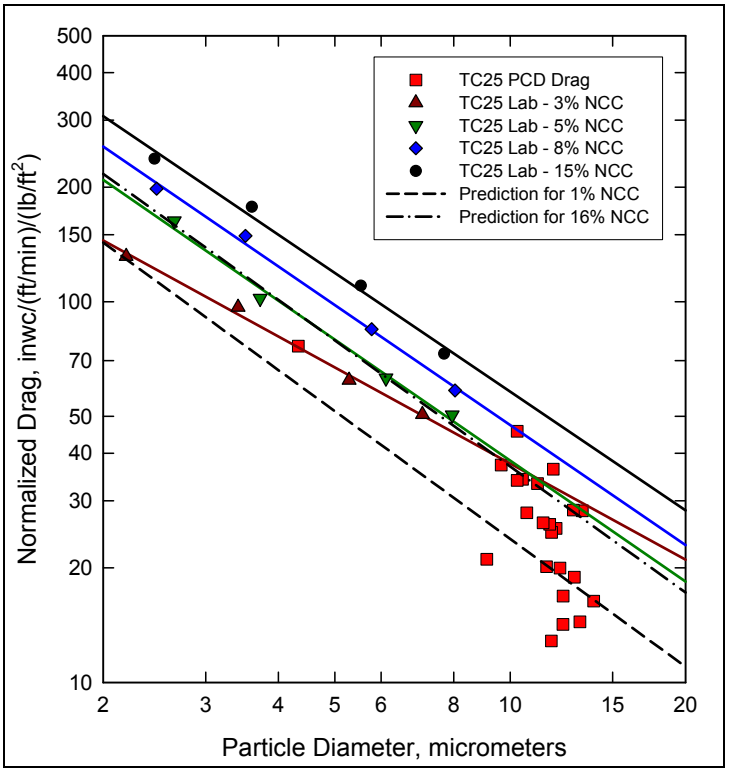

Figure 5-5. Lab-Measured Drag as a Function of Particle Size.

The relationship between drag, particle size, and carbon content was evaluated by calculating a multiple regression. The resulting equation relating these values for the TC25 data is shown by the equation:

Drag = 10^(2.475 – (1.100 * Log(MMD)) + 0.0120 * NCC), with an $r^2 = 0.972$

Figure 5-5 graphically represents the relationship by the two dashed lines which represent the output of the equation for carbon contents of 1 percent and 16 percent over a range of particle sizes. The 1 percent and 16 percent are the high and low range of the in-situ values. Also shown on Figure 5-5 are the individual values of PCD transient drag calculated for each of the in-situ samples from Table 5-4. The results of regression predictions for each individual value of PCD transient drag are shown in the rightmost column of Table 5-4. These calculations use the MMD and carbon content of each in-situ sample to predict the transient drag of the PCD during that test. The lab predictions show good agreement with the actual PCD data for both the individual values and the average for TC25.

5.2.5.2 Comparison of Lab Measurements with Transient Drag

Average laboratory and PCD drag values for all gasification test campaigns are summarized in Table 5-5. The comparison shows excellent overall historical agreement (average difference of about 10 percent). For TC25, the difference was slightly better than average, at negative 8 percent. The average results for all gasification test campaigns are plotted in Figure 5-6 and continue to show data points scattered around the perfect agreement line.

Table 5-5. Average Drag Values Determined from PCD and Lab Measurements.

Run	Coal	Average Transient Drag Determined from PCD Performance, inwc/(lb/ft^2)/(ft/min)	Average Drag Determined from RAPTOR Lab Measurements, inwc/(lb/ft^2)/(ft/min)	Difference from Mean Value*, %
GCT2	PRB	29	21	-33.5
GCT3	PRB	80	93	14.5
GCT4	PRB	66	57	-15.2
TC06	PRB	89	81	-9.6
TC07	PRB	48	50	4.3
TC08	PRB	47	50	7.3
TC09	Hiawatha	29	23	-21.8
TC10	PRB	45	58	25.2
TC11	Falkirk Lignite	16	36	76.2
TC12	PRB	58	61	4.7
TC13	Freedom Lignite	34	39	13.6
TC14	PRB	47	42	-13.0
TC15	PRB	55	76	33.3
TC16	PRB + Limestone	49	52	4.8
TC16	Lignite + Dolomite	26	42	47.1
TC17	IL Basin	25	19	-27.8
TC18	PRB	59	82	32.6
TC19**	PRB	64	72	11.8
TC20**	PRB	78	108	32.3
TC21**	ND Lignite	19	32	51.0
TC22**	MS Lignite	27	40	38.8
TC23**	ND Lignite	65	62	-4.7
TC24**	Utah Bit.	28	18	-43.5
TC25**	MS Lignite	26	24	-8.0
Average		46	52	9.2

* D = (R1-R2)/(R1+R2)/2*100
** Technique modified to use carbon content of lab drag sample

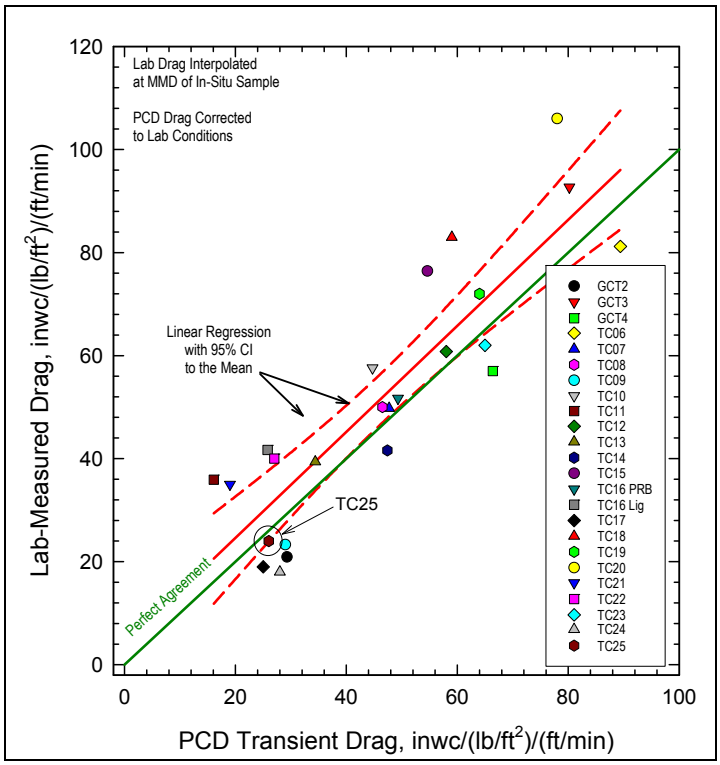

Figure 5-6. Comparison of PCD Transient Drag with Lab Measurements.

5.3 Analysis of PCD Filter Element Condition

After completion of TC25, a complete set of measurements was made on the filter elements to determine the effect of exposure hours on pressure drop and to determine if corrosion was affecting the two types of elements installed.

During shutdown of the gasifier at the end of TC25, attempts were made to preserve the transient filter dustcake by stopping the PCD backpulse system at the same time coal feed was stopped. Nonetheless, by the time the filter elements were removed from the PCD, most of the transient dustcake had fallen off, so flow resistance measurements could not be made with that dustcake intact. For flow testing, the elements were air blown to remove the remaining vestiges of transient dustcake and achieve a uniform residual dustcake for testing. The TC25 dustcakes were dry and easy to remove.

FEAL Elements. The pressure drop at fixed face velocity (3 ft/min) was measured on each of the filter elements both with the light residual dustcake and after pressure washing to remove all particulate. The data are plotted as a function of gasification exposure hours for the FEAL filter elements in Figure 5-7. These are not historical data in the plot, but are a snapshot of the filter elements actually installed in TC25. Displaying the data this way prevents differences in dustcake drag or porosity from biasing the age comparison. The FEAL filter elements continued to show the previously established trend of increasing pressure drop with time for both clean and dirty elements.

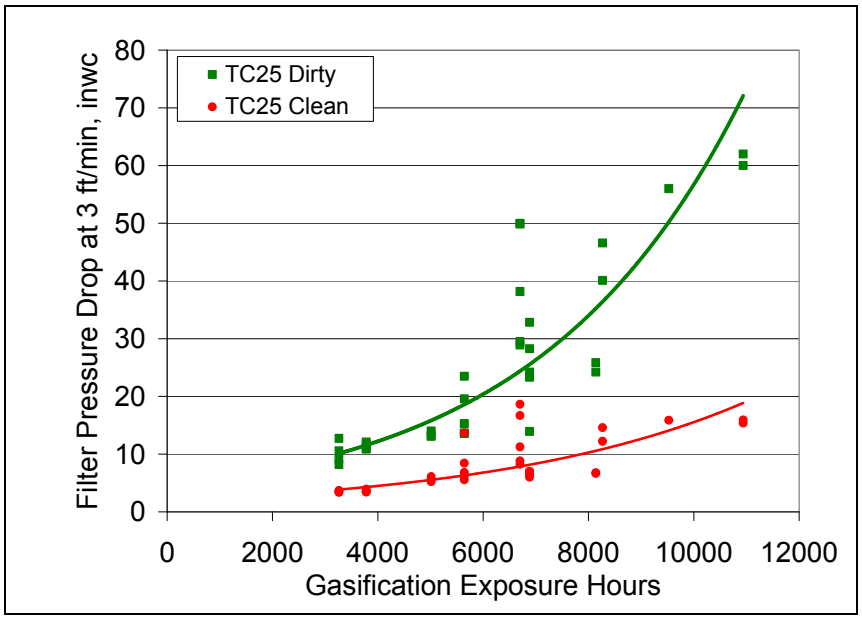

Figure 5-7. Pressure Drop versus Exposure Hours for Iron Aluminide Filter Elements.

A microscope photo of the oldest FEAL filter element with 10,938 hours in gasification is shown in Figure 5-8. Even after pressure washing, the normal porous structure of the filter is completely obscured by corrosion product. The black and gray areas are covered with iron compounds with different degrees of sulfation, while the brown spots are regions of pitting corrosion.

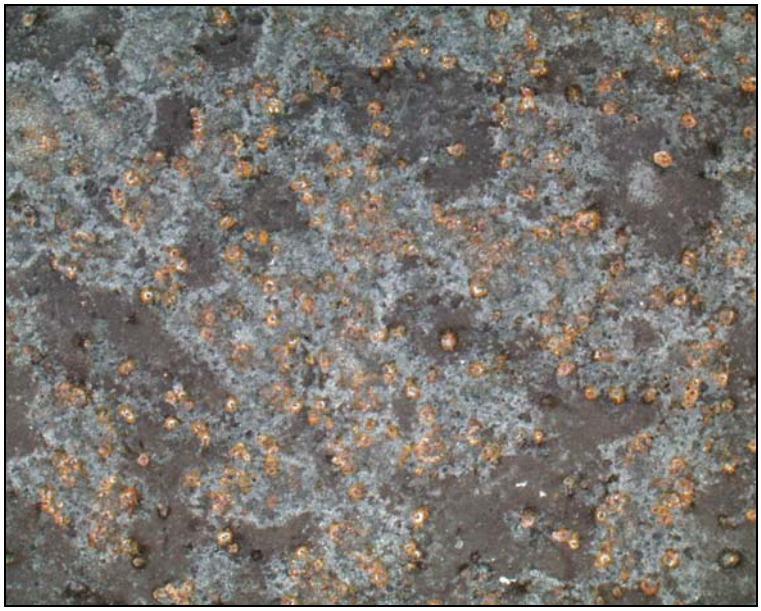

Figure 5-8. Iron Aluminide Filter Media with 10,938 Gasification Hours.

Dynalloy HR-160 Elements. The two types of Pall Dynalloy fiber filter elements evaluated in TC25 were composed of coarse and fine HR-160 alloys. The fine fiber element has higher collection efficiency at the expense of higher pressure drop and possibly greater corrosion tendency.

The pressure drop at 3 ft/min face velocity for the fine fiber Dynalloy elements are shown in Figure 5-9. These elements all have the same current age (979 hours) so to indicate a range of hours the data from both TC24 and TC25 are shown. This plot appears to indicate that the dirty pressure drop of the HR-160-F elements increased with time. However, since the dustcakes were collected from different test campaigns, the data could be biased.

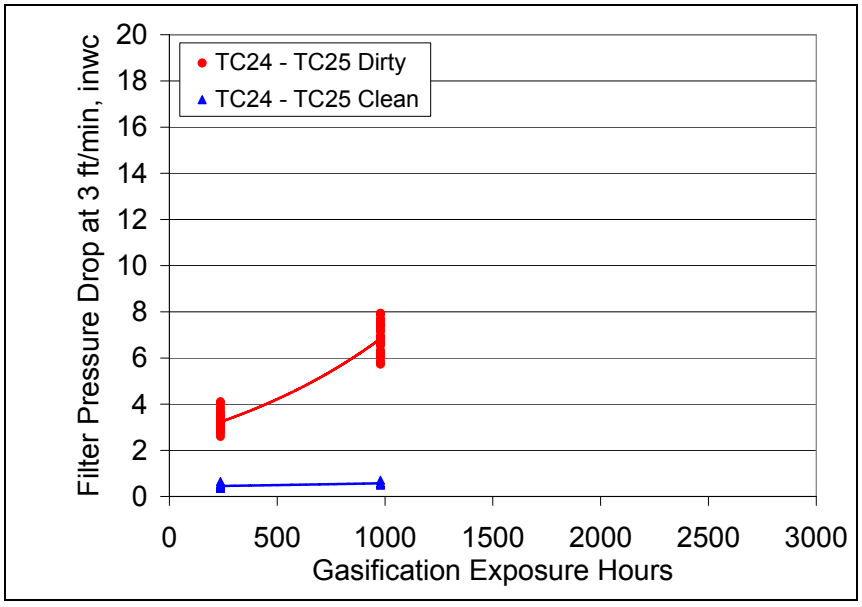

Figure 5-9. Pressure Drop versus Exposure Hours for HR-160-F Filter Elements.

Although the HR-160 was thought to be corrosion resistant, corrosion of these elements was observed. Figure 5-10 is a microscope photo of a portion of a coarse fiber filter element showing pit-like corrosion with pits that are about 50 microns in size and appear to be growing together. Although the fine fiber elements have less than 1,000 hours of operation, Figure 5-11 shows discolored areas that indicate corrosion is occurring with small pits forming at the center of some areas.

Figure 5-10. Coarse Fiber HR-160 Dynalloy Filter Media with 2,698 Gasification Hours.

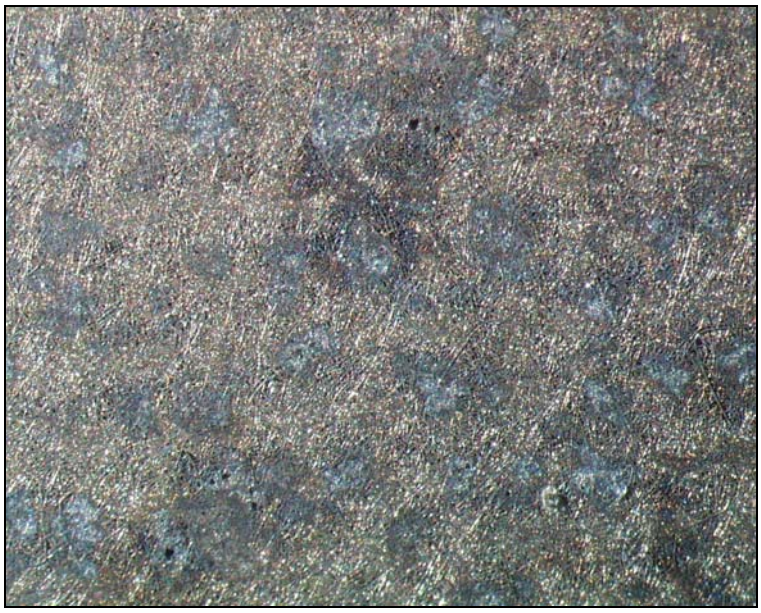

Figure 5-11. Fine Fiber HR-160 Filter Media with 979 Gasification Hours.

6.0 Advanced Syngas Cleanup

Several tests performed in TC25 continued to advance syngas cleanup technologies. Testing of WGS systems and catalysts continued, as did exploratory CO_2 capture testing and testing of a hydrocarbon cracking catalyst, mercury sorbent, and a sulfur sorbent. The syngas cleanup unit was also used to support researchers from the Department of Energy (DOE) National Energy Technology Laboratory, Media Process Technology, and Johnson Matthey.

6.1 Water-Gas Shift Reaction Testing

Two water-gas shift catalysts, the Sud-Chemie T-2822 shift catalyst and the Johnson Matthey Katalco K8-11 shift catalyst, were tested during TC25 using catalytic filter elements. Table 6-1 lists the catalyst properties.

Table 6-1. Water-Gas Shift Catalyst Properties.

Catalyst Supplier	Sud-Chemie	Johnson Matthey
Catalyst Trade Name	T-2822	Katalco K8-11
Chemical Composition		
Aluminum Oxide Content, wt %	50-70	----
Magnesium Oxide Content, wt %	15-35	----
Molybdenum Oxide Content, wt %	5-15	10
Cobalt Oxide Content, wt %	1-10	4
Calcium Oxide Content, wt %	3-7	----
Physical Properties		
Form	Powder	Powder
Size, microns	100-200	100-200
Density, lb/ft^3	50	50

During the outage preceding TC25, the catalysts were pre-sulfided using a gas mixture of 5 percent H_2S and 95 percent H_2 to convert the cobalt and molybdenum, the active ingredients of the catalyst, to the sulfided form. The pre-sulfidation was performed at 200 psig and 660°F. After the pre-sulfidation was completed, the catalysts were pulverized and sieved to a particle size of nominally 100 to 200 microns and then packed in iron aluminide filter elements, with an annular surface layer approximately 1 cm thick.

In the PCD, three filter elements were installed with inner tubes filled with the Sud-Chemie catalyst. In the syngas cleanup unit, two filter elements, one with an inner tube filled with the Sud-Chemie shift catalyst and one with an inner tube of the Johnson Matthey shift catalyst, were installed in parallel pressure vessels.

During the first portion of the test campaign, TC25A, after about 8 hours of gasification operation, thermocouples on the three PCD catalytic elements indicated a thermal upset with temperatures up to 1,340°F, as shown in Figure 6-1. High temperatures were not measured on any other PCD filter elements. The thermal transient occurred during the time of combustion mode operation for removal of syngas cooler fouling when the maximum oxygen concentration in the PCD inlet gas was 1.8 mole percent.

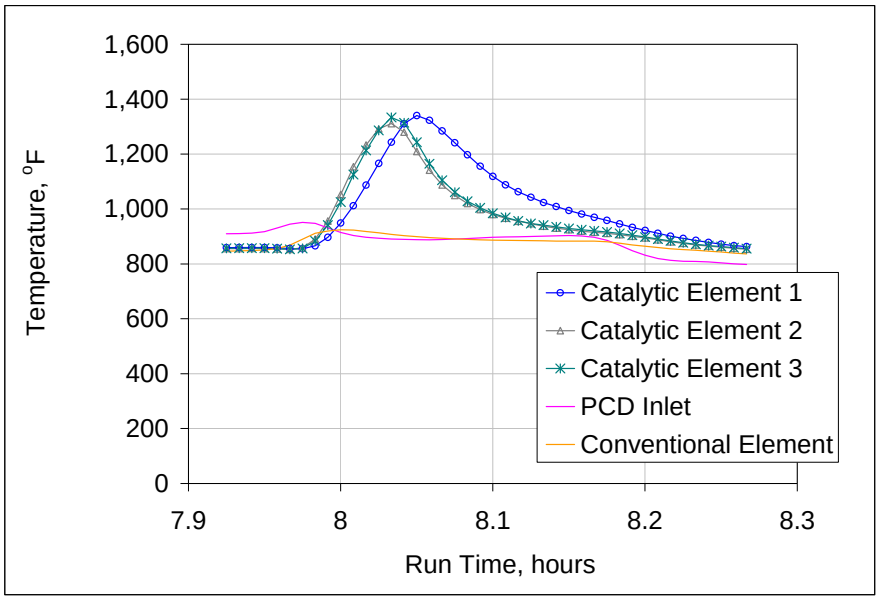

Figure 6-1. Catalytic Filter Element Temperatures during Thermal Upset at Hour 8.

During the subsequent system restart at around Hour 39, the same PCD catalytic filter element thermocouples indicated a dramatic temperature increase, which is shown in Figure 6-2. As before, the thermocouples on these filter elements read higher than any other PCD thermocouples. These thermocouples exceeded the high end of their scale of 2,000°F. This temperature upset occurred at low risk conditions for gasification ash combustion when the measured oxygen level was below three percent. Thermal upsets also occurred in the syngas cleanup unit with the Sud-Chemie catalyst-filled element (the Johnson Matthey catalyst-filled element was isolated from syngas flow during TC25A).

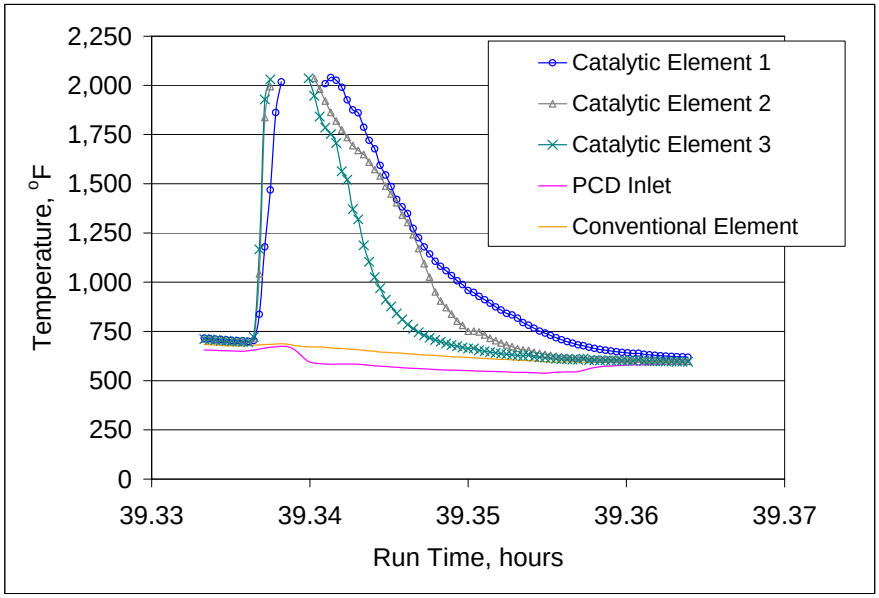

Figure 6-2. Catalytic Filter Element Temperatures during Thermal Upset at Hour 39.

The thermal transients were apparently caused by a highly exothermic reaction of sulfide oxidation. After the catalyst material (which had been pre-sulfided prior to installation) was exposed to a reducing atmosphere, it became highly reactive with oxygen. The catalytic elements were removed from the PCD during the 8-day outage. The catalytic elements remained in the syngas cleanup unit for further testing, since in the cleanup unit, the elements could be easily isolated from process gas flow during re-starts when oxygen is present at low concentrations.

During TC25B, the catalytic elements were successfully tested in the syngas cleanup unit with syngas for about 538 hours. While operating on syngas, the operating conditions were varied to evaluate the performance of the catalytic filter elements. Table 6-2 gives the operating conditions during the catalyst testing.

Table 6-2. Water-Gas Shift Catalyst Test Conditions.

Syngas Exposure Time	162	372
Operating Pressure, psig	150 to 200	150 to 200
Operating Temperature, ^{o}F	600 to 800	650 to 780
Syngas Flow Rate, lb/hr	50	25 to 50
Steam Flow Rate, lb/hr	5	0
Filter Element Face Velocity, ft/min	2.4 to 3	0.8 to 2.5
H_2O-to-CO molar ratio	2.4 to 3.4	0.8 to 1.2
Shift Conversion, percent	22 to 94	20 to 50

Figure 6-3 and Figure 6-4 show the Sud-Chemie and Johnson Matthey Katalco catalysts, respectively, before and after testing in TC25B.

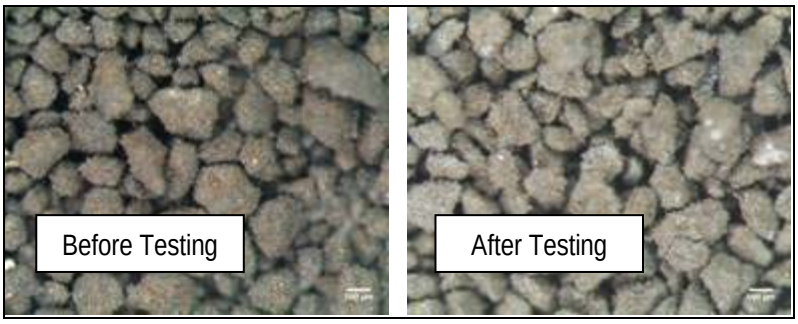

Figure 6-3. Photomicrographs of Sud-Chemie T-2822 Shift Catalyst before and after Testing.

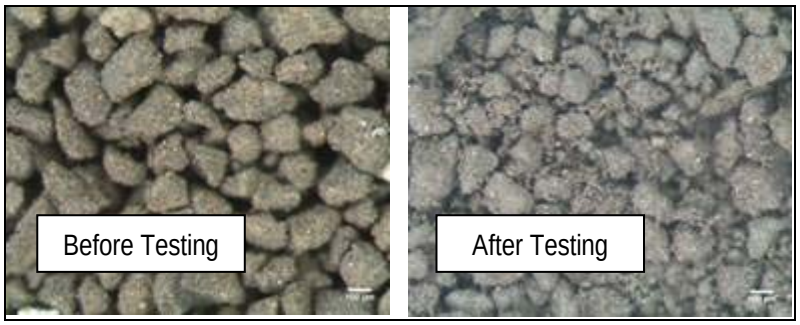

Figure 6-4. Photomicrographs of Johnson Matthey Katalco K8-11 Shift Catalyst before and after Testing.

A combination of gas chromatographs and process analyzers provided continuous data acquisition for the water-gas shift tests. The new Rosemount process gas chromatograph (GC) measured syngas quality (H_2, N_2, CO, CO_2, and CH_4). This instrument also controlled the sample conditioning system. A Rosemount XStream process analyzer was used as a fast response analysis for CO and CO_2. During TC25, an older ABB process GC served as a redundant instrument. Results from all three instruments were virtually identical. A modified Baldwin chiller unit, controlled by the Rosemount GC, allowed for continuous monitoring of the shift tests, which was the first time continuous measurements had been possible at the syngas cleanup unit.

6.2 NETL Fuel Cell Module

Testing of the NETL fuel cell module continued in TC25. The fuel cell module is a multi-cell array mobile platform developed to test different solid oxide fuel cells in parallel on coal-derived syngas. The unit is designed to enable testing for up to 12 individual fuel cells simultaneously over a range of electric load conditions for extended periods of time to provide data on the influence of trace coal contaminants such as arsenic, phosphorous, selenium, and mercury on fuel cell performance. This information is critical for development of fuel cells for coal-based power generation.

NETL researchers continued field modifications to resolve previously identified issues related to gas seals and current leads. Following the repair, a brief 30 hour test run was successfully conducted on hydrogen and syngas on 3 of the 12 cells in the multi-cell array (MCA). With this success, the MCA was shut down in order to rebuild all 12 cells. The test was successfully restarted with hydrogen as the fuel with 11 of 12 cells in good condition. Unfortunately, an attempt to load the current failed due to an unexpected high resistance possibly due to corrosion on the cathode nickel current leads. Replacement of the nickel leads with sliver leads resolved this issue, and testing was restarted with hydrogen fuel. After the cells stabilized, the testing transitioned to syngas fuel for the remaining duration.

Analyses for low-level sulfur compounds, hydrocarbon content, and feed gas composition were supplied during this testing effort. The chiller unit mentioned previously also allowed for continuous sulfur and composition analyses during fuel cell operation. FTIR samples taken once a day to verify the operation of a hydrocarbon cracking catalyst. Other services included

provision of compressed gas supplies (hydrogen and liquid argon) and the construction of gas manifolds for the uninterrupted delivery of hydrogen.

After NETL researchers resolved issues previously identified, the solid oxide fuel cell (SOFC) MCA was restarted successfully with hydrogen fuel and switched to conditioned syngas on August 3. Fuel cell operation on syngas lasted for a total of 200 hours and shut down on August 11 as planned. The experience gained and data collected through the combination of the exposure and electrochemical tests under real syngas conditions should yield valuable information to assist in future SOFC technology development fueled on coal derived gas. During this period, a NETL modified gas chromatograph with inductively coupled plasma/mass spectrometer was also successfully commissioned for accurately measuring syngas trace metals at the sub-ppm level.

The fuel cell module was dismantled on August 12 and shipped back to NETL on August 13. This marked the conclusion of the SOFC test project. During this test period (January 31 through August 13, 2008), PSDF provided support to the project on many fronts including project coordination, equipment installation and dismantling, gas supply and analysis, test stand repair, and necessary test monitoring.

6.3 MPT Carbon Molecular Sieve Membrane

An advanced CMS membrane was installed and commissioned during TC25. Working in cooperation with NETL, researchers at MPT developed the membrane which is highly selective for hydrogen and particularly well suited for coal-derived syngas. The membrane separates hydrogen from syngas with membrane materials that were extensively lab-tested. The objective of the testing at PSDF was to evaluate material stability of the membrane under gasification conditions with particulate-free coal-derived syngas containing both major and minor contaminants including hydrocarbons at the parts per million level. Figure 6-5 is a photograph of the CMS membrane installed at the PSDF.

Figure 6-5. MPT Carbon Molecular Sieve Membrane Installed at PSDF.

MPT successfully tested a single tube of a carbon molecular sieve (CMS) membrane on syngas with and without hydrogen augmentation. For the raw syngas without enrichment, the hydrogen concentration was 8 to 10 percent in the feed and 40 to 50 percent in the permeate. For enriched

syngas, the hydrogen concentration was 18 to 20 percent in the feed and 80 to 90 percent in the permeate. Overall, the CMS membrane was exposed to nearly 19 hours of syngas or hydrogen-enriched syngas. Although there were some minor operational upsets and difficulties in conducting in-situ gas permeance and selectivity measurements, the permeance and selectivity results suggest that the membrane performance was stable during the test and unaffected by contaminants in syngas. MPT will complete further data analysis. This was the first time a hydrogen-selective membrane was successfully operated on untreated coal-derived syngas.

PSDF personnel provided gas analytical services and sampling support to MPT for this test. Initially, a process gas GC was modified for the analysis of the membrane permeate stream (which contained high hydrogen concentrations). After it was determined that accurate monitoring could not be achieved with a single GC, an automated laboratory GC (Agilent 5890) was put into service to monitor the membrane reject stream. New sample conditioning systems were fabricated and installed when it was determined that the conditioning systems of the membrane skid were not adequate. Smaller "knock out" pots were also supplied to supplement the membrane skid's internal conditioning system. These smaller cylinders decreased the analysis response time by 80 percent. Other services and equipment provided to MPT include overall project coordination, skid installation and removal, and various interface connections.

6.4 CO_2 Capture

During TC25, about 40 absorption and 40 regeneration tests were performed with the bench scale Parr reactor to capture syngas CO_2. During testing, parameters such as temperature, pressure, gas flow rate, and solvent concentration were varied to determine the optimum operating conditions for the absorber and regenerator. The syngas CO_2 concentration varied from nominally 10 to 15 percent. Preliminary analysis showed promise for high CO_2 capture efficiencies.

Prior to syngas testing, kinetic tests with the bench scale reactor was performed using bottled CO_2 and nitrogen. The data generated from the bench scale reactor was used as a guide for the design of the continuous pilot reactor for CO_2 capture. In addition, preliminary solvent/additive screening tests for CO_2 capture from flue gas at low pressure were performed. Initial results indicate that the amines such as piperazine were the most effective solvents for CO_2 capture from flue gas at low pressure.

6.5 Johnson Matthey Mercury Sorbent

A Johnson Matthey mercury sorbent composed of palladium was tested in the syngas cleanup unit to capture mercury and other trace metals present in the syngas. Pre-treatment of the catalyst involved reduction for 2.5 hours with hydrogen (diluted with nitrogen) at a pressure of 100 psig, temperature ranging from about 70 to 200^{o}F (increased at a rate of 3.6^{o}F/min), and a space velocity of 2,120 hr^{-1}.

The mercury sorbent was exposed to 500°F syngas over a period of 330 hours, 260 hours with hydrocarbon cracking and 70 hours without hydrocarbon cracking. The syngas samples were collected in the inlet and outlet of the mercury sorbent reactor for the analysis of mercury and other trace metals using a modified EPA Method 29. Depending on sample flow, sampling times

ranged from 2 to 5 hours. Results indicated high capture of mercury. During the test, the pressure ranged from 150 to 200 psig, the syngas flow ranged from 25 to 50 lb/hr, and the space velocity varied from 1,500 to 3,700 hr^{-1}.

6.6 Sorbent Trap Testing

In addition to collecting trace metal samples using the modified EPA Method 29 (impinger trains) during the Johnson Matthey mercury sorbent tests, sorbent traps were used in a parallel gas stream. The modified sorbent trap collection method (EPA Method 30B), which consists of one Appendix K sorbent trap and two impingers, was used on five different occasions. The Appendix K trap, shown in Figure 6-6, was designed primarily to trap mercury, but data suggests that it may capture other trace metals as well (notably arsenic and selenium). This method offers significantly reduced sampling and preparation times compared to the impinger collection with EPA Method 29. Further testing of the sorbent trap method will be conducted in future gasification test campaigns to confirm that this method is viable for trace metals testing in this environment.

Figure 6-6. Sorbent Trap Used for Trace Metals Testing.

6.7 Sulfur and Hydrocarbon Removal

Syngas was treated with sulfur removal and hydrocarbon cracking prior to its use in the fuel cell and mercury sorbent test skids. A sulfur sorbent from Synetix, Puraspec 2010, which is composed mainly of zinc oxide, was effective in reducing the syngas sulfur levels below the detection limit, typically 1 to 1.5 ppm. Use of the Sud-Chemie hydrocarbon cracking catalyst FCR-4 reduced concentrations of organics such as benzene and naphthalene by 90 to 99 percent.

7.0 Support Equipment

7.1 Startup Burner

The start-up burner operated for 87 hours during TC25 with no major operational issues. Initial start-up of TC24A was delayed by problems lighting the pilot flame of the start-up burner. A new shroud assembly was installed prior to operation and after several failed attempts to light the pilot, the burner pilot assembly was modified. Modifications to the start-up burner were successful in improving the ability to light the start-up burner pilot. A shroud was installed around the burner igniter to improve the fuel/air mixture that was being ignited. Figure 7-1 shows the start-up burner pilot assembly before and after the modifications. The extended burner tip and improved vapozier performance ensured easier lighting during TC25 and improved start-up burner performance.

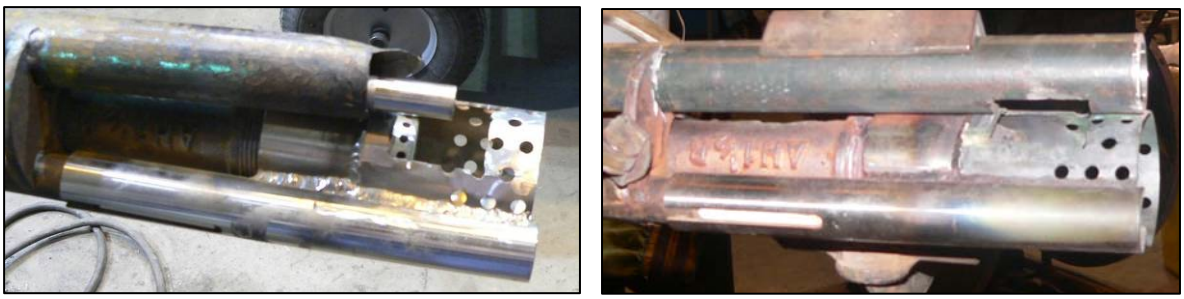

Figure 7-1. Start-up Burner Pilot Assembly.

7.2 Recycle Gas Compressor

The recycle syngas compressor supplied syngas for gasifier aeration for 372 hours during TC25. The system operated well and experienced no major problems. Steady state operating conditions at the recycle gas compressor outlet, as well as the recycle syngas flow rate to the gasifier are shown in Figure 7-2. There were six compressor related trips during TC25. During start-up, there were initially issues with vibration; however, this issue did not warrant any downtime.

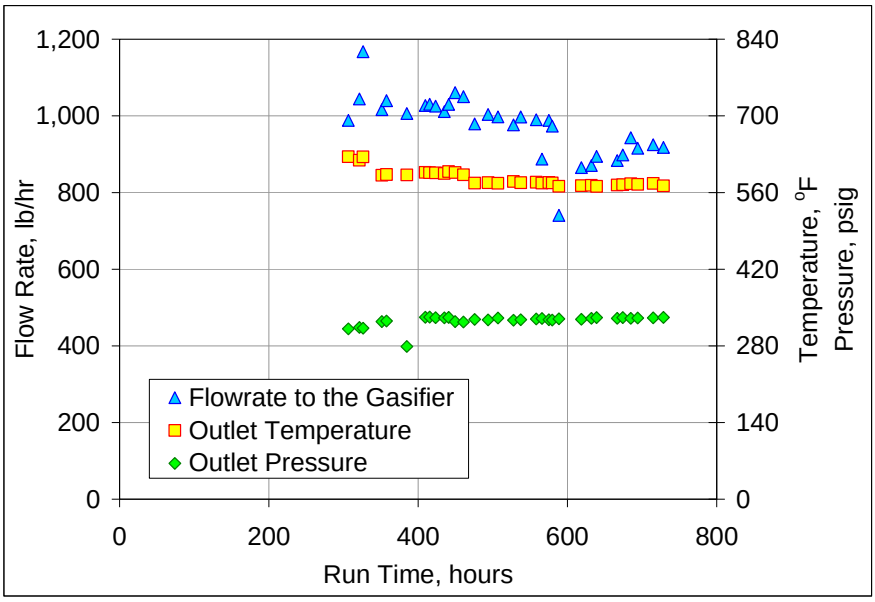

Figure 7-2. Recycle Gas Compressor Operating Conditions.

8.0 CONCLUSIONS

Test campaign TC25 allowed characterization of gasifier operation and performance with high moisture lignite from Mississippi after processing in the new fluid bed dryer system. The test campaign demonstrated steady operation and provided the opportunity for further testing of instrumentation enhancements, hot gas filter materials, and advanced syngas cleanup technologies. The PSDF site was also made available for testing of the National Energy Technology Laboratory's fuel cell module and Media Process Technology's hydrogen selective membrane with syngas from the Transport Gasifier.

Lessons Learned. Information gained from TC25 operation included:

- Fluid bed dryer lessons learned:
  - Quantification of coal particle size attrition through dryer
  - Quantification of oversize removal rate
  - Dryer operation at different hot water supply temperatures
  - Minimization of nitrogen consumption
  - Confirmation of design mass and energy balances
  - Development of PLC controls
  - Identified steps to stop a CO increase
  - Demonstrated that cleanout of the fluidization windbox was unnecessary
  - Improved the nitrogen makeup and nitrogen vent control
- PDAC design concepts proved viable, although additional modification needs were indicated by an unsteady coal feed rate.
- Modifications to original coal feed line showed that lower velocity was sufficient to reliably feed coal. Design of coal feed lines should avoid areas of expansion around valves and fittings.
- The higher efficiency achieved in the gasifier solids separation devices increased carbon conversion as compared to operation with Mississippi lignite in TC22.
- The effect of temperature on carbon conversion was less significant at higher cyclone collection efficiencies.
- Weakness of the Sureflow 88 seal leg refractory caused an obstruction in the gasifier solids circulation path.
- Parametric testing of the fraction of air flow fed from the lower mixing zone showed that maximizing the temperature differential between the lower and upper mixing zones resulted in unstable solids flow to the CCAD ash removal system. Data showed that the optimal percentage of air flow to the lower mixing zone was about 35 percent.
- Operational data showed a strong linear relationship between gasifier solids circulation rate and standpipe level.
- Tar formation and deposition in the primary gas cooler is difficult to clean but can be successfully removed during startup by operating with excess oxygen.
- HR-160 tipped thermowells tested at the PSDF are less expensive and more robust than ceramic thermowells, and, as a result, are the preferred material choice for gasifier thermocouples.
- Use of a modified chiller unit allowed for continuous monitoring of the water-gas shift tests—the first time continuous measurements had been possible at the syngas cleanup unit.

- The Sud-Chemie water-gas shift catalyst used in the PCD and the syngas cleanup unit was highly reactive with oxygen after being in a reduced environment.
- The NETL fuel cell module successfully operated with syngas for over 200 hours, and data collected through the combination of the exposure and electrochemical tests under real syngas conditions can be used for future SOFC technology development.
- Operation of the Johnson Matthey mercury sorbent showed high rates of mercury removal.
- Testing the Media and Process Technology hydrogen selective CMS membrane indicated stable operation with syngas. Performance of the membrane was not affected by syngas contaminants during the testing.
- The iron aluminide PCD filter elements continued to show the previously established trend of increasing pressure drop with time for both clean and dirty elements. Corrosion with the older elements (>10,000 hours of exposure) is characterized by obscuring with corrosion product of the normal porous structure. Areas of the media were covered with iron compounds with different degrees of sulfation, and with regions of pitting corrosion.
- Although the coarse fiber HR-160 material was thought to be corrosion resistant, corrosion of these filter elements has become evident as pitting in elements with over 2,000 hours of exposure.
- Fine fiber HR-160 filter elements with less than 1,000 hours of operation showed corrosion, with discolored areas and small pits forming at the center of some of the discolored areas.
- The shroud installed around the burner igniter improved the fuel/air mixture characteristics for igniting. The extended burner tip and improved vaporzier performance allowed easier lighting during TC25 and improved start-up burner performance.

APPENDIX A OPERATING HISTORY

System commissioning of the KBR Transport Reactor train and the first five test campaigns (TCs) were performed in combustion mode. Approximately 5,000 hours of combustion operation were completed from 1996 to 1999. The system was transitioned to gasification operation in late 1999. Four gasification commissioning tests (GCTs), each lasting nominally 250 hours, were completed by early 2001. At the conclusion of TC25, 20 gasification test campaigns were completed, each nominally 250 to 1,500 hours in duration, for a total of about 11,500 hours of coal gasification operation. Powder River Basin subbituminous coal is the most extensively tested fuel, although several bituminous and lignite coals have also been tested. The Transport Gasifier has operated successfully in both air-blown and oxygen-blown modes.

Table A-1 summarizes the gasification testing completed at the conclusion of TC25. The table lists the duration, number of hours on coal, fuel type, and major objectives of each test. More information about the individual test campaigns may be found in the test campaign reports, located on the PSDF website, http://psdf.southernco.com.

Table A-1. Gasification Operating History.

Test	Start Date	Duration (hrs)	Fuel Type*	Comments
GCT1	September 1999	233	PRB, Illinois #6, Alabama	First gasification testing
GCT2	April 2000	218	PRB	Stable operations
GCT3	February 2001	184	PRB	Loop seal commissioning
GCT4	March 2001	242	PRB	Final gasification commissioning test
TC06	July 2001	1,025	PRB	First long duration test campaign
TC07	April 2002	442	PRB, Alabama	Lower mixing zone commissioning
TC08	June 2002	365	PRB	First oxygen-blown testing First on-line failsafe testing
TC09	September 2002	309	Hiawatha	New mixing zone steam system
TC10	October 2002	416	PRB	Developmental coal feeder commissioning
TC11	April 2003	192	Falkirk Lignite	First lignite testing
TC12	May 2003	733	PRB	Fuel cell testing
TC13	September 2003	501	PRB, Freedom Lignite	Syngas to combustion turbine
TC14	February 2004	214	PRB	Syngas to combustion turbine CFAD commissioning
TC15	April 2004	200	PRB	Improved oxygen feed distribution
TC16	July 2004	835	PRB, Freedom Lignite	Fuel cell testing High pressure O_2-blown operation
TC17	October 2004	313	PRB, Illinois Basin	Bituminous coal testing
TC18	June 2005	1,342	PRB	Recycle gas compressor commissioning
TC19	November 2005	518	PRB	CCAD commissioning
TC20	August 2006	870	PRB	Gasifier configuration modifications
TC21	November 2006	388	Freedom Lignite	First lignite test following the gasifier modifications
TC22	March 2007	543	Mississippi Lignite	High moisture lignite testing
TC23	August 2007	481	PRB, Freedom Lignite	High sodium lignite testing
TC24	February 2008	237	Utah	First bituminous coal test following the gasifier modifications
TC25	July 2008	742	Mississippi Lignite	Fluid bed dryer commissioning

*Note: PRB is a subbituminous coal; Illinois #6, Alabama, Hiawatha, Utah, and Illinois Basin coals are bituminous coals.

APPENDIX B STEADY STATE OPERATING PERIODS AND MAJOR OPERATING PARAMETERS

There were 42 steady state operating periods during TC25, all during TC25B. These periods are given in Table B-1, along with the major operating parameters for each period. The steady state periods are defined based on maintaining gasifier operating conditions within defined ranges.

All of the steady state periods were in air-blown gasification mode with high moisture lignite from the Mississippi Red Hills mine. Recycle syngas operation occurred in 32 of the operating periods. The coal feed rates were calculated from the feeder weigh cells, and the air, steam, and recycle syngas flow rates were taken from flow indicators. The PCD solids rates were determined from the in-situ sampling at the PCD inlet, and the ash removal rates for CCAD were determined by a system ash balance.

Table B-1. Steady State Operating Periods and Major Operating Parameters (Page 1 of 2).

Steady State Operating Period	Start Time (2008)	End Time (2008)	Run Time Hours	Gasifier Mixing Zone Temp., °F	Gasifier Outlet Press., psig	Coal Feed Rate, lb/hr	Air Feed Rate, lb/hr	Steam Feed Rate, lb/hr	Recycle Syngas Flow Rate, lb/hr	Syngas Rate, lb/hr	PCD Inlet Temp., °F	Gasifier Solids Removal Rate, lb/hr	PCD Solids Removal Rate, lb/hr
TC25-1	07/17 00:00	07/17 05:15	122	1,680	200	4,330	12,150	210	0	23,520	730	460	370
TC25-2	07/18 04:45	07/18 09:30	150	1,690	200	4,340	12,410	100	0	23,790	740	270	380
TC25-3	07/19 19:30	07/20 01:15	189	1,750	210	4,000	12,680	90	0	23,790	750	230	320
TC25-4	07/20 12:30	07/20 16:30	206	1,770	210	3,920	12,610	160	0	23,620	760	250	300
TC25-5	07/20 19:30	07/21 03:30	215	1,760	210	3,980	12,920	180	0	24,280	760	260	280
TC25-6	07/21 19:00	07/21 23:15	236	1,790	210	4,040	12,970	20	0	24,030	760	230	280
TC25-7	07/22 02:45	07/22 09:00	245	1,800	210	4,100	13,280	80	0	24,370	770	250	300
TC25-8	07/22 17:15	07/22 21:45	259	1,660	210	3,710	11,160	160	0	22,160	720	170	300
TC25-9	07/23 09:00	07/23 13:15	274	1,670	190	3,500	11,510	610	0	22,310	720	190	310
TC25-10	07/23 14:45	07/23 18:15	280	1,730	210	3,580	12,210	540	0	23,290	740	180	300
TC25-11	07/24 17:15	07/24 21:45	307	1,750	210	4,200	12,990	370	1,020	24,570	760	260	280
TC25-12	07/25 08:30	07/25 12:15	321	1,750	210	3,790	12,720	560	1,050	23,760	760	220	270
TC25-13	07/25 13:15	07/25 17:15	326	1,750	210	3,650	11,590	450	1,080	21,960	740	190	280
TC25-14	07/26 14:00	07/26 18:45	351	1,750	210	3,680	12,150	100	1,030	23,010	740	150	350
TC25-15	07/26 19:00	07/27 02:15	358	1,750	210	3,730	12,190	180	1,030	23,060	740	150	370
TC25-16	07/27 23:30	07/28 04:15	385	1,770	180	4,350	13,500	10	1,010	25,060	760	180	450
TC25-17	07/29 00:15	07/29 04:45	410	1,760	240	4,340	12,900	50	1,020	24,150	750	310	330
TC25-18	07/29 05:15	07/29 11:30	415	1,740	240	4,290	12,840	120	1,030	23,890	750	360	290
TC25-19	07/29 14:30	07/29 18:15	423	1,750	240	4,180	13,120	270	1,030	24,360	750	340	280
TC25-20	07/30 02:00	07/30 06:15	435	1,760	260	3,780	12,040	330	1,010	22,390	740	290	260
TC25-21	07/30 07:30	07/30 12:30	441	1,760	260	3,940	11,690	120	1,020	21,740	740	330	260

Table B-1. Steady State Operating Periods and Major Operating Parameters (Page 2 of 2).

Steady State Operating Period	Start Time (2008)	End Time (2008)	Run Time Hours	Gasifier Mixing Zone Temp., °F	Gasifier Outlet Press., psig	Coal Feed Rate, lb/hr	Air Feed Rate, lb/hr	Steam Feed Rate, lb/hr	Recycle Syngas Flow Rate, lb/hr	Syngas Rate, lb/hr	PCD Inlet Temp., °F	Gasifier Solids Removal Rate, lb/hr	PCD Solids Removal Rate, lb/hr
TC25-22	07/30 16:00	07/30 21:15	450	1,760	210	4,230	13,270	220	1,060	24,600	760	340	290
TC25-23	07/31 04:00	07/31 07:30	461	1,760	210	4,260	13,310	280	1,050	24,600	760	280	340
TC25-24	07/31 18:45	07/31 22:45	476	1,750	210	4,510	13,100	80	950	24,540	750	350	350
TC25-25	08/01 12:45	08/01 16:45	494	1,710	210	4,410	12,390	350	1,000	23,920	730	310	400
TC25-26	08/02 02:00	08/02 06:00	507	1,770	210	4,270	12,890	0	1,000	24,470	760	270	380
TC25-27	08/02 21:15	08/03 04:15	528	1,740	210	4,180	12,570	190	990	23,800	750	330	370
TC25-28	08/03 04:30	08/03 16:30	538	1,740	210	4,020	12,510	60	1,000	23,650	750	280	360
TC25-29	08/04 05:00	08/04 09:30	558	1,740	210	3,970	12,020	100	1,010	22,750	750	280	340
TC25-30	08/04 13:00	08/04 17:00	566	1,750	210	3,780	11,690	180	930	22,070	740	260	340
TC25-31	08/04 22:15	08/05 02:15	575	1,750	210	4,190	13,030	120	980	24,340	750	340	320
TC25-32	08/05 02:15	08/05 07:15	580	1,760	210	4,210	13,130	160	970	24,420	760	350	320
TC25-33	08/05 11:15	08/05 16:00	589	1,770	210	4,250	12,770	50	730	23,360	760	370	310
TC25-34	08/07 16:00	08/07 19:15	619	1,760	210	4,880	14,070	0	870	26,180	770	330	410
TC25-35	08/08 03:45	08/08 10:30	632	1,750	210	4,460	13,440	100	880	25,120	760	270	410
TC25-36	08/08 12:15	08/08 16:00	639	1,750	210	4,330	13,060	0	890	24,470	760	250	420
TC25-37	08/09 15:45	08/09 19:45	667	1,770	210	3,860	12,630	180	910	23,430	760	170	450
TC25-38	08/09 21:15	08/10 05:15	674	1,760	210	4,100	12,500	80	910	23,470	750	200	460
TC25-39	08/10 06:45	08/10 17:15	685	1,770	210	4,280	13,240	40	930	24,440	770	230	470
TC25-40	08/10 17:15	08/11 01:15	694	1,780	210	4,240	13,370	140	920	24,670	770	220	480
TC25-41	08/11 10:15	08/12 02:00	715	1,780	210	3,950	12,590	100	930	23,410	760	190	450
TC25-42	08/12 05:15	08/12 10:00	729	1,760	210	3,860	12,140	150	920	22,300	750	250	380

APPENDIX C MATERIAL AND ENERGY BALANCES

The material and energy balances showed reasonable accuracy given the diversity of the measurements used for their calculation. A gasifier mass balance for the TC25 steady state operating periods is shown in Figure C-1. The mass balance documents the accuracy of the solids and gas rates at the inlet and outlet of the gasifier. The data agreed within 5 percent.

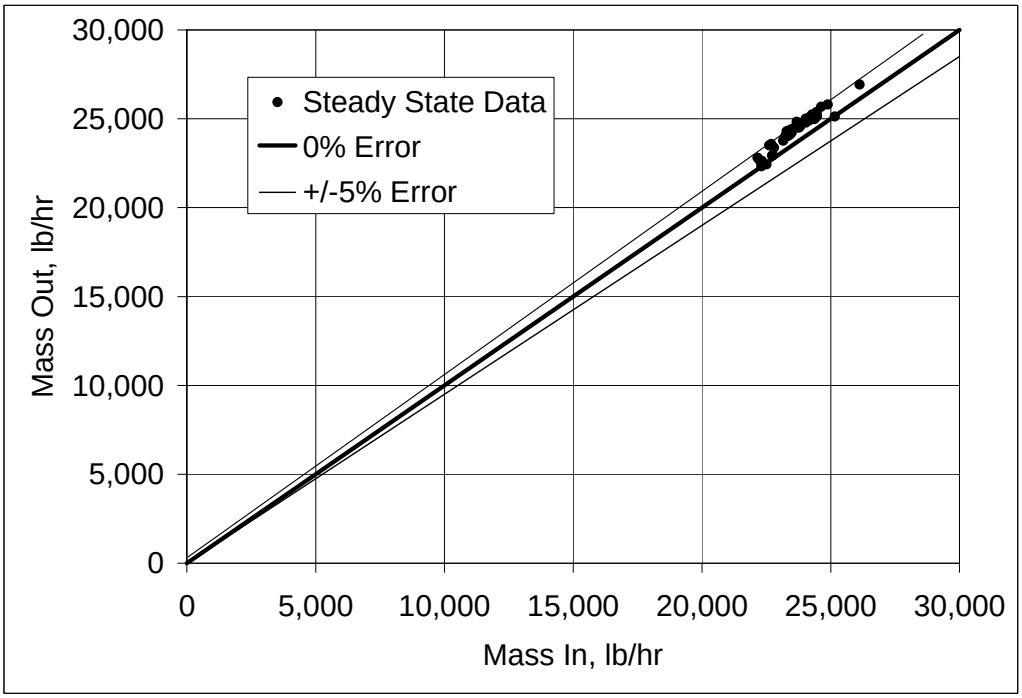

Figure C-1. Mass Balance.

Figure C-2 plots the overall energy balance for the gasifier. Calculations for the energy balance assumed a gasifier heat loss of 3.5 MMBtu/hr. This balance verifies the accuracy of the gasification efficiencies, and shows agreement within about 20 percent.

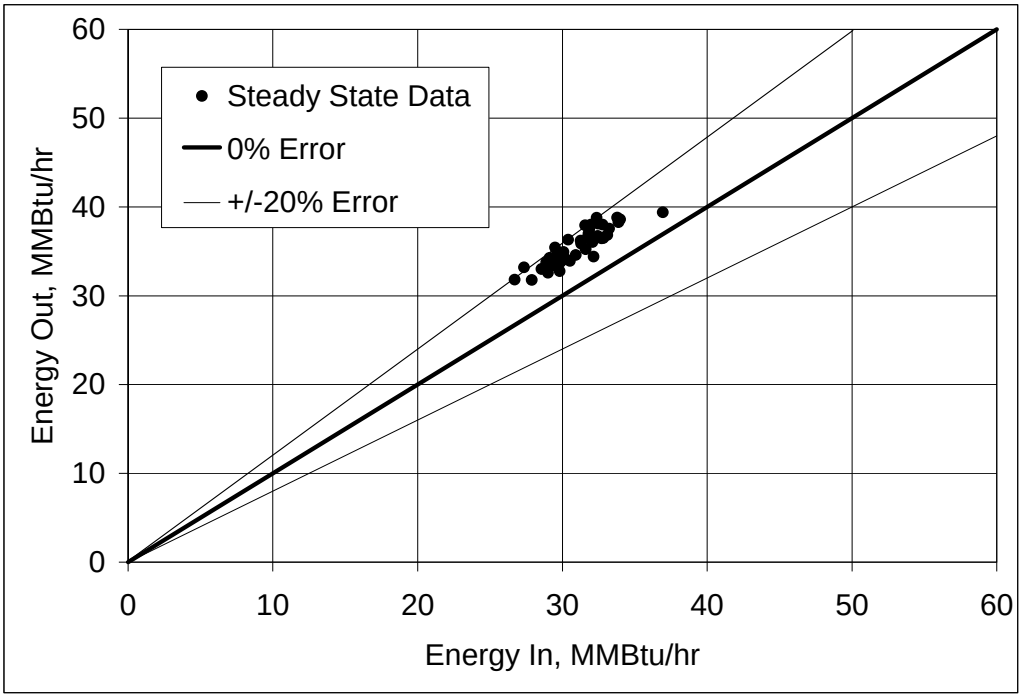

Figure C-2. Energy Balance.

Figure C-3 gives the carbon balance, which documents the accuracy of the carbon conversions. The carbon balances for all but one of the steady periods fell within a 15 percent error range.

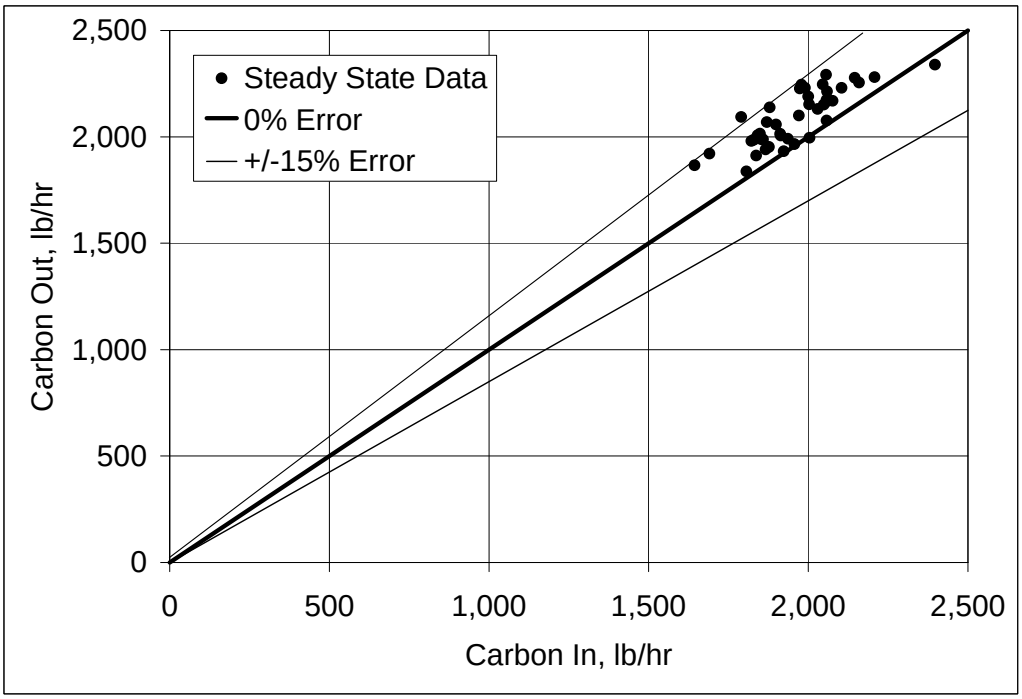

Figure C-3. Carbon Balance.

APPENDIX D LIST OF ACRONYMS AND ENGINEERING UNITS

Acronyms

ASTM—American Society for Testing and Materials
CCAD—Continuous Coarse Ash Depressurization
CFAD—Continuous Fine Ash Depressurization
CMS—Carbon Molecular Sieve
DOE—Department of Energy
EDS—Energy Dispersive X-Ray Spectrometry
EPA—Environmental Protection Agency
FEAL—Iron Aluminide
FTIR—Fourier Transform Infrared
GCT—Gasification Commissioning Test
IGCC—Integrated Gasification Combined Cycle
GC—Gas Chromatograph
LHV—Lower Heating Value
LMZ—Lower Mixing Zone
LOI—Loss on Ignition
MCA—Multi-Cell Array
MMD—Mass Median Diameter
MPT—Media and Process Technology
NCC—Non-Carbonate Carbon
NETL—National Energy Technology Lab
PCD—Particulate Control Device
PDAC—Pressure Decoupled Advanced Coal
PPC—Process Particle Counter
PRB—Powder River Basin
PSD—Particle Size Distribution
PSDF—Power Systems Development Facility
SEM—Scanning Electron Microscope
SMD—Sauter Mean Diameter
SOFC—Solid Oxide Fuel Cell
SRI—Southern Research Institute
TC—Test Campaign
UMZ—Upper Mixing Zone
WGS—Water Gas Shift

Engineering Units

Btu—British thermal units
cm—centimeter
^{o}F—degrees Fahrenheit
ft—feet
ft^3—cubic feet
g/cm^3 or g/cc—grams per cubic centimeter
hr— hours
inH_2O—inches of water
in—inches
inwc—inches of water column
lb—pounds
mg/acm—milligrams per actual cubic meter
min—minutes
mm—millimeters
MMBtu—million British thermal units
mol—mole
µm—microns or micrometers
MW—megawatts
ppm—parts per million
ppmv—parts per million by volume
ppmw—parts per million by weight
psi—pounds per square inch
psig—pounds per square inch gauge
s or sec—second
SCF—standard cubic feet
wt—weight